Handbook of NATURAL GAS TRANSMISSION and PROCESSING

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This book is dedicated to all of those persons who have come before us as family, scientists, and engineers and from whom we have derived comfort and knowledge.

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FOREWORD

The de-carbonization of fuels, a process that started two centuries ago and has little to do with often strident environmental ideologues will position natural gas as the next premier fuel of the world economy. The world, headed by North America, is shifting towards natural gas for power generation as a substitute for coal and oil. Nuclear, while it would be the obvious competitor still seems to suffer from adverse publicity and real large up-front costs and economy of scale requirements. Eventually, the real breakthrough for natural gas use must be its major contribution as a fuel used for transportation, not necessarily directly as compressed natural gas for engine combustion but indirectly, providing the necessary electricity in electrified transportation. There are plentiful natural gas supplies in the world. They are also diverse, compared to oil, and this makes natural gas far more attractive from a geopolitical point of view. Russia, a non-OPEC member has the largest proven reserves. Emerging or speculated consumption by China and India, the always large U.S. energy appetite, along with maturing U.S. domestic production are the primary reasons for the huge expected demand for natural gas transportation from producers to consumers. Natural gas, compared to oil, is far more cumbersome to handle, process, transport and store. There are two common means of transport: pipelines and liquefied natural gas (LNG). Compressed natural gas intended to be transported in sea-going vessels may also emerge. Pipelines are not practical in almost all cases of sea distances longer than 500 miles. This is the reason for the recent major buildup of LNG facilities throughout the world. In the consuming nations, once LNG trade emerges in a large scale the natural gas prices are likely to be kept reasonable for decades.

While several books exist for natural gas engineering in upstream there has been a void for a comprehensive book that captures the entire chain of natural gas including transmission and processing. This book follows an interdisciplinary approach and discusses all relevant aspects of gas gathering, processing, and transmission. The book emphasizes engineering rather than science. It makes use of available field data and converts them into a form that can be used easily by both operating engineers and researchers. The book also presents a comprehensive chapter on gas pricing. Even though it has been long discussed, few ever ventured into presenting a theory of gas pricing, let alone a comprehensive econo-political basis for it. This chapter serves as a reminder that it is never too late to correct the practice of the past – the fact that there is no current international gas pricing mechanism does not mean we should not advance a theory on the topic.

Natural gas is an important aspect of any environmental initiative. Considering this, another chapter of this book describes efforts on the part of the natural gas industry to lessen the environmental impact of natural gas processing and discusses the environmental effects of the use of natural gas, including comparing the emissions from natural gas to other fossil fuels. Finally, the book delves into some other peripheral but important areas such as automation, operation, simulation, asset management, optimization and project management. The unique requirements of the natural gas industries in these areas are discussed and presented in the context of state of the art techniques and technology applications.

This book is a major contribution to the professional literature in this very crucial area of the emerging energy scene. This book will be very useful to upstream natural gas engineers, researchers, and, especially, to engineers working in the mid-stream and down-stream sectors of natural gas utilization. Because a great deal of research materials have also been included, the book can also become a textbook for graduate university courses or advanced industry training.

Professor Michael J. Economides
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PREFACE

Natural gas is an emergent fuel of choice for environmentally aware due to the lower noxious emissions compared with other fossil fuels. Exploration activity by major multinational oil and gas companies is aimed increasingly to find gas in remote locations and in ever deeper ocean depths. Once a gas field has been discovered, the gas accumulation must be developed, produced, gathered, processed, and transported to the consumer. Transport of gas by pipelines to distant delivery points presents unique challenges of flow through long conduits. Processing of the gas to meet delivery specifications is required and this requires process systems design for each production facility.

With this book we have sought to provide a comprehensive technical text that reflects a broad spectrum of natural gas transmission and processing experience. This handbook has been written to assist both the novice and the experienced technical professional in natural gas engineering practice. Emphasis is placed on developing the fundamental concepts and tools of the subject and on discussing basic applications for engineering purposes. To keep things interesting for the expert, we have sought to include the most current developments reported in the latest published works in the research journals. The selection of some of the material also reflects our own research involvement in problems related to natural gas engineering. We have also taken advantage of some authoritative publications and papers published in the recent years from different sources as well as standards and recommendations published by several research institutions. Therefore, readers are strongly recommended to study the references indicated in the text to find more details on each subject.

We hope you find the book comprehensive, clear and useful. If, through the use of this book, the engineer at study or in practice can gain a clearer insight into the solution of certain practical operating or design problems, which may arise from day to day, the authors will feel that they have been well repaid for their efforts.

Saeid Mokhatab
William A. Poe
James G. Speight

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This book could not have been written without the help of many people whom I have had the good fortune to meet during the period I have spent as an active participant in the rectification operation. Justice demands that their coauthorship be recognized. First, I to acknowledge the coauthorship of Dr. Thomas Wines, Senior Marketing Manager of Pall Corporation, East Hills, New York, USA, in writing the coalescer section of the phase separation chapter and Dr. Wayne Monnery, President of Chem-Pet Process Tech, Calgary, Alberta, Canada, for giving permission to reproduce his material in the three-phase separators designing section of the aforementioned chapter, where I will always be grateful to them for this. I express my sincere gratitude to Dr. Rainer Kurz, Manager of Systems Analysis and Field Testing of Solar Turbines Inc. in San Diego, CA, USA, for his great contribution in writing the gas compression chapter of the book.

Words cannot express my special thanks to Dr. Kosta J. Leontaritis, President and CEO of AsphWax, Inc. in Houston, TX, USA, for using the results of his research studies, which are reported in parts or whole throughout the wax section of the book. I owe a debt of gratitude to Mr. Tony Cleveland, President of Cleveland Engineering Services, Ltd. in Calgary, Alberta, Canada, who has contributed in writing the compressor stations section of the book. I particularly thank Professor Bill Svrcek and Professor Mark Trebble of the University of Calgary for their constractive comments and suggestions. I am specially grateful to Professor Michael J. Economides who accepted to write its foreword. I also express my sincere thanks and appreciation to all following peoples for the technical assistance, detailed comments, and suggestions received from them during the preparation of this book: **Mr. Roger Fernandez**, Team Leader of Natural Gas STAR Program, U.S. Environmental Protection Agency (EPA), Washington, DC, USA; **Mr. Sidney P. Santos**, Senior Consultant, Petrobras Gas & Energy, Rio de Janeiro, Brazil; **Dr. Robert J. Wilkens**, Assistant Professor of Chemical

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Petroleum Engineering Department, University of Wyomonig, Laramie, WY, USA; **Dr. Srdjan Nesic**, Professor of Chemical Engineering and Director of Institute for Corrosion and Multiphase Flow Technology, Ohio University, Athens, OH, USA; **Dr. Hossein Golshan**, Senior Engineer, Transcanada Pipelines Ltd., Calgary, Alberta, Canada; **Mr. Craig S. Alexander**, Lead Consultant, UTSI International, Texas, USA; **Dr. R.A. Cottis**, Editor of Journal of Corrosion Science & Engineering and Reader in Corrosion and Protection Centre, UMIST, Manchester, UK; **Mr. Alok Chakraborty**, Deputy Superintending Engineer, Oil & Natural Gas Corporation Limited, Gujarat, India; **Dr. Michael Nikolaou**, Professor of Chemical Engineering, University of Houston, Houston, TX, USA; **Dr. Peter Breithaupt**, Sr. Business Advisor, Shell Gas & Power Intl. B.V., The Hague, The Netherlands; **Dr. James F. Lea**, Kerr McGee Professor of Petroleum Engineering, University of Oklahoma, Norman, OK, USA; **Dr. Christine Ehlig-Economides**, Professor and Holder of the Albert B. Stevens Endowed Chair in Petroleum Engineering, Texas A&M University, TX, USA; and **Dr. Rashid Hasan**, Professor and Head of Chemical Engineering Department, University of Minnesota-Duluth, Duluth, MN, USA.

The usual absolution applies: none of these individuals is responsible for errors in facts or judgement that may remain.

If this book helps in further developments in the gas industry, all of our assistants named above have extensively influenced the achievement of this goal.

Saeid Mokhatab
31/December/2005

ENDORSEMENTS

We are entering into an economic cycle where natural gas is going to play the leading role as energy source than what it played in the past century. It has been called the Gas Economy. This book intends to educate in the areas of gas transmission and processing discussing new technology in these areas. I recommend that you consider this practical book for quick reference in your day-to-day business.

Jorge H. Foglietta
Director of Technology Development and Process Engineering
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This is an excellent document that combines all facets of engineering within the Oil and Gas business from source to supply. As well as supporting our engineering design industry this work will offer a comprehensive education to our process engineers of the future.

Dr John H. Hargreaves
Chief Process Engineer
KBR Production Services, Aberdeen, UK

Natural gas is in the ascendant as a green fuel of the future. Gas usage will continue to increase worldwide in the coming years. In this regard it is fitting that technical books covering gas processing and transport will be required reading for technical specialists. This highly readable book contains a mine of information on gas transmission and processing. It is recommended for the layman and the technical specialist.

Dr. Satinder Purewal
Leader - Business and Leadership,
Shell International Exploration and Production B.V., The Netherlands

Natural gas will not only continue to be a significant part of the energy resource for many years to come, but its use will also be increasing. This book documents the latest technology in all areas of natural gas engineering. It is a useful book for every engineer, scientist, and researcher who has ever faced the challenge of investigating natural gas gathering, processing, and transport. I recommend it highly, as a reference and textbook.

Dr. Brian F. Towler
Head, Department of Chemical and Petroleum Engineering
University of Wyoming, USA

This book is a useful reference to have for producers, transporters, processors and commercial users. It covers all technical aspects of natural gas transmission and processing. I recommend that if you work in the natural gas area, you have this reference available.

Dr. James F. Lea
Kerr McGee Professor of Petroleum Engineering
University of Oklahoma, USA

This book covers subject areas missed by other books on natural gas engineering that have emphasized mainly the upstream aspects. The emergence of natural gas in a global market underscores the importance of transmission and processing, and I highly recommend this book as a technical resource.

Dr. Christine A. Ehlig-Economides
Professor, Albert E. Stevens Endowed Chair in Petroleum Engineering
Texas A&M University, USA

The ongoing emergence of natural gas as the premier fuel of the world economy makes this book eminently topical and useful to all professionals working along the natural gas chain from exploration and production to the midstream and downstream. It fills a considerable void.

Dr. Michael J. Economides
Professor of Chemical Engineering
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Saeid Mokhatab is a senior gas-engineering consultant specializing in design and operations of natural gas transmission pipelines and processing plants. He has participated in several international projects related to his areas of specialization, and has published more than 50 academic and industrial oriented papers, reports, and books. Mokhatab serves as a member of the petroleum engineering editorial advisory board for Gulf Professional Publishing, an imprint of Elsevier Inc., in North America and is a member of the editorial board for *Journal of Petroleum Science and Technology*, a contributing editor for Hydrocarbon Processing and World Oil magazine, an associate editor for the *International Journal of Modeling and Simulation*, a technical editor for *SPE Projects, Facilities and Construction Journal*, a member of the editorial advisory board for Chemical Engineering Progress magazine in the areas of process control, simulation, and troubleshooting, as well as a referee in several scientific journals and international pipeline conferences. He has also been a guest editor for *ASME Journal of Energy Resources Technology* for the period of August 2003 to March 2006. Mokhatab served on the board of SPE London Section during 2003-6 and is currently a member of SPE Distinguished Lecturer Committee, Offshore Technical Committee for Pipeline Systems Division (PSD) of ASME's International Petroleum Technology Institute, ASME/Offshore Technology Conference's General Program Committee, ASCE Pipeline Research Committee, and SPE.

William A. Poe is a business consultant for Invensys Process Systems - Global Solutions in Houston, Texas, USA. He has over 24 years of experience in chemical and gas processing plants design and operations and has developed business relationships with several domestic and international gas processors, chemical plants, and gas transmission pipelines owners. His main area of focus the past decade and a half has been automation, multivariable predictive control (MPC), advanced process control (APC), and optimization design and implementation for the natural gas processing and treating, sulfur recovery, ammonia synthesis, methanol synthesis, liquefied natural gas, and power generation industries. Working under contracts to the Gas Research Institute, he developed new multivariable control applications in the natural gas, ammonia, and methanol industries. Poe developed APC and Optimization Master Plans for international companies such as Saudi Aramco, Statoil, and PDVSA, as well as automation and advanced process control feasibility studies for about 100 natural gas processing, power generation, ammonia, and methanol plants worldwide. He spent over a decade in natural gas processing operations and engineering experience with Shell Oil Company. Poe has project management experience with Shell Oil and GE Continental Controls with multimillion-dollar projects. He assumed the technical leadership of Continental Controls, a privately held company, through acquisition by GE. After joining GE as part of the Continental Controls acquisition, he became vice president of this division of GE, which included direction of product development, projects, technical sales support, and customer service. He has authored over 25 papers, presented at numerous technical conferences, and received the GE Innovators Award in 1999.

James G. Speight is a senior fuel consultant as well as an adjunct professor of chemical and fuels engineering at the University of Utah, USA. He is recognized internationally as an expert in the characterization, properties, and processing of conventional and synthetic fuels and as a chemist with more than 35 years of experience in thermal/process chemistry, thermodynamics, refining of petroleum, heavy oil, and tar sand bitumen, and physics of crude with emphasis on distillation, visbreaking, coking units, and oil-rock or oil catalyst interactions. Speight is currently editor in chief for the *Journal of Petroleum Science and Technology, Energy Sources-Part A: Recovery, Utilization, and Environmental Effects,* and *Energy Sources-Part B: Economics, Planning, and Policy*. He is also the author/editor/compiler of more than 25 books and bibliographies related to fossil fuel processing and environmental issues.

Speight was chief scientific officer and then chief executive officer of the Western Research Institute, Laramie, WY, USA, from 1984 to 2000. During this period he led a staff of more that 150 scientists, engineers, and technicians in developing new technology for gas processing, petroleum, shale oil, tar sand bitumen, and asphalt. Speight has considerable expertise in evaluating new technologies for patentability and commercial application. As a result of his work, he was awarded the Diploma of Honor, National Petroleum Engineering Society, for outstanding contributions to the petroleum industry in 1995 and the Gold Medal of Russian Academy of Sciences (Natural) for outstanding work in the area of petroleum science in 1996. He has also received the Specialist Invitation Program Speakers Award from NEDO (New Energy Development Organization, Government of Japan) in 1987 and again in 1996 for his contributions to coal research. In 2001, he was also awarded the Einstein Medal of the Russian Academy of Sciences (Natural) in recognition of outstanding contributions and service in the field of geologic sciences. More recently (2005) he has been awarded the Gold Medal-Scientists without Frontiers, Russian Academy of Sciences in recognition of continuous encouragement of scientists to work together across international borders.

CHAPTER 1

NATURAL GAS FUNDAMENTAL

1.1 INTRODUCTION

Natural gas is used primarily as a fuel and as a raw material in manufacturing. It is used in home furnaces, water heaters, and cooking stoves. As an industrial fuel, it is used in brick, cement, and ceramic-tile kilns; in glass making; for generating steam in water boilers; and as a clean heat source for sterilizing instruments and processing foods. As a raw material in petrochemical manufacturing, natural gas is used to produce hydrogen, sulfur, carbon black, and ammonia. The ammonia is used in a range of fertilizers and as a secondary feedstock for manufacturing other chemicals, including nitric acid and urea. Ethylene, an important petrochemical, is also produced from natural gas.

Natural gas is considered as an environmentally friendly clean fuel, offering important environmental benefits when compared to other fossil fuels. The superior environmental qualities over coal or crude oil are that emissions of sulfur dioxide are negligible or that the levels of nitrous oxide and carbon dioxide emissions are lower. This helps to reduce problems of acid rain, ozone layer, or greenhouse gases. Natural gas is also a very safe source of energy when transported, stored, and used.

This chapter covers, the essential technical points of natural gas. The objectives of the chapter are to give the reader an introduction to natural gas by describing the origin and composition of natural gas, gas sources, phase behavior and properties, and transportation methods.

1.2 NATURAL GAS HISTORY

The discovery of natural gas dates from ancient times in the Middle East. Thousands of years ago, it was noticed that natural gas seeps ignited by lightning created burning springs. In Persia, Greece, or India, people built temples around these eternal flames for their religious practices. However, the energy value of natural gas was not recognized until approximately 900 BC in China, and the Chinese drilled the first known natural gas well in 211 BC. In Europe, natural gas was unknown until it was discovered in Great Britain in 1659, although it was not commercialized until about 1790. In 1821 in Fredonia, United States, residents observed gas bubbles rising to the surface from a creek. William Hart, considered as America's father of natural gas, dug the first natural gas well in North America (Speight, 1993).

Historically, natural gas was discovered as a consequence of prospecting for crude oil. Natural gas was often an unwelcome by-product, as natural gas reservoirs were tapped in the drilling process and workers were forced to stop drilling to let the gas vent freely into the air. Now, and particularly after the crude oil shortages of the seventies, natural gas has become an important source of energy in the world. Throughout the 19th century, natural gas was used almost exclusively as a source of light and its use remained localized because of lack of transport structures, making it difficult to transport large quantities of natural gas long distances. There was an important change in 1890 with the invention of leak-proof pipeline couplings, but transportation of natural gas to long-distance customers did not become practical until the 1920s as a result of technological advances in pipelines. Moreover, it was only after World War II that the use of natural gas grew rapidly because of the development of pipeline networks and storage systems.

1.3 NATURAL GAS ORIGIN AND COMPOSITION

Natural gas exists in nature under pressure in rock reservoirs in the Earth's crust, either in conjunction with and dissolved in heavier hydrocarbons and water or by itself. It is produced from the reservoir similarly to or in conjunction with crude oil. Natural gas has been formed by the degradation of organic matter accumulated in the past millions of years. Two main mechanisms (biogenic and thermogenic) are responsible for this degradation (Rojey *et al.*, 1997). Biogenic gas is formed at shallow depths and low temperatures by the anaerobic bacterial decomposition of sedimentary organic matter. In contrast, thermogenic gas is formed at deeper depths by (1) thermal cracking of sedimentary organic matter into hydrocarbon liquids and gas (this gas is cogenetic with oil and is called "primary" thermogenic gas) and (2) thermal cracking of oil at high temperatures into gas ("secondary" thermogenic gas) and pyrobitumen. Biogenic gas consists almost entirely of methane. In contrast, thermogenic gas can also contain significant concentrations of ethane, propane, butanes, and heavier hydrocarbons. Knowing whether a natural gas show is biogenic gas or thermogenic gas can have critical implications for the presence of liquid hydrocarbons in a basin. Gas geochemistry readily reveals whether a gas is biogenic or thermogenic.

The principal constituent of natural gas is methane. Other constituents are paraffinic hydrocarbons such as ethane, propane, and the butanes. Many natural gases contain nitrogen as well as carbon dioxide and hydrogen sulfide. Trace quantities of argon, hydrogen, and helium may also be present. The composition of natural gas can vary widely. Table 1-1 outlines the typical makeup of natural gas before it is refined. Natural gas can also contain a small proportion of C_5^+ hydrocarbons. When separated, this fraction is a light gasoline. Some aromatics such as benzene, toluene, and xylenes can also be present, raising safety issues due to their toxicity. Natural gas can contain other contaminants too. Acid contaminants such as mercaptans—R-SH, carbonyl sulfide (COS), and carbon disulfide (CS_2)—may be present in small quantities. Mercury can also be present either as a metal in vapor phase or as an organometallic compound in

Table 1-1
Typical Composition of Natural Gas

Name	Formula	Volume (%)
Methane	CH_4	>85
Ethane	C_2H_6	3–8
Propane	C_3H_8	1–2
Butane	C_4H_{10}	<1
Pentane	C_5H_{12}	<1
Carbon dioxide	CO_2	1–2
Hydrogen sulfide	H_2S	<1
Nitrogen	N_2	1–5
Helium	He	<0.5

liquid fractions. Concentration levels are generally very small, but even at very small concentration levels, mercury can be detrimental due its toxicity and its corrosive properties (reaction with aluminium alloys).

Natural gas is considered "dry" when it is almost pure methane, having had most of the other commonly associated hydrocarbons removed. When other hydrocarbons are present, the natural gas is "wet".

The composition of natural gas varies depending on the field, formation, or reservoir from which it is extracted. Since the composition of natural gas is never constant, there are standard test methods by which the composition of natural gas can be determined and thus prepared for use. However, it is not the intent to cover the test methods in any great detail as they are available elsewhere (Speight, 2002).

1.4 GAS SOURCES

Natural gas produced from geological formations comes in a wide array of compositions. The varieties of gas compositions can be broadly categorized into three distinct groups: (1) nonassociated gas that occurs in conventional gas fields, (2) associated gas that occurs in conventional oil fields, and (3) continuous (or unconventional) gas. Some types of unconventional gas resources include "tight gas" or "tight sands gas," which is found in low-permeability rock; "coalbed methane (CBM)," which is natural gas that has been formed along with the geological processes that formed coal; "natural gas from geopressurized aquifers," which refers to gas dissolved under high pressure and at high temperatures in brines located deep beneath the Earth's surface; "gas hydrates," which are ice-like structures of water and gas located under the permafrost; and "deep gas," which is found at levels much deeper than conventional gas. Although there is no scientific consensus, some believe deep gas originated from inorganic sources and that it exists everywhere as a result of the geological processes that formed the earth. Of the unconventional gas sources, the one most important to the gas transportation industry is coal bed methane.

1.4.1 Non-associated Gas

Nonassociated gas (sometimes called "gas well gas") is produced from geological formations that typically do not contain much, if any, higher boiling hydrocarbons (gas liquids) than methane. Nonassociated gas can contain nonhydrocarbon gases such as carbon dioxide and hydrogen sulfide. Nonassociated gas is directly controllable by the producer; one just turns the valves. The gas flows up the well under its own energy, through the wellhead control valves, and along the flow line to the treatment plant. Treatment requires the temperature of the gas to be reduced to a point dependent upon the pressure in the pipeline so that all liquids that would exist at pipeline temperature and pressure condense and are removed.

1.4.2 Associated Gas

Associated gas is produced during crude oil production and is the gas that is associated with crude oil. Crude oil cannot be produced without producing some of its associated gas, which comes out of solution as the pressure is reduced on the way to and on the surface. Properly designed crude oil well completions and good reservoir management are used to minimize the production of associated gas so as to retain the maximum energy in the reservoir and thus increase ultimate crude oil recovery. Crude oil in the reservoir with minimal or no dissolved associated gas is rare and as dead crude oil is often difficult to produce as there is little energy to drive it.

After the production fluids are brought to the surface, they are separated at a tank battery at or near the production lease into a hydrocarbon liquid stream (crude oil or *gas condensate*), a produced water stream (brine or salty water), and a gaseous stream. The gaseous stream is traditionally very rich (*rich gas*) in *natural gas liquids* (NGLs). Natural gas liquids include ethane, propane, butanes, and pentanes and higher molecular weight hydrocarbons (C_6^+). The higher molecular weight hydrocarbons product is commonly referred to as *natural gasoline*.

When referring to natural gas liquids in the natural gas stream, the term *gallons per thousand cubic feet* of gas is used as a measure of hydrocarbon richness. Depending on its content of heavy components, natural gas can be considered as rich (5 or 6 gallons or more of recoverable hydrocarbons per cubic feet) or lean (less than 1 gallon of recoverable hydrocarbons per cubic feet). However, the terms *rich gas* and *lean gas*, as used in the gas-processing industry, are not precise indicators of gas quality but only indicate the relative amount of natural gas liquids in the gas stream.

In the case of associated gas, crude oil may be assisted up the well bore by gas lift (Speight, 1993). Thus, gas is compressed into the annulus of the well and then injected by means of a gas lift valve near the bottom of the well into the crude oil column in the tubing. At the top of the well the crude oil and gas mixture passes into a separation plant that drops the pressure down to nearly atmospheric in two stages. The crude oil and water exit the bottom of the lower pressure separator, from where it is pumped to tanks for separation of the crude oil and water. The gas produced in the separators and the gas that comes out of solution with the produced crude oil (surplus gas) are then treated to separate out the NGLs that are treated in a gas plant to provide propane and butane or a mixture of the two [liquefied petroleum gas[1], LPG]. The higher boiling residue, after the propane and butane are removed, is condensate, which is mixed with the crude oil or exported as a separate product. The gas itself is then *dry* and, after compression, is suitable to be injected into the natural gas system where it substitutes for natural gas from the nonassociated gas reservoir. Pretreated associated gas from other fields enters the system at this stage (Manning and Thompson, 1991). Another use for the gas is as fuel for the gas turbines on site. This gas is treated in a fuel gas plant to ensure that it is clean and at the correct pressure. The start-up fuel gas supply will be from the main gas system, but facilities exist to collect and treat low-pressure gas from the various other plants as a more economical fuel source.

Worldwide, governments are mandating that producers stop flaring associated gas, as their citizens perceive that it is a waste of a valuable nonrenewable resource. There are often regulatory restrictions on when produced gas can be reinjected, or flared, with an understanding that any reinjected gas must eventually be produced. When such restrictions occur, oil production must be stopped until this associated gas can somehow be exported or reinjected.

1.4.3 Coal Bed Methane

Coal bed methane is the generic term given to methane gas held in coal and released or produced when the water pressure within the buried coal

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[1] LPG is a hydrocarbon mixture containing propane and *n*-butane. To a lesser extent, *iso*-butane may also be present. The most common commercial products are propane, butane, or some mixture of the two and are generally extracted from natural gas or crude petroleum.

is reduced by pumping from either vertical or inclined to horizontal surface holes. The methane is predominantly formed during the coalification process whereby organic matter is slowly transformed into coal by increasing temperature and pressure as the organic matter is buried deeper and deeper by additional deposits of organic and inorganic matter over long periods of geological time. This is referred to as thermogenic coal bed methane. Alternatively, and more often (but not limited to) in lower rank and thermally immature coals, recent bacterial processes (involving naturally occurring bacteria associated with meteoric water recharge at outcrop or subcrop) can dominate the generation of CBM. This is referred to as late-stage biogenic coal bed methane.

During the coalification process, a range of chemical reactions takes place that produce substantial quantities of gas. While much of this gas escapes into the overlying or underlying rock, a large amount is retained within the forming coal seams. However, unlike conventional natural gas reservoirs, where gas is trapped in the pore or void spaces of a rock, such as sandstone, methane formed and trapped in coal is actually adsorbed onto the coal grain surfaces or micropores and held in place by reservoir (water) pressure. Therefore, because the micropore surface area is very large, coal can potentially hold significantly more methane per unit volume than most sandstone reservoirs.

The amount of methane stored in coal is closely related to the rank and depth of the coal; the higher the coal rank and the deeper the coal seam is presently buried (causing pressure on coal) the greater its capacity to produce and retain methane gas. Because coal has a very large internal surface area of over 1 billion square feet per ton of coal, it can hold on average three times as much gas in place as the same volume of a conventional sandstone reservoir at equal depth and pressure. In order to allow the “absorbed” gas to be released from the coal it is often necessary to lower the pressure on the coal. This generally involves removing the water contained in the coal bed. After the gas is released from the internal surfaces of the coal it moves through the internal matrix of the coal until it reaches natural fracture networks in the coal known as cleats. The gas then flows through these cleats or fractures until it reaches the well bore.

Gas derived from coal is generally pure and requires little or no processing because it is solely methane and not mixed with heavier hydrocarbons, such as ethane, which is often present in conventional natural gas. Coal bed methane has a slightly higher energy value than some natural gases.

1.5 NATURAL GAS PHASE BEHAVIOR

The natural gas phase behavior is a plot of pressure vs temperature that determines whether the natural gas stream at a given pressure and temperature consists of a single gas phase or two phases: gas and liquid. The phase behavior for natural gas with a given composition is typically displayed on a phase diagram, an example of which is shown in Figure 1-1. The left-hand side of the curve is the bubble point line and divides the single-phase liquid region from the two-phase gas–liquid region. The right-hand side of the curve is the dew point line and divides the two-phase gas–liquid region and the single-phase gas region. The bubble point and dew point lines intersect at the critical point, where the distinction between gas and liquid properties disappears. Note that dew point temperatures are possible at a given pressure, and two dew point pressures are possible at a given temperature. This phenomenon is known as retrograde condensation. The maximum pressure at which liquids can form (P_{max}) is called the cricondenbar, and the maximum temperature at which liquids can form (T_{max}) is called the cricondentherm.

The natural gas phase behavior is a function of the composition of the gas mixture and is strongly influenced by the concentration of the heavier hydrocarbons, especially C_6^+. The presence of heavier hydrocarbons will increase the phase envelope and failure to include them in a phase calculation will underpredict the phase envelope.

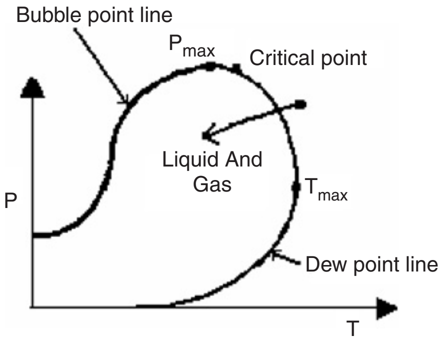


Figure 1-1. Pressure-temperature diagram for a typical natural gas mixture.

1.6 NATURAL GAS PROPERTIES

1.6.1 Chemical and Physical Properties

Natural gas is colorless, odorless, tasteless, shapeless, and lighter than air (see Table 1-2). The natural gas after appropriate treatment for acid gas reduction, odorization, and hydrocarbon and moisture dew point adjustment would then be sold within prescribed limits of pressure, calorific value, and possibly *Wobbe index* (often referred to as the *Wobbe number*). The Wobbe index (calorific value divided by the specific gravity) gives a measure of the heat input to an appliance through a given aperture at a given gas pressure. Using this as a vertical coordinate and the flame speed factor as the horizontal coordinate, a combustion diagram can be constructed for an appliance, or a whole range of appliances, with the aid of appropriate test gases. This diagram shows the area within which

Table 1-2
Properties of Natural Gas

Properties	Value
Relative molar mass	17–20
Carbon content, weight %	73.3
Hydrogen content, weight %	23.9
Oxygen content, weight %	0.4
Hydrogen/carbon atomic ratio	3.0–4.0
Relative density, 15°C	0.72–0.81
Boiling point, °C	−162
Autoignition temperature, °C	540–560
Octane number	120–130
Methane number	69–99
Stoichiometric air/fuel ratio, weight	17.2
Vapour flammability limits, volume %	5–15
Flammability limits	0.7–2.1
Lower heating/calorific value, MJ/kg	38–50
Stoichiometric lower heating value, MJ/kg	2.75
Methane concentration, volume %	80–99
Ethane concentration, volume %	2.7–4.6
Nitrogen concentration, volume %	0.1–15
Carbon dioxide concentration, volume %	1–5
Sulfur concentration, weight % ppm	<5
Specific CO_2 formation, g/MJ	38–50

variations in the Wobbe index and flame speed factor of gases may occur for the given range of appliances without resulting in incomplete combustion, flame lift, or the lighting back of preaerated flames. This method of prediction of combustion characteristics is not sufficiently accurate to eliminate entirely the need for the practical testing of new gases.

Since natural gas as delivered to pipelines has practically no odor, the addition of an odorant is required by most regulations in order that the presence of the gas can be detected readily in case of accidents and leaks. This odorization is provided by the addition of trace amounts of some organic sulfur compounds to the gas before it reaches the consumer. The sulfur compound, a chemical odorant (a *mercaptan* also called a *thiol* with the general formula R-SH and the odor of rotten eggs), is added to natural gas so that it can be smelled if there is a gas leak. The standard requirement is that a user will be able to detect the presence of the gas by odor when the concentration reaches 1% of gas in air. Since the lower limit of flammability of natural gas is approximately 5%, this requirement is equivalent to one-fifth the lower limit of flammability. The combustion of these trace amounts of odorant does not create any serious problems of sulfur content or toxicity.

The following section discusses important gas properties, including specific gravity, compressibility factor, formation volume factor, density, isothermal compressibility, and viscosity.

1.6.2 Gas-Specific Gravity

Specific gravity of gas is defined as

$$\gamma_{\mathrm{g}} = \frac{M}{M_{\mathrm{air}}} \tag{1-1}$$

where M_{air} is the molecular weight of air, which is equal to 29. Once we can calculate the value of the molecular weight of the mixture, we can calculate the specific gravity of the mixture. For a gas mixture, we can calculate the molecular weight as

$$M = \sum_{i=1}^{n} y_{\mathrm{i}} M_{\mathrm{i}} \tag{1-2}$$

where M_i is the molecular weight of component i, y_i is the mole fraction of component i, and n is the total number of components.

Various gas properties, including the molecular weights for pure components, are given in Table 1-3.

1.6.3 Ideal and Real Gas Laws

The volume of a real gas is usually less than what the volume of an ideal gas would be, and hence a real gas is said to be supercompressible. The ratio of the real volume to the ideal volume, which is a measure of the amount the gas deviates from perfect behavior, is called the supercompressibility factor, sometimes shortened to the compressibility factor. It is also called the gas deviation factor and is given the symbol Z. The gas deviation factor is, by definition, the ratio of the volume actually occupied by a gas at a given pressure and temperature to the volume it would occupy if it behaved ideally.

The real gas equation of state is then written as

$$PV = ZnRT \tag{1-3}$$

where P is the pressure, V is the volume, T is the absolute temperature, Z is the compressibility, n is the number of kilomoles of the gas, and R is the gas constant.

The gas deviation factor, Z, is close to 1 at low pressures and high temperatures, which means the gas behaves as an ideal gas at these conditions. At standard or atmospheric conditions the gas Z factor is always approximately 1.

The theory of corresponding states dictates that the Z factor can be uniquely defined as a function of reduced pressure and reduced temperature. The reduced pressure and reduced temperature are defined as

$$P_r = \frac{P}{P_c} \text{ and } T_r = \frac{T}{T_c} \tag{1-4}$$

where P_r and T_r are reduced pressure and reduced temperature, respectively, and P_c and T_c are critical pressure and critical temperature of the gas, respectively. The values of critical pressure and critical temperature can be estimated from the following equations if the composition

Table 1-3
Physical Constants for Pure Components (Whitson and Brule, 2000)

Compound	Formula	Molecular weight	Critical constants			
			P_c psia [kp_a]	T_c °K [°K]	V_c ft^3/lb [m^3/kg]	Z_c
Methane	CH_4	16.043	667.8[4,604]	343[190.6]	0.0991[0.0062]	0.28884
Ethane	C_2H_6	30.070	707.8[4,880]	549.8[369.8]	0.0788[0.00492]	0.2843
Propane	C_3H_8	44.097	6163[4,249]	665.7[369.8]	0.0737[0.0046]	0.2804
n-Butane	C_4H_{10}	58.124	550.7[3,797]	765.3[425.2]	0.702[0.00438]	0.2736
Isobutane	C_4H_{10}	58.124	529.1[3,648]	734.7[408.2]	0.0724[0.00452]	0.2824
n-Pebtabe	C_5H_{12}	72.151	488.6[3,369]	845.4[469.7]	0.0675[0.00422]	0.2623
Isopentane	C_5H_{12}	72.151	490.4[3,381]	828.8[460.4]	0.0679[0.00424]	0.2701
Neopentane	C_5H_{12}	72.151	464.0[3,199]	781.11	0.0674[0.00421]	0.2537
n-Hexane	C_6H_{14}	86.178	436.9[3,012]	913.4[507.4]	0.0688[0.0043]	0.2643
n-Heptane	C_3H_{16}	100.205	396.8[2,736]	972.5[5403]	0.0691[0.00432]	0.2633
n-Octane	C_8H_{18}	114.232	360.6[2,486]	1,023.9[568.8]	0.0690[0.0043]	0.2587
n-Nonane	C_9H_{20}	128.30	332[2,289]	1,0703[594.6]	0.0684[0.00427]	0.2536
n-Deccane	$C_{10}H_{22}$	142.30	304.[2,096]	1,111.8[617.7]	0.0679[0.00424]	0.2462
Ethylene	C_2H_4	28.054	729.8[5,032]	508.6[282.6]	0.0737[0.0046]	0.2765
Propene	C_3H_6	42.081	699.[4,613]	6569[364.9]	0.0689[0.0043]	0.2752
Acetylene	C_2H_2	26.038	890.4[6,139]	555.3[308.5]	0.0695[0.00434]	0.2704
Carbon dioxide	CO_2	44.010	1071.[7,382]	547.6[304.2]	0.0342[0.00214]	0.2742
Hydrogen sulfide	H_2S	34.076	1306[9,005]	672.4[373.6]	0.0459[0.00287]	0.2831
Sulfur dioxide	SO_2	64.059	1145.[7,894]	775.5[430.8]	0.0306[0.00191]	0.2697
Nitrogen	N_2	28.013	493[3,399]	227.3[126.3]	0.0514[0.00321]	0.2916
Water	H_2O	18.015	3208.[22,105]	1,1650[647.2]	0.0500[0.00312]	0.2350

of the gas and the critical properties of the individual components are known:

$$P_{\mathrm{c}} = \sum_{i}^{n} P_{\mathrm{Ci}} y_{\mathrm{i}} \text{ and } T_{\mathrm{c}} = \sum_{i}^{n} T_{\mathrm{ci}} y_{\mathrm{i}} \tag{1-5}$$

where P_{ci} and T_{ci} are the critical pressure and critical temperature of component i, respectively; and y_{i} is the mole fraction of component i.

Once critical properties of the mixture are calculated as stated in Eq. (1-5), we can use Eq. (1-4) to calculate the reduced properties of the mixture.

The values of critical pressure and critical temperature can be estimated from its specific gravity if the composition of the gas and the critical properties of the individual components are not known. The method uses a correlation to estimate pseudocritical temperature and pseudocritical pressure values from the specific gravity. There are several different correlations available. The most common is the one proposed by Sutton (1985), which is based on the basis of 264 different gas samples. Sutton (1985) used regression analysis on raw data to obtain the following second-order fits for the pseudocritical properties:

$$P_{\mathrm{pc}} = 756.8 - 131.07\gamma_{\mathrm{g}} - 3.6\gamma_{\mathrm{g}}^{2} \tag{1-6}$$

$$T_{\mathrm{pc}} = 169.2 + 349.5\gamma_{\mathrm{g}} - 74.0\gamma_{\mathrm{g}}^{2} \tag{1-7}$$

These equations are valid over the range of specific gas gravities with which Sutton (1985) worked $0.57 < \gamma_g < 1.68$.

The most commonly used method to estimate the Z factor is the chart provided by Standing and Katz (1942). The Z factor chart is shown in Figure 1-2. The chart covers the range of reduced pressure from 0 to 15, and the range of reduced temperature from 1.05 to 3.

The Z factor chart of Standing and Katz (1942) is only valid for mixtures of hydrocarbon gases. Wichert and Aziz (1972) developed a correlation to account for inaccuracies in the Standing and Katz chart when the gas contains significant fractions of acid gases, specifically carbon dioxide (CO_2) and hydrocarbon sulfide (H_2S). The Wichert and Aziz (1972) correlation modifies the values of the pseudocritical temperature and pressure of the gas. Once the modified pseudocritical properties are obtained, they

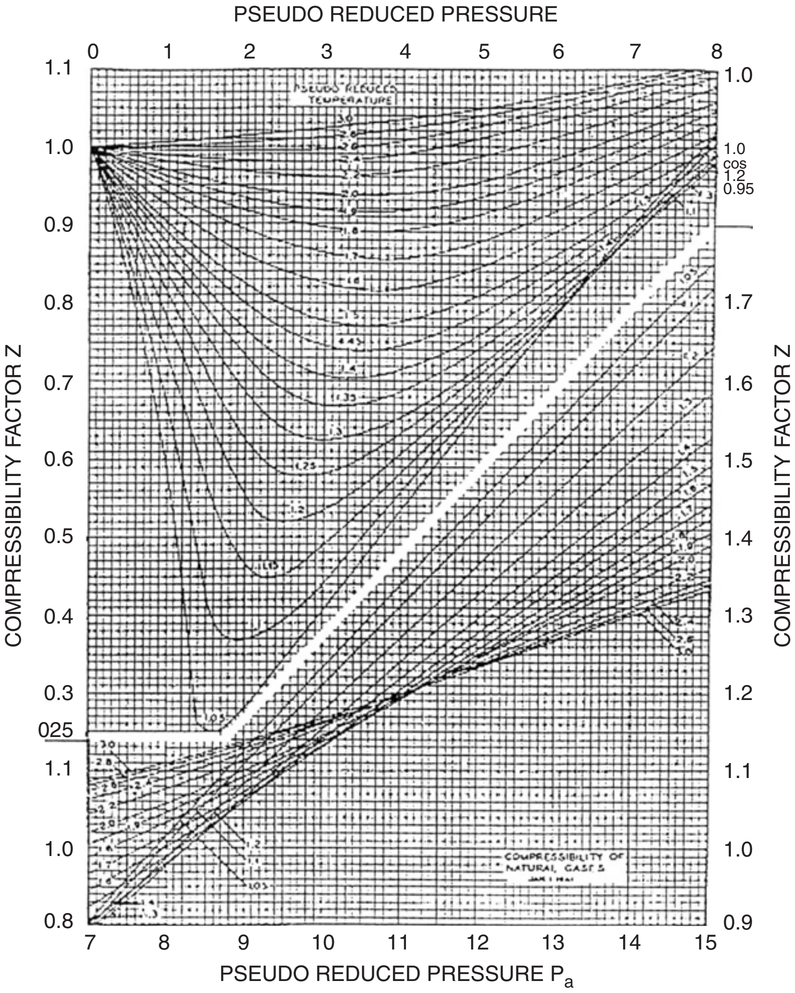


Figure 1-2. Compressibility of natural gases as a function of reduced pressure and temperature (Standing and Katz, 1942).

are used to calculate pseudo-reduced properties and the Z factor is determined from Figure 1-2. The Wichert and Aziz (1972) correlation first calculates a deviation parameter ε:

$$\varepsilon = 120(A^{0.9} - A^{1.6}) + 15(B^{0.5} - B^{4}) \tag{1-8}$$

where A is the sum of the mole fractions of CO_2 and H_2S in the gas mixture and B is the mole fraction of H_2S in the gas mixture.

Then, ε is used to determine the modified pseudocritical properties as follows:

$$T'_{pc} = T_{pc} - \varepsilon \tag{1-9}$$

$$P'_{pc} = \frac{P_{pc} T'_{pc}}{[T_{pc} - B(1 - B)\varepsilon]} \tag{1-10}$$

The correlation is applicable to concentrations of $CO_2 < 54.4$ mol% and $H_2S < 73.8$ mol%. Wichert and Aziz found their correlation to have an average absolute error of 0.97% over the following ranges of data: 154 psia $< P <$ 7026 psia and 40°F $< T <$ 300°F.

1.6.4 Gas Formation Volume Factor

The formation volume factor for gas is defined as the ratio of volume of 1 mol of gas at a given pressure and temperature to the volume of 1 mole of gas at standard conditions (P_s and T_s). Using the real gas law and assuming that the Z factor at standard conditions is 1, the equation for formation volume factor (B_g) can be written as

$$B_g = \frac{V_R}{V_s} = \frac{nZRT}{P} \frac{P_s}{nZ_s RT_s} = \frac{P_s ZT}{T_s P} \tag{1-11}$$

when P_s is 1 atmosphere (14.6959 psia or 101.325 kPa) and T_s is 60°F (519.67°R or 288.71°K), this equation can be written in three well-known standard forms:

$$B_g = 0.0283 \frac{ZT}{P} \tag{1-11-1}$$

where B_g is in ft^3/SCF, P is in psia, and T is in °R. Alternately,

$$B_g = 0.3507 \frac{ZT}{P} \tag{1-11-2}$$

where B_g is in m^3/Sm^3, P is in KPa, and T is in °K.

In some cases, it is more convenient to define the value of B_g in bbl/SCF. The equation can be written as

$$B_g = 0.005\frac{ZT}{P} \tag{1-11-3}$$

where T is in °R and P is in psia.

1.6.5 Gas Density

Gas density is defined as mass per unit volume and so can also be derived and calculated from the real gas law:

$$\rho_g = \frac{m}{V} = \frac{PM}{ZRT} \tag{1-12}$$

Knowing that the molecular weight of gas is the product of specific gravity and molecular weight of air and that the value of R is 10.73 in field units [8.314 in SI units], we can write the equation for density as

$$\rho_g = 2.7\frac{P\gamma_g}{ZT} \tag{1-12-1}$$

where ρ_g is in lbm/ft^3, P is in psia, and T is in °R. Alternately,

$$\rho_g = 3.49\frac{P\gamma_g}{ZT} \tag{1-12-2}$$

where ρ_g is in kg/m^3, P is in kPa, and T is in °K.

The density can also be written as

$$\rho_g = 0.0764\frac{\gamma_g}{B_g} \tag{1-12-3}$$

where ρ_g is in lbm/ft^3 and B_g is in ft^3/SCF. Alternately,

$$\rho_g = 1.224\frac{\gamma_g}{B_g} \tag{1-12-4}$$

where ρ_g is in kg/m^3, and B_g is in m^3/Sm^3.

1.6.6 Isothermal Compressibility of Gases

The isothermal gas compressibility, which is given the symbol c_g, is a useful concept is used extensively in determining the compressible properties of the reservoir. The isothermal compressibility is also called the bulk modulus of elasticity. Gas usually is the most compressible medium in the reservoir. However, care should be taken so that it is not confused with the gas deviation factor, Z, which is sometimes called the supercompressibility factor:

$$c_g = -\frac{1}{V_g}\left(\frac{\partial V_g}{\partial P}\right)_T \tag{1-13}$$

where V and P are volume and pressure, respectively, and T is the absolute temperature. For ideal gas, we can define the compressibility as

$$c_g = \frac{1}{P} \tag{1-13-1}$$

whereas, for nonideal gas, compressibility is defined as

$$c_g = \frac{1}{P} - \frac{1}{Z}\left(\frac{\partial Z}{\partial P}\right)_T \tag{1-13-2}$$

If the relationship between the Z factor and the pressure at a given temperature is known, we can calculate the compressibility of gas. Since we already know the relationship between Z and P, we can numerically calculate the derivative and, hence, the value of the compressibility.

1.6.7 Gas Viscosity

Just as the compressibility of natural gas is much higher than that of oil, water, or rock, the viscosity of natural gas is usually several orders of magnitude lower than oil or water. This makes gas much more mobile in the reservoir than either oil or water. Reliable correlation charts are available to estimate gas viscosity, and the viscosity of gas mixtures at one atmosphere and reservoir temperature can be determined from the gas mixture composition:

$$\mu_{\mathrm{ga}} = \frac{\sum_{i=1}^{N} y_{\mathrm{i}} \mu_{\mathrm{i}} \sqrt{M_{\mathrm{gi}}}}{\sum_{i=1}^{N} y_{\mathrm{i}} \sqrt{M_{\mathrm{gi}}}} \tag{1-14}$$

where μ_{ga} is the viscosity of the gas mixture at the desired temperature and atmospheric pressure, y_{i} is the mole fraction of the *i*th component, μ_{ga} is the viscosity of the *i*th component of the gas mixture at the desired temperature and atmospheric pressure, M_{gi} is the molecular weight of the *i*th component of the gas mixture, and N is the number of components in the gas mixture. This viscosity is then multiplied by the viscosity ratio to obtain the viscosity at reservoir temperature and pressure.

1.7 QUALITY

The amount of energy that is obtained from the burning of a volume of natural gas is measured in British thermal units (Btu). The value of natural gas is calculated by its Btu content. The energy content of natural gas is variable and depends on its accumulations, which are influenced by the amount and types of energy gases they contain: the more noncombustible gases in a natural gas, the lower the Btu value. In addition, the volume mass of energy gases that are present in a natural gas accumulation also influences the Btu value of natural gas. The more carbon atoms in a hydrocarbon gas, the higher its Btu value. Btu analyses of natural gas are done at each stage of the supply chain. Gas chromatographic process analyzers are used in order to conduct fractional analysis of the natural gas streams, separating natural gas into identifiable components. The components and their concentrations are converted into a gross heating value in Btu-cubic foot.

Normally, natural gas as it is when extracted is not suitable for pipeline transportation or commercial use before being processed. Pipelines set their specifications for the quality of natural gas. In any case, natural gas must be processed in order to remove unwanted water vapor, solids, or other contaminants and to get those hydrocarbons that have a higher value as separate products.

1.8 TRANSPORTATION

Gas, as a result of the storage difficulties, needs to be transported immediately to its destination after production from a reservoir (Cranmore and Stanton, 2000a). There are a number of options for transporting natural gas energy from oil and gas fields to market (Rojey *et al*., 1997; Thomas and Dawe, 2003). These include pipelines, liquefied natural gas (LNG), compressed natural gas (CNG), gas to solids (GTS), i.e., hydrates, gas to power (GTP), i.e., electricity, and gas to liquids (GTL), with a wide range of possible products, including clean fuels, plastic precursors, or methanol and gas to commodity (GTC), such as aluminum, glass, cement, or iron. The following section examines some of these technical methods by which natural gas energy can be transported and covers many of the essential points needed to enter the discussion.

1.8.1 Pipelines

Pipelines are a very convenient method of transport but are not flexible as the gas will leave the source and arrive at its (one) destination (Cranmore and Stanton, 2000b). If the pipeline has to be shut down, the production and receiving facilities and refinery often also have to be shut down because gas cannot be readily stored, except perhaps by increasing the pipeline pressure by some percentage.

In the last decade, on average, over 12,000 miles per year of new gas pipelines have been completed;most are transnational. If political stability can be guaranteed, pipelines may be able to provide a long-term solution for transportation. An example of this approach is a proposed deep water pipeline from Oman to India (EIA, 2002). However, the cost of building such a pipeline remains unclear. Subsea lines over 2000 miles have, until recently, been regarded as uneconomic because of the subsea terrain making pipeline installation and maintenance expensive and any recompression along the route difficult, but changes are in the air! If technical and economic hurdles can be overcome, these pipelines can become effective.

1.8.2 Liquefied Natural Gas

Liquefied natural gas technology has been proven to be effective since the mid-1970s. LNG is the liquid form of natural gas. Gas cooled to approximately −162°C liquefies and has a volume approximately 1/600 that of gas at room temperature. However, facilities for liquefying natural gas require complex machinery with moving parts and special refrigerated ships for transporting the liquefied natural gas to market (Cranmore and Stanton, 2000b). The costs of building a liquefied natural gas plant have lowered since the mid-1980s because of greatly improved thermodynamic efficiencies, making liquefied natural gas a major gas export method worldwide, and many plants are being extended or new ones are being built in the world.

Large cryogenic tanks are needed to store the liquefied natural gas; typically these may be 70 m in diameter, 45 m high, and hold over 100,000 m^3 of liquefied natural gas. At the consumer end, an infrastructure for handling the reprocessing of vast quantities of natural gas from LNG is required, which is also expensive and vulnerable to sabotage.

The current largest specially built refrigerated tankers can carry 135,000 m^3 of liquefied natural gas, equivalent to 2.86 billion scf of gas, but are very expensive. This makes it difficult for liquefied natural gas to use smaller isolated (offshore) reserves and to serve small markets commercially because it is this large capacity and continuous running that keep thermodynamic efficiency high and costs to a minimum. Thus small volumes of intermittent gas are not economically attractive to the major gas sellers for liquefied natural gas facilities. However, a small well-insulated liquefied natural gas container trade is being investigated, and, if successful, small quantities of liquefied natural gas may be able to be delivered from liquefied natural gas storage, just like the gasoline tankers of today. Even so, liquefied natural gas must be stored for periods of time (months) without significant boil-off losses, which is difficult.

LNG will likely play an increasing role in the development of giant gas fields, as most countries, especially net oil importers, are keen on developing their gas reserves, however stranded, for greater energy independence and extending domestic oil reserves where applicable, as well as for environmental reasons.

1.8.3 Compressed Natural Gas

Gas can be transported in containers at high pressures, typically 1800 psig for a rich gas (significant amounts of ethane, propane, etc.) to roughly 3600 psig for a lean gas (mainly methane). Gas at these pressures is termed *compressed natural gas*. Compressed natural gas is used in some countries for vehicular transport as an alternative to conventional fuels (gasoline or diesel). The filling stations can be supplied by pipeline gas, but the compressors needed to get the gas to 3000 psig can be expensive to purchase, maintain, and operate.

An alternative approach has dedicated transport ships carrying straight long, large-diameter pipes in an insulated cold storage cargo package. The gas has to be dried, compressed, and chilled for storage onboard. By careful control of temperature, more gas should be transported in any ship of a given payload capacity, subject to volume limitation and amount and weight of material of the pipe (pressure and safety considerations). Suitable compressors and chillers are needed, but would be much less expensive than a natural gas liquefier and would be standard so that costs could be further minimized. According to the proponents, the terminal facilities would also be simple and hence less expensive. Two new types of CNG transport are being promoted by their respective companies and are discussed next (Fischer, 2001):

"VOTRANS" is a new type of CNG marine-transport technology from EnerSea Transport, L.L.C. Its engineering studies indicate that it can move up to 2 Bcf per ship over distances up to 4000 miles at significantly lower total costs than LNG. The technology comprises large-diameter pipe structures manifolded together in tiers, essentially a sea-going pipeline. To maintain temperature, the pipe structures are contained within a nitrogen-filled, insulated chamber. It can store CNG more efficiently at significantly lower compression (∼40% compared to LNG), increase vessel capacities, reduce costs, and transport both lean and rich gas. Finally, VOTRANS minimizes gas losses during processing and transport to less than 7% compared to as much as 20% for LNG.

"Coselle" CNG technology is from Cran & Stenning Technology Inc. The system uses conventional, 10.6-mile-long, 6-in diameter, 1/4-in wall thickness line pipe in large coils (coselles). Such a CNG carrier may have 108 coselles with a 330-MMcfg capacity. Stored gas temperature is 50°F at 3000 psi. American Bureau of Shipping and Det Norske Veritas have concluded that a Coselle CNG carrier is "at least as safe as other gas carriers." These ships can be loaded at relatively simple marine facilities, including offshore buoy moorings, through flexible hoses connected to onshore or on-platform compressor stations (Stenning, and Cran, 2000).

"Coselle" and "VOTRANS" are two would-be commercial, high-pressure gas storage and transport technologies for CNG. Technical and economic analyses of these two technologies were done by Economides *et al*. (2005).

Compressed natural gas technology provides an effective way for shorter-distance transport of gas. The technology is aimed at monetizing offshore reserves, which cannot be produced because of the unavailability of a pipeline or because the LNG option is very costly. Technically, CNG is easy to deploy with lower requirements for facilities and infrastructure. Results show that for distances up to 2500 miles, natural gas can be transported as CNG at prices ranging from \$0.93 to \$2.23 per MMBTU compared to LNG, which can cost anywhere from \$1.5 to \$2.5 per MMBTU depending on the actual distance. At distances above 2500 miles the cost of delivering gas as CNG becomes higher than the cost for LNG because of the disparity in the volumes of gas transported with the two technologies (Economides *et al*., 2005).

1.8.4 Gas to Solid

Gas can be transported as a solid, with the solid being gas hydrate (Børrehaug and Gudmundsson, 1996; Gudmundsson, 1996; Gudmundsson and Børrehaug, 1996; Gudmundsson *et al*., 1995, 1997). Natural gas hydrate is the product of mixing natural gas with liquid water to form a stable water crystalline ice-like substance. NGH transport, which is still in the experimental stage, is believed to be a viable alternative to liquefied natural gas or pipelines for the transportation of natural gas from source to demand.

Gas to solids involves three stages: production, transportation, and regasification. Natural gas hydrates are created when certain small molecules, particularly methane, ethane, and propane, stabilize the hydrogen bonds within water to form a three-dimensional, cage-like structure with the gas molecule trapped within the cages. A cage is made up of several water molecules held together by hydrogen bonds. Hydrates are formed from natural gas in the presence of liquid water, provided the pressure is above and the temperature is below the equilibrium line of the phase diagram of the gas and liquid water. The solid has a snow-like appearance.

In the oil/gas industry, natural hydrates are a pipeline nuisance and safety hazard. Considerable care must be taken by the operators to ensure that these hydrates do not form, as they can block pipelines if precautions, such as methanol injection, are not taken. However, vast quantities of gas hydrate have been found in permafrost and at the seabed in depths below 500 m (1500 ft) and, if exploited properly, could become the major energy source in the next 30 years.

For gas transport, natural gas hydrates can be formed deliberately by mixing natural gas and water at 80 to 100 bar and 2 to 10°C. If the slurry is refrigerated to around −15°C, it decomposes very slowly at atmospheric pressure so that the hydrate can be transported by ship to market in simple containers insulated to near-adiabatic conditions. At the market, the slurry is melted back to gas and water by controlled warming for use after appropriate drying in electricity power generation stations or other requirements. The hydrate mixture yields up to 160 m^3 of natural gas per ton of hydrate, depending on the manufacture process. The manufacture of the hydrate could be carried out using mobile equipment for onshore and ship for offshore using a floating production, storage, and off-loading vessel with minimal gas processing (cleaning, etc.) prior to hydrate formation, which is attractive commercially.

Conceptually, hydrate slurry production is simply mixing chilled water and gas. In practice, processed gas is fed to a hydrate production plant, where a series of reactors convert it into hydrate slurry. Each reactor further concentrates the hydrate slurry. It is then stored and eventually offloaded onto a transport vessel. At the receiving terminal, the hydrate is dissociated and the gas can be used as desired. The water can be used at the destination if there is a water shortage or returned as ballast to the hydrate generator; because it is saturated with gas, will not take more gas into solution.

The hydrate mixture can be stored at normal temperatures (0 to −10°C) and pressures (10 to 1 atmosphere) where 1 m^3 of hydrate should contain about 160 m^3 gas per m^3 of water. This concentration of gas is attractive, as it is easier to produce, safer, and less expensive to store compared to the 200-m^3 per 1 m^3 of compressed gas (high pressure ca. 3000 psig) or the 637-m^3 gas per 1 m^3 of liquefied natural gas (low temperatures of −162°C).

Gas storage in hydrate form becomes especially efficient at relatively low pressures where substantially more gas per unit volume is contained in the hydrate than in the free state or in the compressed state when the pressure has dropped. When compared to the transportation of natural gas by pipeline or as liquefied natural gas, the hydrate concept has lower capital and operating costs for the movement of quantities of natural gas over adverse conditions. Thus, gas hydrate is very effective for gas storage and transport as it eliminates low temperatures and the necessity of compressing the gas to high pressures. Dry hydrate pellets yield about 160 m^3 of gas at standard conditions from 1 m^3 of hydrate compared to the 637-m^3 per 1 m^3 of liquefied natural gas. This is a considerable volume penalty (and hence transport cost) if considered in isolation; with less expensive ships for hydrate transport, the process could be economic.

1.8.5 Gas to Power

Currently, much of the transported gas destination is fuel for electricity generation. Electricity generation at or near the reservoir source and transportation by cable to the destination(s) (GTP) is possible. Thus, for instance, offshore or isolated gas could be used to fuel an offshore power plant (may be sited in less hostile waters), which would generate electricity for sale onshore or to other offshore customers. Unfortunately, because installing high-power lines to reach the shoreline appears to be almost as expensive as pipelines, that gas to power could be viewed as defeating the purpose of an alternative less expensive solution for transporting gas. There is significant energy loss from the cables along the long-distance transmission lines, more so if the power is AC rather than DC; additionally, losses also occur when the power is converted to DC from AC and when it is converted from the high voltages used in transmission to the lower values needed by the consumers.

Some consider having the energy as gas at the consumers' end gives greater flexibility and better thermal efficiencies because the waste heat can be used for local heating and desalination. This view is strengthened by economics, as power generation uses approximately 1 million scf/day of gas for every 10 MW of power generated so that even large generation capacity would not consume much of the gas from larger fields and thus not generate large revenues for the gas producers. Nevertheless, gas to power has been an option much considered in the United States for getting energy from the Alaskan gas and oil fields to populated areas.

There are other practical considerations to note such as if the gas is associated gas, then if there is a generator shutdown and no other gas outlet, the whole oil production facility might also have to be shut down or the gas released to flare. Also, if there are operational problems within the generation plant the generators must be able to shut down quickly (in around 60 s) to keep a small incident from escalating. Additionally, the shutdown system itself must be safe so that any plant that has complicated processes that require a purge cycle or a cool-down cycle before it can shut down is clearly unsuitable (Ballard, 1965). Finally, if the plant cannot shut down easily and/or be able to start up again quickly (perhaps in an hour), operators will be hesitant to ever shut down the process for fear of financial retribution from the power distributors.

1.8.6 Gas to Liquids

In GTL transport processes, the natural gas is converted to a liquid, such as syncrude methanol and ammonia, and is transported as such (Knott, 1997; Skrebowski, 1998; Thomas, 1998; Gaffney Cline and Associates, 2001). The technology of converting natural gas to liquids is not new. In the first step, methane is mixed with steam and converted to syngas or synthetic gas (mixtures of carbon monoxide and hydrogen) by one of a number of routes using suitable new catalyst technology (Cranmore and Stanton, 2000a). The syngas is then converted into a liquid using a Fischer-Tropsch process (in the presence of a catalyst) or an oxygenation method (mixing syngas with oxygen in the presence of a suitable catalyst). The produced liquid can be a fuel, usually a clean-burning motor fuel (syncrude) or lubricant, or ammonia or methanol or some precursor for plastics manufacture (e.g., urea, dimethyl ether, which is also used as a transportation fuel, LPG substitute, or power generation fuel, as well as a chemical feedstock). Hundreds of modifications and patents have been applied to this complex, energy-intensive process, and further developments continue to the present day. Most recent modifications generally involve lowering capital expenditures and the overall energy required for processing, especially through the use of proprietary catalysts and the manner in which oxygen is added to the system.

Methanol is a gas-to-liquids option that has been in commission since the mid-1940s. While methanol produced from gas was originally a relatively inefficient conversion process, optimized technology has improved the efficiency. Methanol can be used in internal combustion engines as a fuel, but the current market for methanol as a fuel is limited, although the development of fuel cells for motor vehicles may change this. Methanol is best used as a basic chemical feedstock for the manufacture of plastics.

Other GTL processes are being developed to produce clean fuels, e.g., syncrude, diesel, or many other products, including lubricants and waxes, from gas but require a complex (expensive) chemical plant with novel catalyst technology.

1.8.7 Gas to Commodity

Commodities such as aluminium, glass, bricks, cement, and iron bars all require large quantities of energy in their making. In the gas-to-commodity concept, the gas is converted to thermal or electrical power, which is then used in the production of the commodity, which is then sold on the open market. It is the energy from the gas, heat via electricity or direct combustion, and not the components of the gas-to-liquids concept that is used. The gas energy is, in essence, transported via the commodity, but there are many market risks, which should be fully assessed. The cost of a GTC plant is very high and raw materials for conversion to commodities, e.g., bauxite, silica sand, and limestone, may be difficult to import to sites with reliability. Therefore, much thought has to be given before embarking on the project(s) and monetizing the gas by this route (Thomas and Dawe, 2003).

As discussed earlier, there are a number of options of exporting natural gas energy from oil and gas fields to market. Any gas energy export route requires a huge investment in infrastructure and long-term “fail-proof” contracts, covering perhaps 20 years or more. But which is the best way to monetize the gas? Gas-rich countries are currently in this challenging debate. There could be options for handling niche markets for gas reserves that are stranded (no market) and for associated gas (on- or offshore) that cannot be flared or reinjected or for small reservoirs that cannot otherwise be exploited economically. Transportation of natural gas as a hydrate or CNG is believed feasible at costs less than for LNG and where pipelines are not possible. The competitive advantage of GTS or CNG over the other nonpipeline transport processes is that they are intrinsically simple, making them much easier to implement at lower capital costs, provided economically attractive market opportunities can be negotiated to the gas seller. The transport options preferred by governments and companies must not only take the economic risks into account, but must also consider the negative effects of possible terrorist activity, political changes, and trade embargos over long periods of time. Thomas and Dawe (2003) cover many of the essential technical points and broad economic pointers needed to enter the discussion of gas-rich states that do not need the gas for domestic use, but wish to monetize their reserves by export.

REFERENCES

Ballard, D., How to operate quick-cycle plants, *Hydrocarbon Proc.* **44**(4), 131 (1965).

Børrehaug, A., and Gudmundsson, J.S., Gas transportation in hydrate form, "In Proceeding of EUROGAS 96," pp. 35–41, Trondheim (June 3–5, 1996).

Cranmore, R.G., and Stanton, E., Natural gas, In Modern Petroleum Technology, (R.A. Dawe, ed.), Chapter 9. Wiley, Hoboken, NJ (2000a).

Cranmore, R.G., and Stanton, E., Transport, In Modern Petroleum Technology, (R.A. Dawe, ed.,) Chapter 10. Wiley, Hoboken, NJ (2000b).

Economides, M.J., Sun, K., and Subero, G., "Compressed Natural Gas (CNG): An Alternative to Liquid Natural Gas (LNG)," SPE 92047, Paper presented at the Asia Pacific Oil and Gas Conference and Exhibition, Jakarta, Indonesia (April 5–7, 2005).

EIA, "Natural Gas Monthly" (August 1997), downloaded from Energy Information Administration (EIA) web site: http://www.eia.gov (2002).

Fischer, P.A., How operators will bring worthless gas to market, *Natural Gas*, Part 8, Monetizing Stranded Gas, World Oil **222**, 11 (2001).

Gaffney Cline and Associates, "GTL Discussion," Paper prepared for the Gas to Liquids Taskforce, Australian Department of Industry, Science & Resources, Commonwealth of Australia (June 2001).

Gudmundsson, J.S., "Method for Production of Gas Hydrate for Transportation and Storage," U.S. Patent No. 5, 536, 893 (1996).

Gudmundsson, J.S., Hveding, F., and Børrehaug, A., Transport of natural gas as frozen hydrate, "In Proceeding of the 5th International Offshore and Polar Engineering Conference," Vol. I, pp. 282–288, The Hague, The Netherlands (June 11–16, 1995).

Gudmundsson, J.S., and Børrehaug, A., Frozen hydrate for transport of natural gas, "In Proceeding of the 2nd International Conference on Natural Gas Hydrates," pp. 415–422, Toulouse (June 2–6, 1996).

Gudmundsson, J.S., Andersson, V., and Levik, O.I., "Gas Storage and Transport Using Hydrates," paper presented at the Offshore Mediterranean Conference, Ravenna (March 19–21, 1997).

Knott, D., Gas-to-liquids projects gaining momentum as process list grows, *Oil Gas J*, 16–21 (June 23, 1997).

Manning, F.S., and Thompson, R.E., "Oil Field Processing of Petroleum," Vol. 1. Pennwell Publishing Company, Tulsa, OK (1991).

Rojey, A., Jaffret, C., Cornot-Gandolph, S., Durand, B., Jullin, S., and Valais, M., "Natural Gas Production, Processing, Transport," Editions Technip, Paris, France (1997).

Skrebowski, C., Gas-to-liquids or LNG? *Petroleum Rev.* 38–39 (Jan. 1998).

Speight, J.G., "Gas Processing: Environmental Aspects and Methods," Butterworth Heinemann, Oxford, England (1993).

Speight, J.G., "Handbook of Petroleum Product Analysis," Wiley, Hoboken, NJ (2002).

Standing, M. B., and Katz, D. L., Density of natural gases, *Trans. AIME* **142**, 140–149 (1942).

Stenning, D.G., and Cran, J.A., "Coselle CNG Economics and Opportunities: A New Way to Ship Natural Gas by Sea," Paper presented at the Gastech 2000 Conference, Houston, TX (Nov. 14–17, 2000).

Sutton, R. P., "Compressibility Factor for High Molecular Weight Reservoir Gases," SPE 14265. Paper presented at the SPE Annual Technical Conference and Exhibition, Las Vegas, NV (Sept. 22–25, 1985).

Thomas, M., Water into wine: Gas-to-liquids technology the key to unlocking future reserves, *Euroil* (May 17–21, 1998).

Thomas, S., and Dawe, R.A., Review of ways to transport natural gas energy from countries which do not need the gas for domestic use, *Energy Int. J.* **28**, 1461–1477 (2003).

Whitson, C. H., and Brule, M. R., "Phase Behavior," SPE Monograph No. 20, Richardson, TX (2000).

Wichert, E., and Aziz, K., Calculate Z's for sour gases, *Hydrocarbon Proc.* **51**, 119–122 (May 1972).

CHAPTER 2

NATURAL GAS ENERGY PRICING

2.1 INTRODUCTION

The pricing of natural gas as an energy commodity presents a number of features of historic significance. These historical elements relating to the politics and economics of the development of oil and gas resources are highly significant in their own right. Indeed, natural gas is playing an increasingly important role as an alternative to crude oil and the gasoline and numerous other by-products refined from petroleum. At the same time, a number of nonhistorical components also enter into the pricing of natural gas as an energy commodity. Some of these elements, especially those implicated in environmental pollution, have an intangible connection to gas pricing, although one would be hard-pressed to ferret out its exact contribution to consumers' expense. Appropriate intervention by the working engineer can do wonders for preserving the environment and all of us living in it from serious harm in the short term and the long term. Nothing including the most precious energy commodity, can have a higher value than human life, which forms the starting point of this chapter. Although natural approaches in general, and their usefulness for addressing environmental problems created by current energy development technologies and approaches in particular, are beyond the scope both of this chapter and of conventional concerns about gas pricing, throughout the rest of this chapter some of the obstacles that have been strewn along and throughout the various paths toward the goal of knowledge-based, all-natural, researched solutions to what some have called "our technological disaster" — solutions that would prove innovative, economically attractive, environmentally appealing, and socially responsible — will be pointed out. The principal goal of this chapter is to examine natural gas energy pricing with the aim of answering the following question: considered from an engineering standpoint, how consistent with their actual potential are the patterns, changes, and trends in the supply of and demand for so-called "nonrenewable" energy commodities?

2.2 ENERGY PRICING, SUPPLY, AND DEMAND

In recent years there has been an increasing recognition that the actual supply and demand for energy commodities diverge significantly from the conventional economic notions and expectations concerning the supply and demand of commodities (Islam and Zatzman, 2004, 2005). Much effort has been expended to define and account for the divergence as an exception relating mainly to the strategic character of energy as a commodity in a globalized marketplace. The problem with an analysis that goes only this far, however, is its failure to explain why current economic theory as a body of theory fails to account for, or predict, the energy exception. Without such comprehension at the level of theory, it becomes difficult, if not impossible, to discover or elucidate any paths that could or might overcome the numerous contradictions and downsides inherent in the present marketing and pricing arrangements in place for energy commodities.

What do such considerations have to do with the price of gas? Consider what happens with the hydrogen sulfide (H_2S) stream recovered from the natural gas-processing scenario. The recovered hydrogen sulfide gas stream itself is considered an object of "waste management." It "may be either vented, flared in waste gas flares or modern smokeless flares, incinerated, or utilized for the production of elemental sulfur or sulfuric acid" (EPA, 1995). Any of the first three options require regulation as potentially hazardous emissions. Again what is being proposed is to burden society with a further problem arising from the original technological decision to "sweeten" the gas as "efficiently" as possible rather than innovate delivery systems that would avert creation in the first place of additional waste management expense. What these waste management expenses actually represent for those selling natural gas is a further cost factored into the final delivered price of the product — to pay for meeting the U.S. Environmental Protection Agency (EPA) standard. (The fourth option, which would not only potentially generate new revenue but do so in a manner that undergirds ongoing production of natural gas with an additional productive purpose as a feedstock generator, is neither discussed nor framed in EPA discussions as part of the planning of natural gas processing.)

If the recovered H_2S gas stream is not to be utilized as a feedstock for commercial applications, the gas is usually passed to a tail gas incinerator in which the H_2S is oxidized to SO_2 and is then passed to the atmosphere out a stack (Mullins, 1975). Future gas engineering projects might want to take note of the fact that smokestack wastes are one of the most endemically costly and health-threatening environmental hazards of modern living and accordingly undertake a waste conversion plan capable of returning a revenue stream that allows the production and transmission costs as well as the delivered price of natural gas either to be reduced or its rate of increase contained. These wastes are especially strongly implicated in the return and increasing occurrence of asthma not only in children, but in many sections of the adult population. In the United States and Canada, the smokestack waste from the Ohio Valley region actually created the well-known problem of “acid rain.” Even as many ingenious methods have been developed to neutralize the quite varied effects of such precipitation on everything from groundwater quality to the health of the southern boreal forest cover of eastern North America in general and its maple and other hardwood tree growth in particular, the geographic region involved takes in the territories of eastern Indiana state, southern Michigan state, and the state of Ohio in the catchment zone of the Ohio Valley; southwestern Ontario and southern and eastern Quebec, in central Canada; northern New York, Vermont, and New Hampshire in the northeastern United States, the catchment area of the St. John River system in Maine, Quebec, and New Brunswick; and the Cobequid Hills of northern Nova Scotia. A careful and reasoned calculation of all the costs and benefits derivable from taking the particulates vented as smokestack waste in this major acid rain-vulnerable zone and recapturing them for production of nanomaterials would likely prove most instructive as well as enlightening on the matter of how inflexible present production, delivery, and pricing arrangements for natural gas actually are.

Science and engineering issues of technological development have become consciously disconnected from rational considerations of the overall aims and needs of human social life, a disconnection that leads to, and feeds, an incoherence that becomes increasingly intolerable in our current “information age,” when information as the building block of useful knowledge and understanding has become available in quantities and detail as never before, but the research and thinking needed to make such a flood of information coherent, comprehensible, and useful remain extremely rare quantities indeed. The working engineer must never forget that there are alternatives, but, remarkably, while the EPA web site generated 5443 "hits" in response to a query for article titles from its libraries that deal with "glycol and natural gas processing," a request for article titles discussing "glycol substitutes" returned a grand total of 2 "hits."

In the course of addressing gas-pricing issues in this chapter, some key elements of the massive disinformation attending the entire notion of "renewable" versus "nonrenewable" energy sources will be mentioned. The price mechanism and price formation for these resources as energy commodities do not depend upon mainly or solely upon any of their inherent material characteristics as raw materials, nor on their particular properties as energy sources. Nor do these price mechanisms and price formation depend purely or even mainly upon the conditions limiting or enabling how humanity accesses them. Not the prices themselves, but certainly the processes of energy commodity price formation and associated price movements remain circumscribed to an uncannily large degree by the evolution of energy policy in the context of world-altering developments at the level of international politics and economics as they emerged at the end of World War I. The decisive role of the overall economic system in these price mechanisms and price formation has been influenced greatly by this context, even into our own time. The leading features of this overall economic system have exercised distorting influences over the engineering approaches taken to the development of these resources. Practical as the concerns must be of the working engineer, an essential barrier to humanity achieving a truly functional as well as just solution to our current energy supply–demand conundrum is to be found in this historic legacy.

2.3 SUSTAINABILITY AND THE INCREASING FASCINATION WITH NATURAL GAS

The entire question of the sustainability of economies dependent on fossil fuels as their principal energy source — be it for transportation, for domestic uses such as heating and cooking, or for the provision of electrical power for industrial or domestic markets — is affected and influenced by changes over time in the pricing structure of the production and delivery of these resources. However, there still seems to have no price at which humanity will give up seeking and harnessing these resources. Thus, as influential as shifts in prices and production costs may appear in the short term, the price mechanism as such, left to its own devices, is unlikely to play a determining role in how choices are made between different energy alternatives for the long term. As the long term is exactly what sustainability is all about, however, and as evidence has accumulated about the finitude of existing conventional reserves that are either in production or under exploration, a great hubbub has broken out over whether the world is about to enter the era after which oil production will have “peaked.” Matters of this kind insert an element beyond conventional notions of risk assessment that become highly problematic for predicting price trends and resource “futures.” At the same time, it seems likely that a sustainability factor will have to be considered for longer term contracts of the kind, for example, recently signed between China and Iran, locking in production and distribution of the output of the world’s largest gas field to date for the next 30 years. Gas supplies in North America are no less affected by sustainability considerations: if present plans materialize in the next 10–15 years, U.S. imports of natural gas from Canada will more than double, but not on the basis of increasing gas production in Canada so much as by increasing pipeline throughout of liquefied natural gas (LNG) brought ashore at termini in Canadian ports on the Atlantic, the Pacific, and in the St. Lawrence River connected by pipelines to receiving centers in the United States. Interestingly also in this connection, recently revised estimates by the U.S. Energy Information Administration about future supplies of natural gas for the United States from Canadian fields have noted a strong likelihood that increasing quantities of Alberta gas are likely to be diverted as feedstock for the heavy oil-refining process, which extracts “synthetic crude” from the bitumen mass of the oil sands. An associated aspect of the sustainability problem arising in regard to Canadian-supplied quantities of natural gas for the U.S. market concerns the increasing *de facto* annexation of Canada and its economic development policy by the United States While consolidation of the pipeline supply from Canada to the United States is bound to improve economies of scale that will drive prices downward for U.S. customers or at least arrest any tendency for the price to rise, the situation viewed from the Canadian side of the relationship may be very different. Already, for example, almost none of the gas piped from the Sable Island gas field off the province of Nova Scotia to New England is diverted to, or even available for, local markets in the Canadian provinces through which the distributing conduit, the maritimes and northeast pipeline, passes en route to Everett, Massachusetts. If the Canadian-accessed share of the increasing gas throughout predicted in the coming years is similarly limited, how much longer can Canadians be expected to support this development without demanding some rearrangement — with difficult-to-anticipate consequences for consumers on both sides of the border?

These scenarios suggest that future gas pricing based on sustaining the *status quo* arrangements — the present set of relationships — may compel consumers to pay a hefty and possibly unacceptable premium in the name of sustainability of energy supplies. Scientifically, and at a more fundamental level, the sustainability promised by shifting away from refined petroleum toward natural gas may be little more than a chimera. Up until now, the world has been sold on the progression “wood–oil–natural gas” as the most concise description of human progress in energy supply, mainly because or in relation to the technologies brought into the process of delivering energy supplies and content from these sources. However, looking at this arrangement in reverse, the raw material from which reservoir sources of natural gas originate is the oldest, geologically ancient, decayed plant material, oil’s sources slightly younger, and wood’s source the newest of all. In terms of natural mechanics, the younger the energy source, the cleaner its combustion relative to the capacity of the surrounding environment to dispose of the waste product. Thus, the increasing refinements in processing and transmission technologies may not be purchasing greater sustainability in the long term at all. Solid answers to these questions require further research, but the questions themselves go to the heart of the question of technology and whether it ensures anything in itself simply because it is available to be applied.

2.4 IS NATURAL GAS ALWAYS “NONRENEWABLE?”

Natural gas is often considered, like crude oil, as being of finite availability, as it is “nonrenewable” (e.g., unlike solar or wind power). This designation slides in alongside the “fossil fuel” designation also used commonly to describe natural gas. To think, however, of natural gas as a “fossil fuel” is actually a starting point for considerable confusion. It is true that natural gas is found frequently in connection with crude oil. It is also true that crude oil is deposited as the decayed remains of ancient flora and that these are found in very ancient geological layers of the earth in which fossilized remains also frequently appear.

As a fuel/energy source, however, the valuable active element of "natural gas" is methane. Methane is present throughout the Milky Way galaxy. It comprises a large portion of entire celestial bodies, such as "gas giant" planets like Jupiter. On Earth, it occurs in many places having no connection whatever with fossil layers in the Earth's geology, for example, as a major by-product of landfills. In the earth itself at very great depths both on land and in the oceans, there are significant quantities of trapped gas seeking a way to be vented. Finally, there is no set formula for "crude oil," which appears in raw material form as an inhomogeneous, usually liquid, mass of carbon and hydrogen atoms in a wide variety of proportions, but the chemical formula of methane is always the same, CH_4, and, moreover, this chemical configuration of a single carbon atom with four available bonding sites filled each by a hydrogen atom is remarkably stable. Only when some pressure confining it, or some level of thermal energy exciting the atoms as a gas, is removed does its physical state become unstable.

"Nonrenewable" is taken to mean, in the case of a generally considered fossil fuel reservoir, that once a reservoir is exhausted commercially — often it is only possible to extract 20 to 40% of the reservoir's estimated crude oil and or gas contents — it cannot be replaced in a time frame that would be useful to humanity within the time frames that normally condition or dictate modern economic existence. It was formed over geological time spans of billions of years; even the hugest oil sources located on this planet would last less than 200 years at current rates of economic activity.

Of course, much natural gas comes from these same reservoirs, formed over the same lengthy geological time span and existing under pressure at various depths in the bowels of the earth. However, whereas crude oil really needed that geological time span to form in liquid state, decaying organic material anywhere can begin to produce methane, the active energetic component of natural gas. In addition to being available in reservoir formations as an energy source, natural gas, can also be extracted as an output of a process of waste conversion.

There is a much homelier example of exactly the same idea for a source of natural gas: landfills. It appears that there is no way modern human societies are not going to produce landfills. One can treat them purely as a waste burden on society or research their possibilities and innovate. Russia, already rated as one of the world's largest single reserves of conventional sources of natural gas, has a further reserve in municipal solid waste landfills. Although enormous quantities continue simply to be incinerated, the largest biogas complex in Europe are two plants on the outskirts of Moscow fed from some of this waste (Izrael *et al*., 1997).

A number of observers of the possible further development of natural gas as a substitute for refined petroleum see the downsides that have become undeniable and that emanate from the cartelized development of the petroleum industry. Some partisans of this line of development assert that natural gas can fill the gap so long as it is developed and made available to end users on a competitive, deregulated, and noncartelized basis. This is in fact a starting point of yet further disinformation. The actual evolution of the expanded production and distribution of natural gas, organized according to the deregulated model pioneered in the United States, is in fact premised on enabling a “soft landing” for Big Oil on the basis of restructuring government incentives to shift investment from Big Oil into natural gas. It is not about independent business entities undertaking free competition and accepting the government as an independent referee–arbitrator–regulator of their relations. If more or less the same financial interests that created Big Oil and brought it to the present impasse end up directing or are otherwise in charge of the processes of substituting natural gas for end uses currently fulfilled by refined petroleum, the interests of consumers will not be advanced one whit; in fact, their powerlessness will have multiplied. There is considerable evidence accumulating of interlocking directorates throughout the energy industry and that the “arm’s length” relationship that is supposed to exist among those invested in Big Oil and entities undertaking major new investments in natural gas processing and marketing is notional. Already by 1993, a study conducted for the EPA but not published until 1999 found a rapid growth of increasing concentration of U.S. natural gas production and markets in the hands of a remarkably tiny percentage of all enterprises involved in the exploration, production, processing, distribution, or storage of natural gas (EPA, 1999). Is this some inadequacy of the deregulated system governing natural gas, is it some “externality” that the deregulation process just never took into account, or is it something the entire deregulation scenario was elaborated to conceal?

Once the fullest and most unbiased inventory of all sources of natural gas — be it methane emissions from waste or disused coal beds, — is compiled, the question as to whether it is indeed economically sounder to continue with conventional approaches or to conduct research into an innovative delivery system can be settled. Elucidating the history of actual development is quite important for shedding light on where thinking became trapped in the past down various blind alleys, especially where various assumptions about crude oil as a nonrenewable fossil fuel were simply transferred and reapplied willy nilly to natural gas.

Another seed of yet further confusion is the very label “natural” in “natural gas.” The methane fresh from the cow and the methane trapped billions of years ago beneath the Earth’s surface are doubtless equally “natural,” but are their actual potentials as sources of useful energy truly equal? Does pointing out that they both contain CH_4 shut down all further discussion of this temporal dimension of the very formation of these compounds?

Overwhelmingly, the main natural gas sources discussed by and within the oil and gas industry, and within the engineering fields of work connected with that industry, are those that can be extracted in connection with crude oil, using pipeline rights of way or related distribution systems developed initially by oil companies, etc. In the context of the political and economic implications of how strategic energy commodities in general are disposed of on the world scale, these same elements are also significant as this entire aspect emerges as a prominent feature of the overall picture. However, in discussions or considerations in purely engineering-based technical writing on the subject, it is still ordinarily and, it must be said, systematically excluded. This separation of engineering from politics and economics has enabled considerable confusion to grow into outright disinformation. This has spread in the form of two key messages: (1) that the oil and gas industry is a hostage to “Nature,” which possesses only finite quantities of usable or accessible energy sources to meet the infinitely expanding needs of humanity, and (2) that humanity is the hostage of the oil and gas industry for the foreseeable future as no serious alternative sources of comparable energy content are available or conceivable.

2.5 U.S. NATURAL GAS: PRICING, MARKETS, RISK MANAGEMENT, AND SUPPLY

2.5.1 Some Features of Current Natural Gas Pricing in the United States

The first and most outstanding feature of natural gas pricing is that there has yet to emerge anything like a single world price. However, the rates

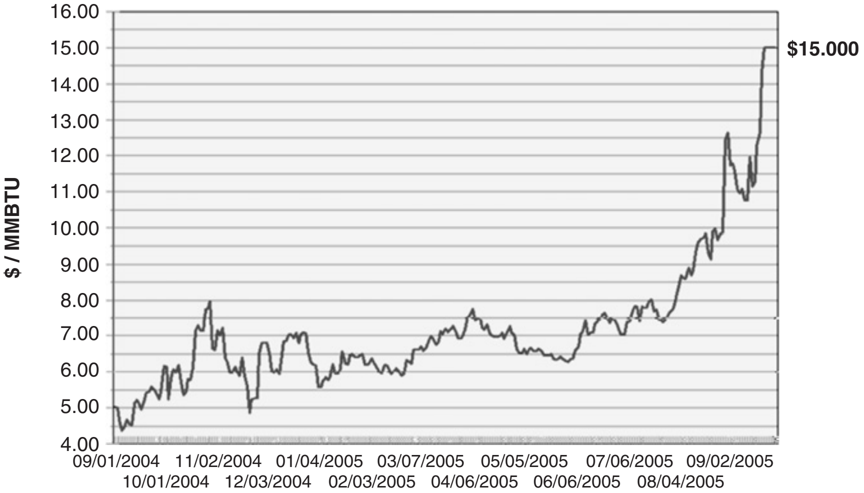


Figure 2-1. Henry Hub natural gas spot market price between September 1, 2004 and September 1, 2005. Source: WTRG Economics.

of increase have tracked upward or downward mostly in conjunction with what was happening with the world oil price.

The temporary tripling of the Henry Hub spot price for natural gas disclosed in Figure 2-1 demonstrates the extreme potential effect of a short-term event — in this case, hurricane Katrina, which, in the end, was found to have caused little consequential damage to the gas pipeline network along the Louisiana coast, whose throughput at Henry, Louisana, forms the basis for the so-called “hub price.” This single event and speculation before, during, and after it about the potentially disastrous consequences of the 2005 hurricane season in the Gulf of Mexico nevertheless enabled sellers into markets basing themselves on the Henry Hub to double the price from the U.S $5.00–7.50/Mcf band in which gas had traded from 1 September 2004 until the third week of August 2005. Such a jump would not be possible if there was a world price for this commodity. In contrast, in the same time period there was an even shorter-lived “speculative premium” that bumped the price of oil from U.S $65 to U.S $70/bbl; it lasted less than 72 h precisely because a world price exists that dampens the impact of such events.

How closely the gas price is hitched to the oil price in the main pricing centers of the United States is clearly disclosed in Figures 2-2 to 2-6,

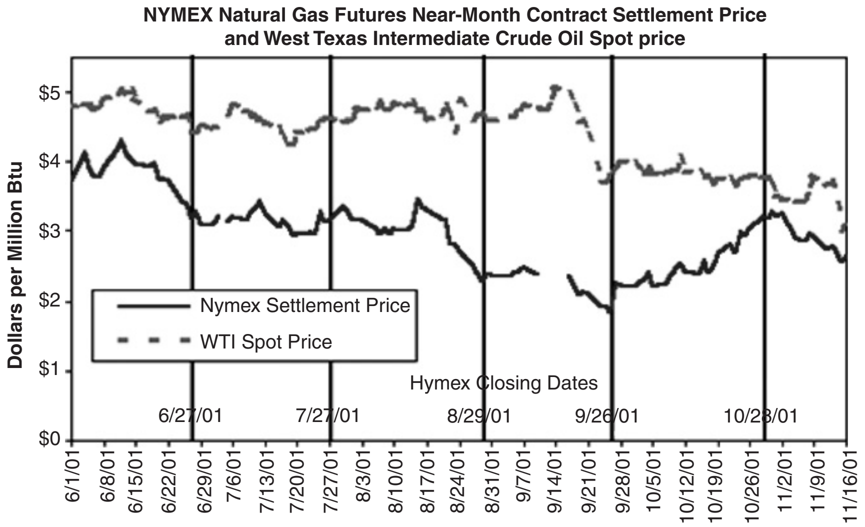


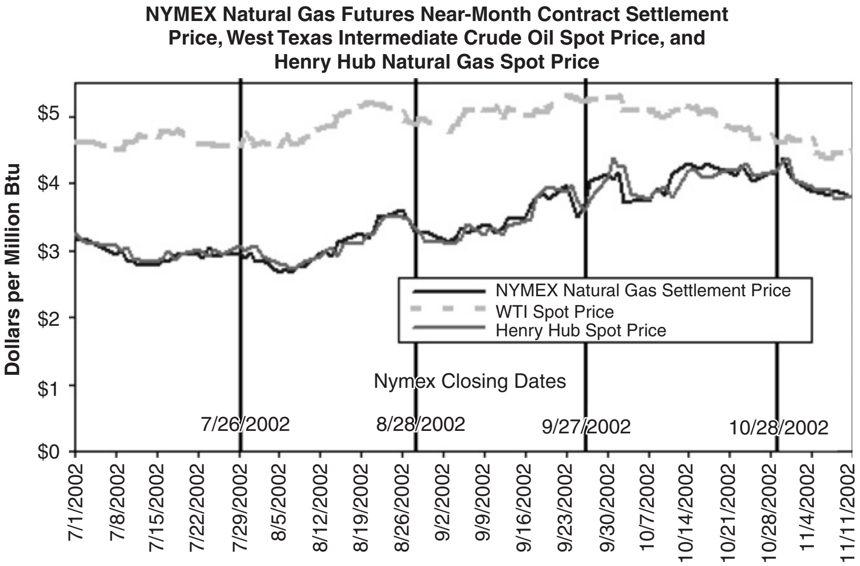
Figure 2-2. Gas and oil prices 1 June 1 November 2001 (based on spot market and futures contracts). The West Texas Intermediate crude oil price, in dollars per barrel, is converted to $MMBtu using a conversion factor of 5.80 MMBtu per barrel. Dates marked by vertical lines are the NYMEX near-month contract settlement dates. The NYMEX was closed on 9/11 9/13. Source: *Reuters Information Service.*

documenting the movement of gas and oil prices in a 2 to 3-month window leading up to the third week of November in 2001, 2002, 2003, 2004, and 2005, respectively.

The close tracking of the oil price by that of gas is very evident. When compared in terms of actual energy content, *i.e.*, dollars per million Btu, the prices are rather close but oil is predictably higher, reflecting the widespread industrial demand for feedstock to refine into end products. The same pattern is seen a year later in November 2002 (Figure 2-3).

By the fall of 2003, the price of gas had risen much closer to that of oil, but the basic pattern of the gas price tracking the oil trend remained intact (Figure 2-4).

Reflecting the effects of supply constrictions and profit taking occasioned by the U.S.-coalition invasion of Iraq, the price of oil rose. The autumn increase in gas prices is a typical example of how weather predictions, especially of a colder than normal or earlier than normal onset of winter, temporarily perturb the overall pattern until the pattern restores itself (Figure 2-5).



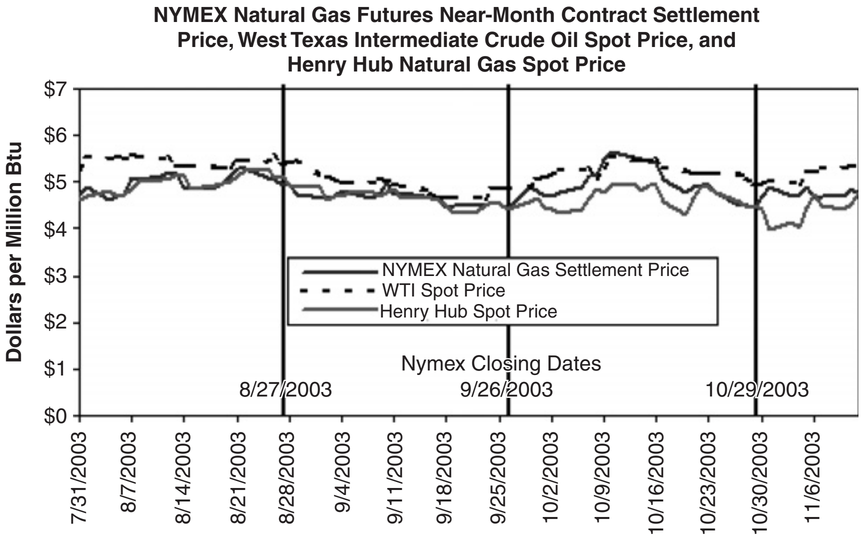
Note: The West Texas Intermediate crude oil price, in dollars per barrel, is converted to \$/MMBtu using a conversion factor of 5.80 MMBtu per barrel. The dates marked by vertical lines are the NYMEX near-month contract settlement dates.

Source: NGI's *Daily Price Index* (http://Intelligencepress.com)

Figure 2-3. Oil and gas prices, 1 July–11 November 2002.

In Figure 2-6, the effects of the ongoing occupation of Iraq by the U.S.-led coalition in destabilizing the supply of oil to the world market from Iraq, and consequently heightening speculation against the future oil price, are seen in the extremely high level maintained by the oil price trend compared to the preceding years (Figures 2-2–2-5). At the end of August 2005, the U.S. Gulf coast, at the center of the main onshore transfer points of oil and gas from the Gulf into the pipeline system that transports large volumes throughout the central and eastern United States, was devastated by hurricane Katrina, an event sufficiently destructive to sink entire portions of the city of New Orleans. Reflecting this development, the gas price topped the oil price for a period, not receding until late October.

Data in these figures suggest that the conventional explanation of an energy market as a price setter and its participant price takers is seriously incomplete, as well as misleading about some crucial underlying influences. The gas trend tracks the oil trend, although the latter reflects a world price that damps down effects of temporary perturbations or incorporates the consequences of longer term ones (like the war in Iraq)


NYMEX Natural Gas Futures Near-Month Contract Settlement Price, West Texas Intermediate Crude Oil Spot Price, and Henry Hub Natural Gas Spot Price

Note: The West Texas Intermediate crude oil price, in dollars per barrel, is converted to $/MMBtu using a conversion factor of 5.80 MMBtu per barrel. The dates marked by vertical lines are the NYMEX near-month contract settlement dates.

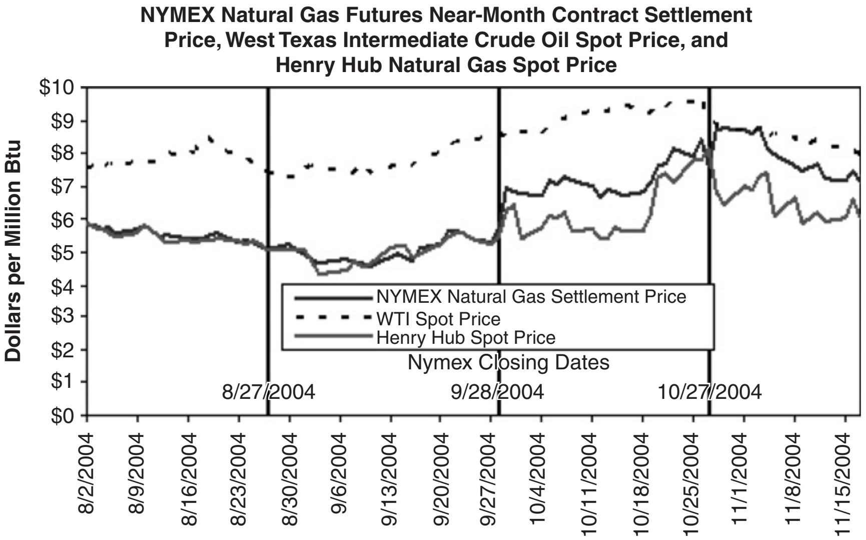
Source: NGI's *Daily Price Index* (http://Intelligencepress.com)

Figure 2-4. Oil and gas prices, 31 July–6 November 2003.

in the general price level. Occasional short-term perturbations in the gas market punctuate the overall picture, but these are actually the moments where the anarchy inherent in the market system unveils its whip hand. Thus, contrary to what was promised and predicted, deregulation of gas and electricity prices has essentially failed to bring about any significant orderliness in gas marketing. Indeed, this conclusion could certainly have been suspected from a brief consideration of the financial earthquake unleashed by the bankruptcy of Enron, a pure "deregulated market" player, and its various consequences and ripple effects throughout corporate boardrooms across North America.

2.5.2 U.S. Energy Markets: The Regulation–Deregulation Nexus

As Table 2-1 illustrates, from the outset, the rising importance of gas within the U.S. energy supply picture could not long be separated from, and was actually decisive in bringing about, challenges to existing regulation. The first 30 years of this process developed from the supply of gas as a customer service being regulated as a residential utility, like electrical



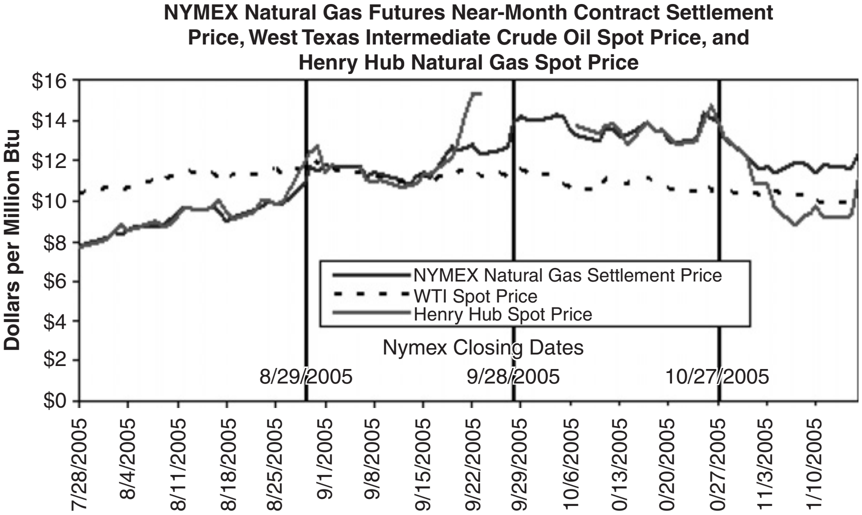
NYMEX Natural Gas Futures Near-Month Contract Settlement Price, West Texas Intermediate Crude Oil Spot Price, and Henry Hub Natural Gas Spot Price

Note: The West Texas Intermediate crude oil price, in dollars per barrel, is converted to $/MMBtu using a conversion factor of 5.80 MMBtu per barrel. The dates marked by vertical lines are the NYMEX near-month contract settlement dates.

Source: NGI's *Daily Price Index* (http://Intelligencepress.com)

Figure 2-5. Oil and gas prices, 2 August–15 November 2004.

service, to its delivery being regulated across state lines on the basis of preventing the emergence and dictate of oligopolistic firms operating as a restraint upon free competition. The exigencies of wartime sequestering of petroleum to meet the demands of modern motorized forms of warfare on land, sea, and in the air opened the prospect of gas being substituted for oil in the supply of electric power to industry and local utilities. By the 1950s, the bias of regulatory effort shifted toward attenuating tendencies among producer oligopolies, pipeline oligopolies, and supplier oligopolies to poach one another's turf. Conventional economics presents natural gas production and distribution as a "natural monopoly," where the costs of business entry and startup, and the likely diminution of any newcomer's margin of profit, discourage the establishment of competing enterprises in the same market. This fails, however, to explain and actually confuses historical facts. Natural gas went from being highly regulated to being deregulated at its industrial structure was transformed from scores of small local distributors to a network dominated by a handful of major pipeline operators.



Note: The West Texas Intermediate crude oil price, in dollars per barrel, is converted to $/MMBtu using a conversion factor of 5.80 MMBtu per barrel. The dates marked by vertical lines are the NYMEX near-month contract settlement dates.

Source: NGI's *Daily Price Index* (http://Intelligencepress.com)

Figure 2-6. Oil and gas prices, 28 July–10 November 2005.

As the point of market saturation and the peak in the increase of the rate of return on investment in natural gas production, processing/transport, and delivery was passed, the pressure grew to consider deregulation. This pressure intensified greatly as it became apparent after 1970 that the United States could no longer profitably increase its domestic production of petroleum. Table 2-2 documents the manner in which, since 1978–1985, deregulation has been introduced.

2.5.3 Energy Price Volatility and Derivatives

It has been clear for some time that when it comes to energy sources and the maintenance of upstream investment and downstream revenue, the market cannot deliver when left to its own devices. One of the main ways developed to overcome this gap has been the development of risk management strategy as a major part of managing corporate-scale interventions in energy markets.

Derivatives, conventionally defined as “financial instruments (contracts) that do not represent ownership rights in any asset but, rather,

Table 2-1
Initial Development and Regulation of Natural Gas in the United States: A Timeline Compiled from Fleay (1998) and Natural Gas Organization (2005)

Development and Regulation of Natural Gas
1859 — Col. Drake, U.S. army, drills first commercial oil well at Spindletop, Pennsylvania. Ambient shows of natural gas are “flared off.” This technique is applied generally in oil fields until the start of World War II.
1900s — Starting in New York and Wisconsin, municipalities enable articulation of gas pipeline networks into urban neighborhoods to provide an alternative to coal for domestic cooking and heating.
1935 — U.S. Federal Trade Commission (FTC) report notes with concern the increasing merger of financial interests from the electric and gas utility industries and recommends intervention to regulate in the interests of consumers.
1938 — U.S. Congress passes Natural Gas Act (NGA) in response to the FTC recommendations and continuing widespread public agitation over cartelization of utilities raising the cost of delivering an essential service on an uncompetitive basis at a time of widespread impoverishment and want. Drawing on the model of the Interstate Commerce Commission Act of 1913, the NGA proposes to arrest their negative consequences of excessive concentration by empowering the Federal Power Commission to regulate interstate development of pipeline networks and sale of gas.
1941/1945 — The United States enters the war against Japan. Within a year, as it becomes clear that domestically produced oil will be needed for the war, especially for the fleet, the U.S. government mandates widespread development of natural gas in the domestic market as a wartime substitute for refined petroleum wherever possible. In the United States, a Supreme Court ruling regulates the wellhead price of gas where the utility and the pipeline are part of the same financial entity.
1954 — Supreme Court renders its decision in *Phillips Petroleum Co. v. Wisconsin* (347 U.S. 672 [1954]). In this decision, the Supreme Court ruled that any natural gas producer selling natural gas into interstate pipelines fell under the classification of “natural gas companies” in the NGA and were subject to regulatory oversight by the FPC, simultaneously streamlining regulation of both the rates at which producers sold natural gas into the interstate market and the rates at which natural gas was sold by interstate pipelines to local distribution utilities. This sets the stage for the issue on which the deregulation melodrama unfolds, as a liberation of the producers from the tributary yoke of the pipeline companies backed by the FPC.
1970 — U.S. domestic oil production reaches a peak from which it has declined without interruption to date. Energy industry planning of domestic supply centers hereafter around natural gas.
1998 — The role of oil in supplying the commercial energy of the entire globe had fallen to about 40%; natural gas had risen to about 22%. Some 60% of the consumption of oil is been taken up by transport systems: rail, sea, and air.

Table 2-2
Timeline of Natural Gas Deregulation in the United States Compiled from LIHEAP Clearinghouse (2005)

A Condensed Deregulation Timeline

1978 — Natural Gas Policy Act ends federal control over the wellhead price of “new” gas as of January 1, 1985, but keeps in place wellhead price controls for older vintages of gas. According to the American Gas Association, the leading natural gas producers’ cartel in the United States, with this administrative change of everything except the wellhead price for existing supplies, “the laws of supply and demand begin to work again in the natural gas industry.”

1985 — Order 436, issued by the Federal Energy Regulatory Commission (FERC), establishes a voluntary program that encourages natural gas pipelines to be “open access” carriers of natural gas bought directly by users from producers. This order brings the separation of pipelines’ merchant and transportation functions and initiates a restructuring of the natural gas industry’s regulatory structure.

1989 — Natural Gas Wellhead Decontrol Act lifts the remaining wellhead price controls on natural gas.

1992 — FERC Order 636 orders interstate natural gas pipelines to “unbundle,” or offer separately, their gas sales, transportation, and storage services. The goal of this order is to ensure that all natural gas suppliers compete for gas purchasers on equal footing.

1996 — The first residential natural gas customer choice programs are implemented. By 1997, local natural gas utilities in 17 states and the District of Columbia had proposed and/or implemented such residential customer choice policies or pilot programs.

1996 — On 31 August and 1 September, legislature in Sacramento, California, the most populous state in the United States, authorizes deregulation of electric power rates to begin 1 January 1998.

1999–2000 — During one of the hottest summers on record in the state of California, a private corporation in Houston, Texas, set up under the name Enron to purchase electric power from deregulated utilities and resell it to customers in other deregulated jurisdictions, corners the California market and people’s air conditioning bills skyrocket. Mustering the facts of the brownouts and related power cuts and outages caused by the unprecedented demand in conditions of such artificially restricted supply as evidence of the need for massive increases in power supplies generally, lobbyists for the power engineering cartel in Washington, DC put the case for expanded federal subsidies and incentives to construct new power plants across the United States. As an unexpectedly long cold snap hits the populous regions of the U.S. eastern seaboard and southeast in late 2000, the FERC permits a fourfold increase in the wellhead price (Henry Hub) of natural gas — the main proposed feedstock for the proposed future plants.

derive (*emphasis added*) their value from the value of some other underlying commodity or other asset," have been around since the late 1800s in the United States as a tradable agricultural commodity, the most famous example being probably the "pork belly" contract (EIA, 2002). Within the envelope of the aforedescribed events, and tending to overstimulate the marketing and pricing of both natural gas and electricity, a derivatives market was developed extensively as an instrument for the hedging of risk among entities investing in areas of the energy market undergoing deregulation, especially electricity and natural gas.

Modernized, streamlined derivatives trading, especially the associated practices of indexing arbitrage and what is known as "program trading" (the preset, computer-regulated sell-off by syndicates of banks, investment dealers, and other large-scale purchasers of shares, usually near the end of a financial quarterly reporting period, of tens and even hundreds of millions of shares of stock in a vast range of entities), were first seen in the stock exchanges of the United States on a large scale during and following the October 1987 stock market crash. However, indictments and convictions shortly thereafter of Michael Milken, Ivan Boesky, and other exchange insiders for abusing the latitude available for those prepared to indulge new areas of risk in the market, far from discouraging their development, generated new interest in taming and stabilizing the playing field to facilitate new entrants. The risks involved are multiplied greatly by (1) the mutual and often reciprocal dependence of different contract performance promises on one another and (2) the legal system's general posture that contracts between legal entities are deemed to have been entered into freely until otherwise proven in a court of law, and therefore it is no outside body's business to inquire into or approve in advance any of the terms or requirements of such contracts.

The single most dramatic recent example of what can happen when "contracts" are simply invented without any intention of performing or completing them was seen in the incredibly rapid expansion and equally dramatic implosion of Enron, the derivatives-based energy commodities trader in Houston, Texas, between 1997 and 2001. After the implosion of Enron the EIA commissioned an authoritative study that exculpated derivatives trading as such of any responsibility. Entitled "Derivatives and Risk Management in Energy Industries," the report documented the litany of disasters that have trailed in the wake of liberating deregulated energy commodity markets into the eager and waiting arms of the derivatives traders:

> Although derivatives meet legitimate needs, they have also been implicated in tremendous losses. For example, Orange County, California, lost $1.7 billion in 1993; Metallgesellschaft lost about $1.3 billion in 1993 in energy trading; and in 1998 the Federal Reserve Bank of New York organized a rescue of Long Term Capital Management in order to avoid disrupting international capital markets. And in 2001 Enron became at that time the largest bankruptcy in American history. Enron was a large user and promoter of derivative contracts. Although Enron's failure was not caused by derivatives, its demise raised significant concerns about counterparty (credit) risk and financial reporting in many energy companies.

The report explains the overall situation of energy commodity risk management and derivatives from the government's point of view as follows in the report's summary:

> Natural gas and oil pipelines and residential natural gas prices are still regulated. Operating under government protection, these industries had little need for risk management before the wave of deregulation that began in the 1980s — about the same time that modern risk management tools came into use.

There are five general types of risk that are faced by all businesses: market risk (unexpected changes in interest rates, exchange rates, stock prices, or commodity prices), credit/default risk; operational risk (equipment failure, fraud); liquidity risk (inability to buy or sell commodities at quoted prices); and political risk (new regulations, expropriation). Businesses operating in the petroleum, natural gas, and electricity industries are particularly susceptible to market risk — or more specifically, price risk — as a consequence of the extreme volatility of energy commodity prices. Electricity prices, in particular, are substantially more volatile than other commodity prices. (EIA, 2002)

Of course, with the increased use of natural gas as a fuel for new electric power plants, the risk–reward "game matrix" that might be developed to predict investment outcomes for the widened variety of possible scenarios became ever more convoluted, with changes in electricity prices capable of leading or following changes in natural gas prices, and vice versa.

Price volatility is caused by shifts in the supply and demand for a commodity. Natural gas and wholesale electricity prices are particularly volatile, for several reasons. Demand shifts quickly in response to weather conditions, and "surge production" is limited and expensive. In addition, electricity and natural gas often cannot be moved to areas where there are unexpected increases in demand, and cheap local storage is limited, especially for electricity. (EIA, 2002)

Table 2-3, comparing spot-market price volatility for selected commodities, demonstrates the potential profitability of short-term investing in certain electricity markets over investing in practically any significantly traded other "commodity play" in U.S. markets.

The conventions of risk management strategy are thus very much in flux in U.S. energy markets, a situation that has become increasingly fluid as interest in the field of econometrics about the relationships elucidated by game theory regarding market uncertainty has developed extensively. John F. Nash was awarded the 1994 Nobel prize in economics for his contributions in this field, and the 2006 Nobel prize for economics was awarded to Thomas Schelling and Robert J. Aumann (Aumann, 1964, 1974; Aumann and Branderburger, 1987; Aumann and Hart, 1992, 1994, 2002) for their work in this field as well.

2.5.4 Natural Gas Supply in North America

The role of the United States in natural gas production, marketing, and pricing differs in significant respects from the picture just described. It has developed a cartelizing model that has successfully and, in many respects, falsely presented itself as a model and high point of deregulated competition. Figure 2-7 gives an idea of how the United States supplies its market and hints at how this develops in a manner that also integrates entire portions of the economies of other countries, notably its geographically contiguous neighbors Canada and Mexico, into the economy of the United States.

2.5.4.1 The Special Position of the United States

Annual market consumption of natural gas in the United States has increased from about 5 trillion cubic feet in 1949 to a peak of more than 22 trillion cubic feet in 1973. After falling back throughout most of the 1980s and 1990s, it resumed a steady upward climb again to its next peak, at about 23.3 trillion cubic feet in 2000 (U.S. Department of Energy, 2005).

Two factors account for most of the postwar expansion: the explosion of the housing market throughout the United States, especially during the

Table 2-3
Spot-Market Price Volatility for Selected Commodities (EIA, 2002)[a]

Commodity	Average annual volatility (%)	Market	Period
Electricity			
California–Oregon border	309.9	Spot–peak	1996–2001
Cinergy	435.7	Spot–peak	1996–2001
Palo Verde	304.5	Spot–peak	1996–2001
PJM	389.1	Spot–peak	1996–2001
Natural gas and petroleum			
Light sweet crude oil, LLS	38.3	Spot	1989–2001
Motor gasoline, NYH	39.1	Spot	1989–2001
Heating oil, NYH	38.5	Spot	1989–2001
Natural gas	78.0	Spot	1992–2001
Financial			
Federal funds rate	85.7	Spot	1989–2001
Stock index, S&P 500	15.1	Spot	1989–2001
Treasury bonds, 30 Year	12.6	Spot	1989–2001
Metals			
Copper, LME grade A	32.3	Spot	January 1989–August 2001
Gold bar, Handy & Harman, NY	12.0	Spot	1989–2001
Silver bar, Handy & Harman, NY	20.2	Spot	January 1989–August 2001
Platium, producers	22.6	Spot	January 1989–August 2001

Continued

Table 2-3
Spot-Market Price Volatility for Selected Commodities (EIA, 2002)[a]—cont'd

Commodity	Average annual volatility (%)	Market	Period
Agriculture			
Coffee, BH OM Arabic	37.3	Spot	January 1989–August 2001
Sugar, World Spot	99.0	Spot	January 1989–August 2001
Corn, N Illinois River	37.7	Spot	1994–2001
Soybeans, N. Illinois River	23.8	Spot	1994–2001
Cotton, East TX and OK	76.2	Spot	January 1989–August 2001
FCOJ, Florida Citrus Mutual	20.3	Spot	September 1998–December 2001
Meat			
Cattle, Amarillo	13.3	Spot	January 1989–August 2001
Pork bellies	71.8	Spot	January 1989–August 1999

[a] Data from Commodity Futures Trading Commission Calculations by Energy Information Administration staff.

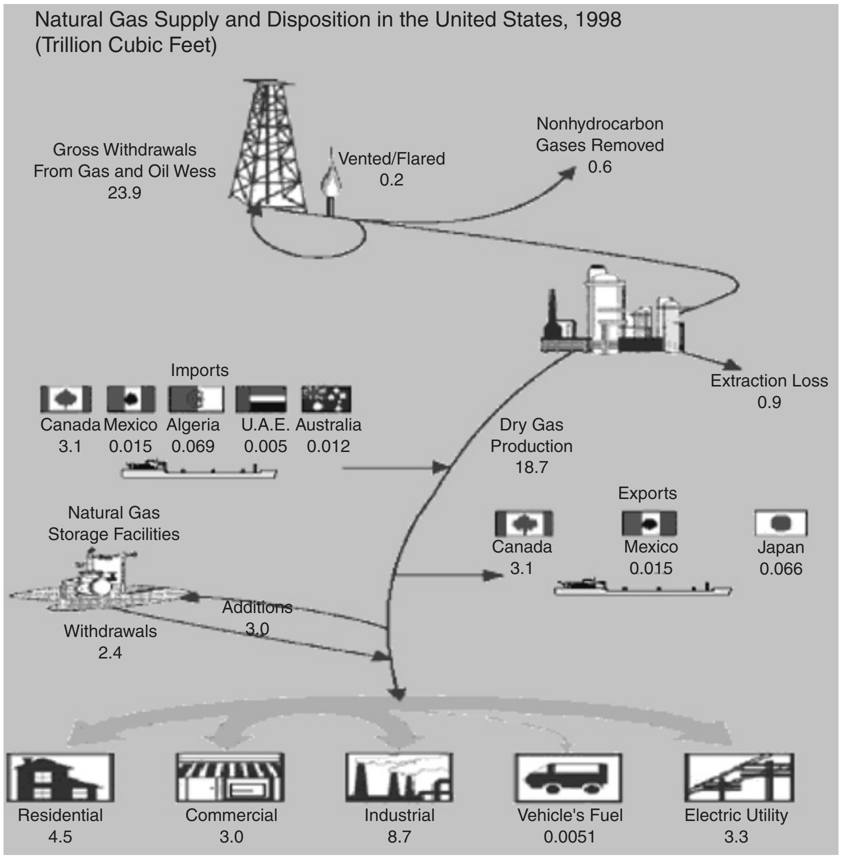


Figure 2-7. United States production, consumption, import, and export of natural gas (1998). Source: Energy information Administration.

main years (1945–1960) of the so-called "baby boom," and the expansion of medium and heavy industry beyond the coal–and steel belts of the U.S. northeast and Great Lakes to new urban growth centers of the west coast, the "right-to-work" states in the south, and the tremendous expansion of population and industry in Texas and Florida.

By the mid-1990s, however, some 70% of new homes in the United States were being heated with natural gas. Essentially this market had reached a saturation point. At the same time, as a result of the so-called "deregulation" of natural gas pricing begun in the United States in 1978 and taking effect after 1985, a situation was created whereby the natural gas producers, after having divested themselves of any significant further financial involvement of pipeline and or other distribution networks and infrastructure, were deemed to be no longer in a cartelizing situation (Udall and Andrews, 2001).

At this point, arguing that they would now need to be able to compensate their operations for the future losses in future revenue that could be anticipated from the decline in household demand, the gas producers became free to combine in setting the wellhead price. This — the so-called "Henry Hub" price, based on delivery at the main hub on the U.S. Gulf Coast, where comes some 60% of U.S.-produced natural gas — quadrupled in 2000. Far from ensuring themselves against future loses of revenue from declining home heating demand, the producers were in fact harvesting the windfall created by deregulation of electricity prices, provision of services and markets in major populated states, starting with California. To keep electric power generation profitable in a deregulated market, natural gas was increasingly substituted for oil and coal. Natural gas would thereafter be produced increasingly as a feedstock commodity for the electricity-generating market, with home heating markets concentrated in the northeast — where population was in long-term decline — contributing a diminishing portion of overall revenue (Udall and Andrews, 2001).

Deregulation of electrical power supply and markets opened massive opportunities for outright swindling of the public on an unprecedented scale. The most dramatic result, which came quickly in 2000–2001, was the rise and collapse of Enron — headquartered in Houston at the center of the natural gas boom, but doing most of its business in California in the form of Ponzi-type "contracts for future supplies of electric power." With its collapse, natural gas prices fell back 50% from their historic highs, but superprofits were garnered in any event from the happy confluence during the winter of 2000–2001 of an extremely cold and extended heating season in the U.S. northeast along with the Enron-induced brownouts and blackouts visited on residents of California (Udall and Andrews, 2001).

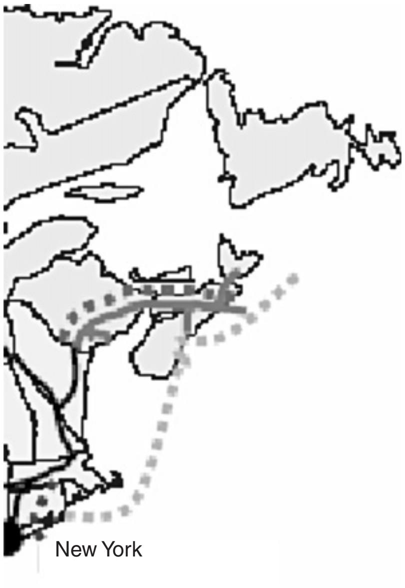
The United States' "long-standing glut of natural gas and electrical capacity, along with the world's spare oil capacity, vanished simultaneously in spring 2000. . . . Fueled by cheap energy, the U.S. economy grew 60 percent since 1986, an astounding five percent in 2000 alone. Gas consumption grew 36 percent over that period (+43.8% over the period 1986 to 2000). But it was the demand for electricity . . . up 5.4 percent in 1998, an astounding rate for such a large economy . . . that has had the biggest impact on gas prices. To meet . . . growing electricity needs, utilities have ordered 180,000 Megawatts of gas-fired power plants to be installed by 2005. It was a logical thing to do: gas is the cheapest, cleanest way to convert fossil fuel to electricity. But if ordering one gas turbine makes perfect sense, ordering 1,000 is a recipe for disaster. No one in the utility industry asked the key question: can we produce enough gas to run all those plants? Many experts think the answer is no. . . .” Reproducing a previously published 20-year projection of natural gas consumption in the United States for the period 1997–2017 estimating relatively moderate increases in annual industrial consumption (from about 8.7 to slightly more than 9 trillion cubic feet), residential use (from 5 to 5.5 trillion cubic feet), and commercial applications (from 3.2 to 3.5 trillion cubic feet), this same source indicates that gas for electricity generation is predicted to nearly triple over the same period, from about 3.2 to more than 9.1 trillion cubic feet per annum (Udall and Andrews, 2001).

The importance of a distribution network cannot be overstated when it comes to natural gas.

> Without a pipeline, natural gas is worthless, a constraint first recognized by the Chinese. They were drilling for gas in 1000 A.D. . . . but their pipeline materials were limited to bamboo. [*sic*] American pipelines today could stretch to the Moon. Most date to post-World II, when Gulf Coast supplies were tied to markets in the Midwest and New England. Since pipelines are prone to corrosion, beer keg-sized diagnostic tools called “pigs” are pushed through the lines to search for weak spots, not always in time. In August, 2000, a pipeline exploded in New Mexico, killing 10 people, and crimping gas deliveries to California. Many aging pipelines need to be rebuilt, replaced, or expanded to deliver more gas to urban areas, where the new fleet of gas-fired power plants will be moored. In December 2000, gas delivered to L.A. briefly fetched $69, equivalent to $400 for a barrel of oil. (Udall and Andrews, 2001)

2.5.4.2 Canada as North American Gas Supplier

All the aforementioned considerations regarding pipeline connections hold particularly profound implications for the development mainly by U.S. companies of natural gas production in The Gully zone of the Sable Island field along the Atlantic coast of the Canadian province of Nova Scotia. Although the gas find was developed entirely with U.S. markets in mind, its estimated life is only about 20 years and the main longer term asset developed with it seems to be the Maritimes and Northeast pipeline



Map 1. How the U.S. natural gas pipeline industry may annex the 200-mile limit off Canada's east coast is disclosed in this detail published by the Canadian Federal Department of Energy and Natural Resources (NRCan Report, 2004).

running from Goldboro, Guysborough County, Nova Scatia, where Sable Island gas is brought ashore. This gas pipeline (the solid line in Map 1) runs across central and northern Nova Scotia through the neighboring province of New Brunswick, across the international border with the state of Maine as far southwest as the Boston-area suburb of Everett, Massachusetts, the nearest U.S. gas pipeline distribution point.

This suggests several hitherto unexamined reasons for the intense rise of U.S. interest in LNG either at portside storage facilities in existing large U.S. urban centers, something that local referenda have defeated repeatedly in a series of downeast ports north and east of the Boston–Portland corridor, or via pipelines that have been built and paid for by others, *e.g.*, the Maritimes and Northeast pipeline.

The most interesting feature in Map 1 is the dotted line. It is the El Paso group's proposed pipeline from the Sable Island gas field to Connecticut. The stop-off point near Shelburne Nova Scatia is not to serve Canadian gas customers. It creates something novel: a means for a private company to sue the Canadian government under Chapter 11 of the North American

Free Trade Agreement should anything happen to the flow of gas to U.S. customers. Neither Shelburne nor the surrounding region in Nova Scotia is a growth center for industry of any kind, now or in the future, and thus there cannot be any other possible interpretation of such a design feature. It would have been a far more direct route, and less expensive, to lay pipeline on the sea bottom directly from the gas field to the New England coast. The only rational explanation not to do so is that the alternative pictured in Map 1 depicts what amounts to a scheme to annex a large portion of the Canadian 200-mile offshore limit directly to the contiguous US 200-mile limit, setting the stage for massive integrated future offshore oil and gas development. As a result of the implicit underlying annexation and integration of the territorial base, a strong fillip would also be given to the forces working hard to overturn the current moratorium on exploratory oil and gas drilling on the George's Bank, a rich fishing zone presently divided by the Canada–U.S. boundaries offshore, a moratorium that Big Oil has been unable to budge for the last decade.

What about objections and alternative proposals from Canadians? A survey of the Canadian energy industry vetted and published in the winter of 2005 by the U.S. Energy Information Administration (EIA, 2005) noted that "while not without controversy, the Canadian LNG terminals have not met with the same level of resistance from local residents and environmentalists that similar facilities in the U.S. have faced" (U.S. Department of Energy, 2005). The reality of the "resistance from local residents and environmentalists that similar facilities in the U.S. have faced" was that local referenda in a series of downeast ports north and east of the Boston–Portland corridor repeatedly defeated bond-issue proposals for such projects. The role of "environmentalists" in these fights was to expose that the smaller and more down at heel the particular port community, the riskier such development could prove in the event of a major "liability event," such as an explosion, spill, or collision. Notwithstanding the possible stubbornness of this or that local crank or crusading environmentalist, it was only after the liability risks were clarified that the community could exercise its franchise. Once this took place, the process proceeded normally. Contrary to the implication planted within the syntax of the EIA description just quoted, the rejection was not because of this local crank or that environmental crusader. Notwithstanding the generally acknowledged objectivity of the U.S. Department of Energy in maintaining the EIA web site as a source of current and reliable information in this field, to suggest that the way matters turned was an aberration caused by individuals hijacking the democratic process is misleading and — had the individuals been named — probably grounds for a libel suit.

One outstandingly principal reason that "Canadian LNG terminals have not met with the same level of resistance" is that local referenda on such matters are available neither in Canadian community traditions nor laws. This fact is unmentioned anywhere in the EIA's online review or other such reviews. While such plebiscites have formed a vital part of the lifeblood of the town-meeting-based "democracy from below" since its invention by New England settlers in the late 1600s, there is in the Canadian system no guaranteed process in the hands of the people to check up on matters arranged between corporate and government bureaucrats. This suggests another possible reason the oil and gas lobby favors going through Canada is that the best way to handle objections is any legal way that would ensure such objections or resistance never attain any legal, public, or widespread airing in the first place.

In connection with the Maritimes and Northeast pipeline, two proposals won approval in principle (i.e., approval pending a formal environmental impact assessment) earlier in 2005 from Canada's National Energy Board:

- A facility that would connect to this pipeline to be owned and operated by Irving Oil Ltd. This company is Canada's largest privately held, vertically integrated oil refiner and marketer and operator of the country's largest refinery (250,000 BPD). It is based at Saint John, NB, where Irving already maintains its own ocean tanker loading and unloading facility known as "Canaport." The estimated throughput is 500 million cubic feet per day (U.S. Department of Energy, 2005).
- A facility capable of delivering twice as much as the Irving project — one billion cubic feet per day — is proposed for at Bear Point, NS outside Point Tupper/Port Hawkesbury, Richmond County NS — near the pipeline's Goldboro terminus — to be owned and operated by Anadarko Petroleum (2005). This is a U.S. company with extensive involvement up to now in the Alberta oil patch in the western canada sedimentary basin, on the other side of the continent from Nova Scotia.

Both proposals raise the curtain on yet another aspect of the development of gas pipeline network distribution between Canada and the United States. This is a link that has become critical for the overall U.S. natural gas supply picture. In total, there are plans to build over 4 billion cubic feet per day of LNG receiving capacity by 2008 in eastern Canada alone. (At this writing in September 2005, there are still several further LNG-terminal proposals on the Pacific and Atlantic coasts waiting to go before the National Energy Board.) Even before the latest Maritimes-based projects mentioned previously, a $1.3 billion, 500-million-cubic-feet-per-day (MMCFD) LNG terminal at Gros Cacouna, Quebec, along the St. Lawrence River to be built by Petro-Canada and TransCanada Pipelines, was approved, apparently in connection with the signing of a deal between Petro-Canada and Russia's Gazprom to feed the Gros Cacouna terminal from Gazprom's Shtokman field (U.S. Department of Energy, 2005). This and other proposals are documented in Table 2-4.

Canada currently supplies about 16% of natural gas consumed in the United States (U.S. Department of Energy, 2005). During the period 2000 through 2003, this represented 57–60% of all the natural gas produced in Canada (NRCan Report, 2004). Clearly Canada is not producing gas for itself and exporting what it cannot consume. Rather it is producing gas for export while domestic consumption develops only incidentally.

Much of Canada's pipeline infrastructure linked to U.S. markets before the 1990s served residential and commercial markets in the larger centers of Alberta as well as southern Ontario and Quebec. However, the Maritimes and Northeast pipeline, one of the newest to the United States and more than 400 km in length, financed in part by generous tax subsidies from the federal and provincial governments involved, was designed without any plan to develop any of the Sable Offshore Energy Project's gas for a potential residential market of about 1.5 million people in central and northern Nova Scotia and southern New Brunswick. To date, there has emerged only a very vague long-term notion of supplying a tiny industrial market in Dartmouth NS and Saint John NB. This points to a new and rather disturbing feature of the increasing embroilment of Canadian resources and territory for this latest phase of U.S. industrial expansion. Popular Canadian standup comedian Dave Broadfoot joked a few years ago that Canadians' role may actually be "evolving [from] being mere hewers of wood and drawers of water' to 'expellers of gas' as well." Table 2-5 documents the pattern of evidence for the serious side of this claim.

NRCan ("Natural Resources Canada"), the Canadian Federal Department of Energy and Natural Resources, which selectively makes public various developments in Canadian energy production and marketing, has become increasingly sophisticated at producing material for public

Table 2-4
Proposed LNG Terminals in Canada (NRCan Report, 2004)

Operator (name)	Location	Send-out capacity (Bcf/d)	Earliest start date	Status
		Projects under review		
Anadarko Petroleum Corporation (Bear Head)	Canso Strait, NS	1.00	2007	Received federal–provincial environmental assessment approval in August 2004
Irving Oil Limited (Canaport)	Saint John, NB	1.00	2007	Received federal–provincial environmental assessment approval in August 2004
Enbridge/Gaz Mètro/Gaz de France (Rabaska)	Beaumont, QC	0.50	2008	Undergoing federal–provincial environmental assessment Process commenced June 2004
Kelitic Petrochemicals	Goldboro, NS	0.50	2008	Undergoing federal–provincial environmental assessment Process commenced August 2004
Galveston LNG	Kitimat, BC	0.61	2008	Undergoing federal–provincial environmental assessment Process commenced August 2004
TransCanada/Petro-Canada (Cacouna Energy Project)	Gros Cacouna, QC	0.50	2009	Undergoing federal–provincial environmental assessment Process commenced September 2004
		Other announced projects		
Westpac Terminals	Prince Rupert, BC	0.3	2009	Conceptual. Project not yet under review
Statia Terminals	Canso Strait, NS	0.50	2009	Conceptual. Project not yet under review
Total Canada		4.91		

Sources: Industry press and company web sites

Table 2-5
Comparative Data on Canadian Natural Gas Production for Export and Domestic Markets (NRCan Report, 2004)

	2003 (Bcf)	2002 (Bcf)	Change (Bcf)	Change (%)
U.S. residential	5,085	4,890	195	4%
U.S. commercial	3,127	3,103	24	1%
U.S. industrial	6,966	7,557	−591	−8%
U.S. electric power	4,929	5,672	−742	−13%
U.S. other[1]	1,769	1,796	−27	−2%
Total U.S. demand	**21,877**	**23,018**	**−1,141**	**−5%**
U.S. LNG exports	64	63	1	2%
U.S. exports to Mexico	333	263	70	27%
Total U.S. gas disposition	22,274	23,344	−1,070	−5%
Canada residential	675	620	55	9%
Canada commercial	518	486	32	7%
Canada industrial	1,029	970	59	6%
Canada electric power	282	261	21	8%
Canada other[2]	410	399	11	3%
Total Canadian demand	**2,914**	**2,736**	**178**	**7%**
Total N.A. Demand	**24,791**	**25,754**	**−963**	**−4%**
Total N.A. Disposition	**25,188**	**26,080**	**−892**	**−3%**

Sources: EIA, StatsCan.
Notes: [a] Includes pipeline and distribution use, lease, and plant fuel and vehicle fuel.
[b] Includes pipeline compressor fuel, processing fuel, and line losses.

consumption that goes far beyond just painting the expanded interest of government and private industry in expanding pipeline infrastructure for gas deliveries to U.S. markets in the best possible light. Their materials deliberately suggest that such developments, far from being done at the Canadian public's cost, are saving the world from "global warming due to excess greenhouse gas emissions." However, is it really without cost to the Canadian host? Not only does the gas carried by some of these pipelines go straight to markets in the United States almost without stopping, e.g., the Maritimes and Northeast line takes more than 97% of Sable-Panuke gas to U.S. markets, with only an inconsequential amount is being distributed to selected industries in Halifax NS, Moncton and Saint John NB — but the Canadian federal and provincial governments involved are collecting only a token royalty to reserve the pipeline's rights of way and fund potential future liabilities, such as environmental damage or property destruction suffered by Canadian residents incidental to maintaining the pipeline. The impression is being created instead that a fate such as "global warming" is important enough to justify enabling private interests, most of them based outside Canada, to obtain access not only to Canadian energy policy-setting structures, but to the treasury as well. Although questions are raised from time to time in the Canadian media about governments' inability to fund necessary social programs adequately, little effort is made to pursue too closely the connection this bears to tax break incentives for new exploration projects and other "giveaways to the oil companies" (McQuaig, 2004).

Meanwhile, however, it may be inferred from the tenor of the reviews published by the most authoritative governmental sources that while future demand in U.S. markets has been pumped up by the stampede to substitute natural gas as the main fuel source for new electric power plants, Canada as a source of natural gas will become increasingly problematic to the extent that synthetic crude oil exports from Canada become a larger part of projected U.S. "energy security" and security of supply. This follows from the fact that the key to expanding production to meet such enhanced demand for synthetic crude will involve diverting more and more of Alberta's natural gas reserves to processing tar sand bitumen, the raw material source of synthetic crude. "Canada's proven natural gas reserves, 56.1 [trillion cubic feet] as of January 2005, only rank 19th in the world. These reserves have decreased by 13.3 percent since 1996, and at current rates, production will completely deplete reserves in 8.6 years." What could cause such a drastic depletion? "The oil sands industry is heavily reliant upon water and natural gas, which is necessary in both the extraction of bitumen from oil sands and the upgrading of bitumen to synthetic oil" (U.S. Department of Energy, 2005). The vulnerability of existing marketing mechanisms of the entire "downstream" sector of the North American oil and gas sector to increases in natural gas prices or sharp reductions in natural gas supply would be distributed, extremely unevenly, with the oil sands industry likely to experience the most critical repercussions.

One interpolation that is consistent with these facts is that, short of new discoveries of natural gas in Canada, the main use of Canadian territory by the U.S. gas market in coming years will be for transiting LNG from its ports and pipelines to U.S. markets. Adding further credibility to that scenario is a published projection of LNG's share of the U.S. gas market rising five times over the next 15 years to 2020 (NRCan Report, 2004). The profitability of such arrangements is clearly premised on utilizing new/future supplies gas increasingly and mainly for electrical power generation and related industrial use, and less and less for residential or other commercial uses (*e.g.*, as an alternative to gasoline).

It would seem clear, then, that one major premise built into meeting expanded the U.S. market for natural gas by supplying LNG is an increasing integration and annexation of Canadian territory, resources, and resource-delivery infrastructure by and for U.S. interests. On the one hand, Anadarko, mentioned earlier, exemplifies strategic foreign investing of this type, possessing interests of its own in Alberta oil and gas, and accumulated knowledge of its actual development trends, since the 1960s (Anadarko Petroleum, 2005). For their part, on the other hand, the Irving interests, although both Canadian owned and somewhat insulated from shareholder takeover attempts (hostile or otherwise) as a result of the status of the principal holding company as a privately held corporation, are already extensively connected to the United States from at least two known directions. Irving Oil has long been integrated into supplying U.S. gasoline and home heating fuel markets in northern New England. The Irving refinery, opened in 1965, was financed to a significant degree by Exxon on the basis of receiving crude for processing from Nigeria, Venezuela, and some other major oil-producing zones outside the Middle East in ocean-going tankers owned or chartered by Exxon Corporation (Bryce, 1978).

Map 2 illustrates the considerable degree to which Canadian natural gas marketing arrangements have already become integrated into the United States.

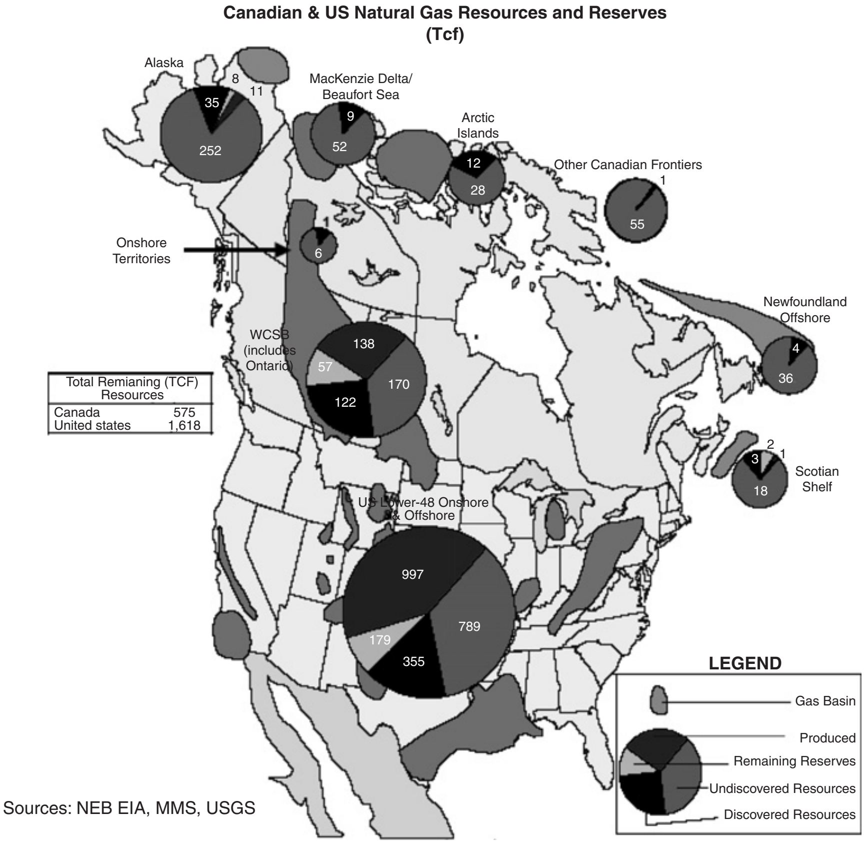
This is a remarkable annexationist feature built into the evolution of U.S. energy policy. It is something not seen at all in the long-term natural gas delivery contracts negotiated earlier in 2005 between the Islamic Republic of Iran and the People's Republic of China or between the gigantic Russian gas syndicate led by Gazprom and its various governmental and corporate customers in the European Union or Japan. Far from identifying this as any kind of problem, let alone investigating its import, considerable confusion has been created. An immediate cause at issue is the prospect that U.S. law, which currently explicitly protects the Alaska wilderness refuge, will be gutted by regulations being planned by the Bush administration based on framing an exception for the energy supply needs of "homeland security." The main threat posed by expanded U.S. petroleum sector investments in exploration and development of new oil and gas fields has been framed in terms of the danger such development could pose for the ecology and natural wilderness spaces of the Arctic zone

Map 2. Important price points of the Canadian side of this integrated market are not cities, but only the toll gates where the gas crosses the border into the United States. (NRCan Report, 2004).

wildlife refuge areas along the Beaufort Sea coastline of eastern Alaska and near the delta of the Mackenzie River in the Canadian Yukon Territory.

Map 3 clearly indicates how focused upon its current state of relative underdevelopment compared to the United States the interest in Canadian gas actually is. The proportion of gas resources in Canada that are undiscovered or not yet in production is far ahead of the same proportions on the U.S. side of the picture.

Given such scenarios, it is difficult to credit how serious headway can be made or expected in the direction of actually defending the ecological needs of the natural wilderness without first upholding the sovereign rights of the peoples involved.



Sources: NEB EIA, MMS, USGS

Map 3. Deregulated free market — or integration and subordination (NRCan Report, 2004)?

2.6 NATURAL GAS IN EURASIA: THE SPECIAL POSITION OF POST-SOVIET RUSSIA

When the situation of natural gas is compared to oil, the contrast could not be more striking. Even for what are commercially defined as natural gas sources, there is no shortage of supply, just some distribution constraints available for some degree of artificial and largely temporary manipulation. Most remarkably, one of the zones of greatest commercial supplies is the territory of the former Soviet Union, in which there never developed the upstream–downstream separation that facilitated U.S. domination, the manipulation of excess demand for oil (EDO), and the ever-expanding gap between the stagnation in real constant-dollar costs of production in oil-exporting countries and the price extracted for products in the consuming countries.

Another factor further transforming the picture stems from the absence of the corporate as well as government-owned sectors of the oil and gas industries of post-Soviet Russia from any of the existing U.S. or U.S.–European production or marketing cartels. The territories of the former Soviet Union (including the separate republics of central Asia which emerged since 1991) possess the largest conventional oil reserves outside OPEC and the second-largest proven reserves of natural gas anywhere on earth. Its emergence as a major world-market “player” in its own right in the production and sale of crude and refined petroleum for markets in Europe, especially for the fastest-growing markets in China and the Asian subcontinent, challenges both the demand-management model used by OPEC and backed by the U.S. oil majors to set and maintain a world oil price. As for natural gas, Russia provides the geographic linchpin for delivering to markets anywhere in the Eurasian land mass. For the more than four billion people living on that land mass, this fact must, over time, erode any notion of any purely local market for natural gas. The global distribution of commercial natural gas and its production are described.

Proven natural gas reserves — those that could be produced economically with the current technology — are illustrated in Figure 2-8 as follows. The former Soviet Union holds the world’s largest natural gas reserves, 38% of the world’s total. Together with the Middle East, which holds 35% of total reserves, they account for 73% of world natural gas reserves. In 2000 total world reserves were 150.19 trillion cubic meters. Global reserves more than doubled since the mid-1980s. The world’s ratio of proven natural gas reserves to production at current levels is about 60 years. This represents the time that remaining reserves would last if the present levels of production were maintained.

As Figure 2-9 illustrates, the world’s main natural gas-producing countries in 2000 were the United States (22.9% of world production) and the Russian Federation (22.5% of total production). Other major producing countries are Canada, United Kingdom, Algeria, Indonesia, Iran, Netherlands, Norway and Uzbekistan. These 10 countries alone accounted for more than 86% of total natural gas production in 2000. North America and the former Soviet Union together accounted for 59% of global production. Total world production in 2000 was 2422.3 billion cubic meters.

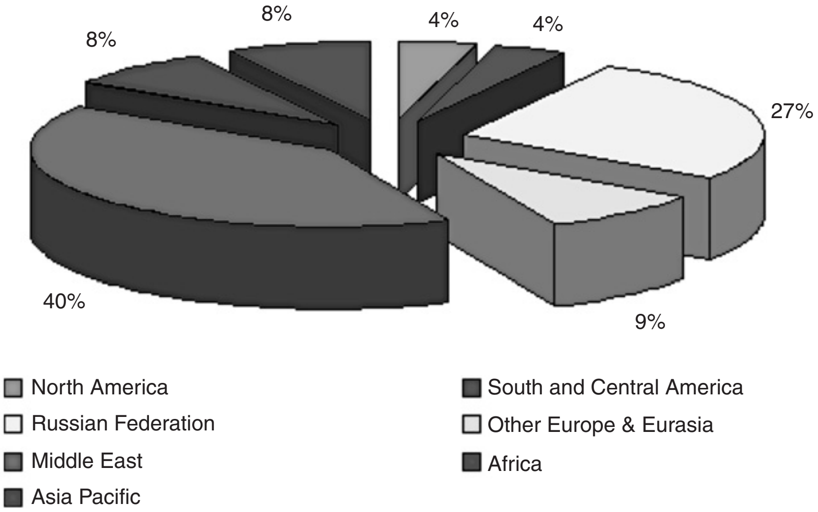


Figure 2-8. Distribution of proven natural gas reserves in 2003. Source: UNCTAD based on data from BP Amoco, Statistical Review of World Energy, June 2004.

Production growth in 2000 was 4.3%, a significantly higher growth rate than the 1990–2000 annual average. Although production increased in all regions, the faster growth was recorded in the Middle East and Africa. During the 1990s, production rose in all regions but the former Soviet Union.

Natural gas accounts for almost a quarter of the world's energy consumption. As shown clearly in the Figure 2-9, consumption of natural gas has increased considerably since the mid-1970s.

As Figure 2-10 illustrates, the world's main consuming countries in 2000 were the United States, accounting for 27.2% of total consumption, and the Russian Federation, with 15.7% of total consumption. North America and the former Soviet Union together consumed 55% of total natural gas. The share of Europe in total natural gas consumption was 19.1%. These three areas account for three-quarters of global consumption. Consumption growth was 4.8% in 2000, with the highest rates of growth registered in Africa (12.8%) and Asia (7.8%). Total world consumption was 2404.6 billion cubic meters.

Post-Soviet Russia's situation represents one of the foremost features of the world picture of commercial production of natural gas. Russia's natural gas industry, a monopoly run by Gazprom, which controls about 86%

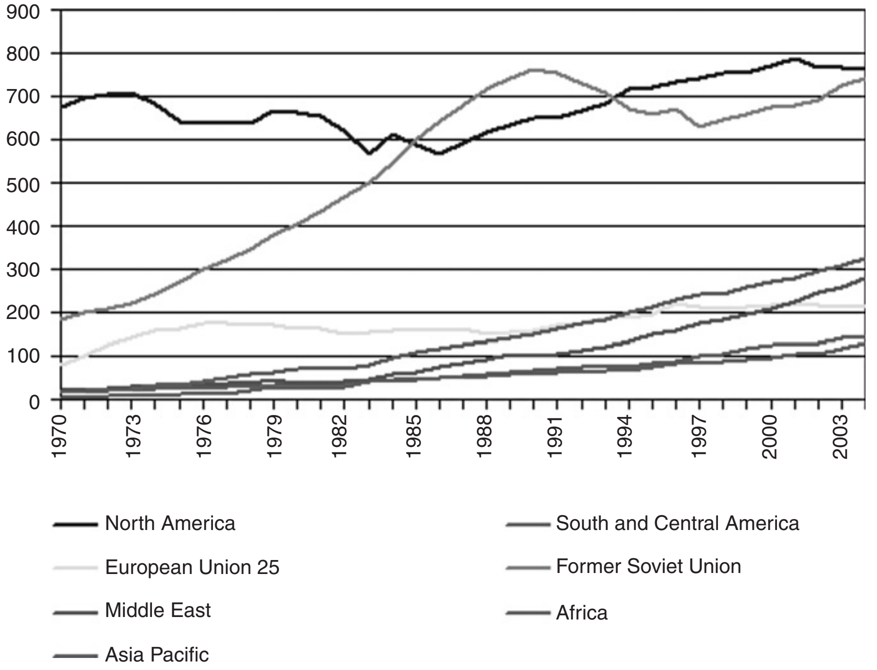


Figure 2-9. Natural gas production (billion cubic meters). Source: UNCTAD based on data from BP Amoco, Statistical Review of World Energy, June 2004.

of the country's marketed production, is the second major source of the world's commercial production of natural gas and the world's leading exporter of natural gas. Furthermore, within Russia, natural gas is the predominant industrial fuel, accounting for nearly half of the country's domestic energy consumption.

Another remarkable piece of the natural gas marketing picture related to this is the situation presented by western Europe, which possesses only about 5% of commercial natural gas resources but has an insatiable appetite for this form of energy. The main producing countries are Netherlands, Norway, and United Kingdom.

The gas industry in Europe consists mainly of downstream activities undertaken by transmission and distribution companies. More than 30% of gas consumption is met by pipeline imports from the former Soviet Union and Algeria, as well as liquefied natural gas imports from North Africa. The nearest and best source of supply is, logically, Russia, as Figure 2-11 shows.

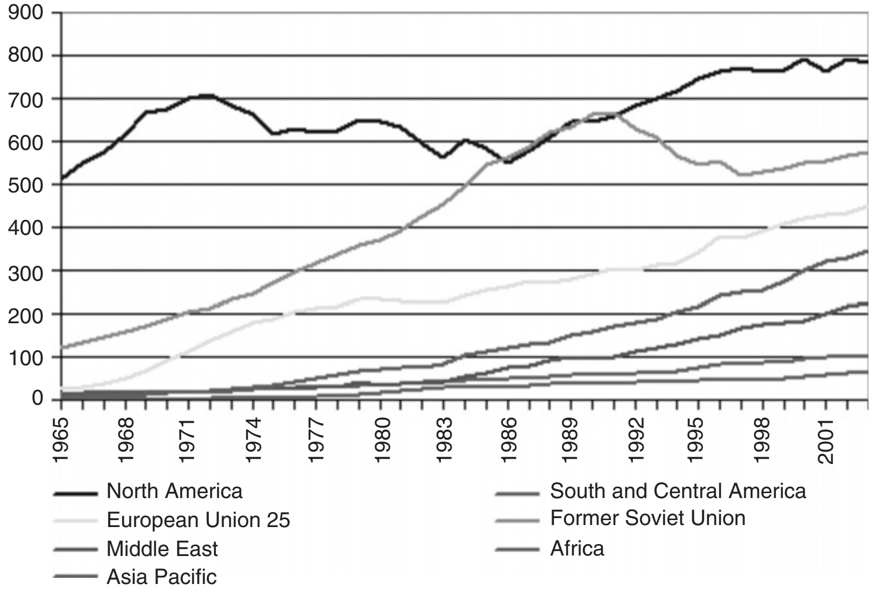


Figure 2-10. Natural gas consumption (billion cubic meters). Source: UNCTAD based on data from BP Amoco, Statistical Review of World Energy, June 2004.

The general picture in international trading of natural gas further illustrates how the dominant positions of the leading producers are mediated by the forces that determine the basic cost structures of international trade, such as shipping cartels. According to Cedigaz, 26.3% of total world marketed natural gas production was traded internationally. LNG tankers trade accounted for 21% of total international trade. The low share of international trade is due to the complexity to transport gas and requiring large investments, while many gas resources are far from consuming centers. The construction and management of pipelines also pose legal and logistical problems. Main exporting countries by pipeline in 2000 were the Russian Federation, Canada, Norway, Netherlands, Algeria, and United Kingdom. The main importing area by pipeline, apart from United States, which took hold of all Canadian exports, was Europe.

LNG trade itself presents an even more checkered picture. This takes place predominantly in Asia–Pacific, with Indonesia, Malaysia, and Australia as exporting countries and Japan as the main importing country.

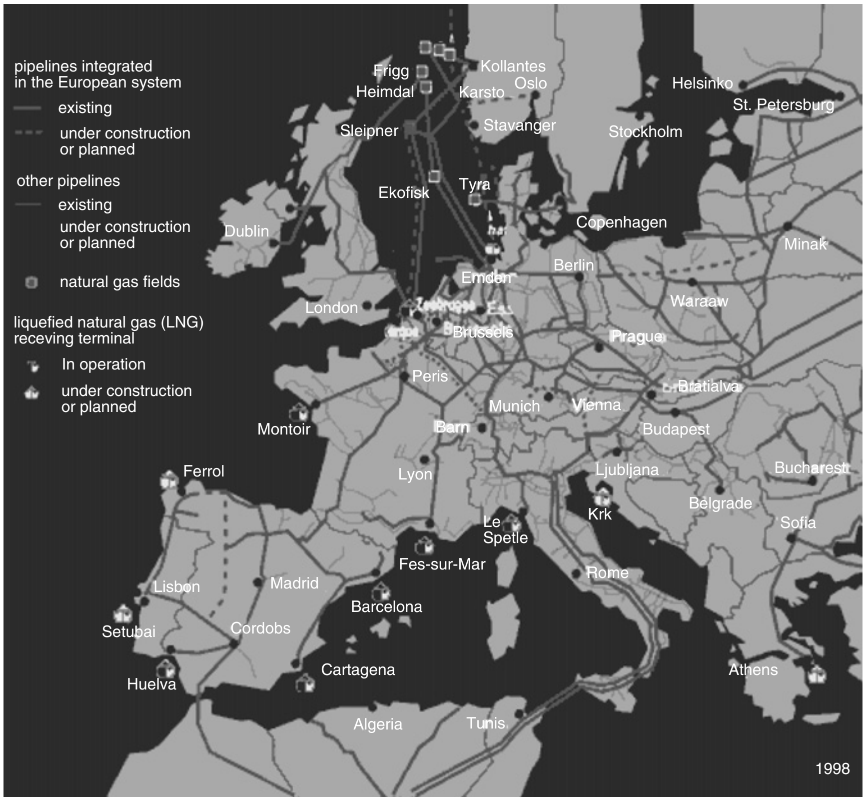


Figure 2-11. Natural gas pipelines in western and eastern Europe. Source: Eurogas Consortium.

Algeria and Qatar are also major exporters of LNG. The world LNG infrastructure as described by the GTI's World LNG Source Book (Sen, 2005) is as follows.

- Thirteen countries have liquefaction facilities: Abu Dhabi, Algeria, Australia, Brunei, Egypt, Indonesia, Libya, Malaysia, Nigeria, Oman, Qatar, Trinidad and Tobago, and the United States.
- Forty-seven receiving terminals are operating in 14 countries: 24 in Japan, 4 in Spain, 5 in the United States, 2 in Korea, 2 in France, 2 in India, and 1 each in Belgium, Dominican Republic, Greece, Italy, Puerto Rico, Taiwan Province of China, United Kingdom, and Turkey.

2.7 LOOKING TO NATURE FOR A NEW MODEL

Normally we would end the chapter on gas pricing right here. However, as much as we may document how it works, the fact remains that the pricing of natural gas up until now has reflected neither the true energy content of the gas nor the true costs of its production and delivery to consumers. Many costs have come to be added that also added further layers of opacity to the problem. For example, there is a problem with hydrate formation in gas pipelines that can severely restrict flow. Tens of billions of dollars annually are expended on toxic chemicals that are injected into pipelines to arrest, break up, or slow down hydrate formation. This cost is passed on to consumers, and then sometimes the pipeline company — but more often society, usually in the form of government — is then also confronted with paying the costs of environmental damage and cleanup from the toxic chemicals. This adds to the gross domestic product by simply increasing spending, but is it improving the standard of living of society as a whole?

The adding of costs in this manner — and with them, further layers of opacity — is reflecting no significant improvement in either of the two most intangible aspects of the energy commodity, namely, its efficiency or quality. Indeed, this process so effectively marginalizes all discussion or consideration of these aspects that the deeper question starts to pose itself: is any of this entire proceeding on the right path? Resolving this issue, however, is by no means as straightforward as it might at first appear. The commonest approach is the familiar one of "cost–benefit" analysis: the costs and benefits of various paths are totted up, compared, and finally one presumably optimal path is selected. Such quantitative comparison has the advantage of being highly tangible, but what if the problem resides precisely in the tangibility of all the paths thus examined, *i.e.*, the fact that they all quantify on the basis of putting dollar values on external manifestations?

There are alternatives to a number of the sources of added cost that follow a path quite different from that which can be directly engineered through adding chemicals. For example, bacteria can also break up hydrates and return to the environment without doing pollution damage. So far no one has found a way to turn such a thing into a multibillion dollar industry that would deliver a useful result. Some conclude from this state of affairs that such a course is uneconomic, but the fact of the matter is that doing this well would in the end be low cost, its essential mechanism (like many natural solutions) quite self-contained and thus unlikely to be spinnable into a grand new profit center. Of course this would also mean that a major increased charge on the final price of gas to the customer would be that much harder to justify in the name of defeating the hydrates menace and saving the environment.

The choices of path cannot be limited to rearranging and manipulating external features, elements that are extraneous to providing an energy source on a continuous and reliable basis. Natural paths are to be preferred over engineered paths, but the driver of the selection process remains "return on investment." Not all engineered paths will deliver an optimum return on investment, but when natural paths are examined, the problem becomes far more complicated. When "return on investment" is the driver, every possible natural path looks hopelessly utopian and impossible to justify.

What is the essence of the problem posed by continuing with energy commodity production on the basis that has developed up to the present? Many question whether any actual social need of society or its individual members is being addressed within the confines of any presently available social or economic model. At the same time, who or where anywhere in this modern world can survive without a system of providing and meeting energy needs? Could it be the very fact that return on investment is the driver is ensuring that, in the end, no actual need or needs of society or its individual members is or are being, or can be, addressed?

Hydrates, like many of the other hazards attending the exploration, production, distribution, and consumption of natural gas, petroleum, and its by-products, become irremovable singularities that can only be dealt with either at the cost of private corporate profit or at public government expense. They are labeled "downsides" and entire forests have been felled to produce the thousands upon thousands of pages of statutes and regulations that are supposed to "manage" the problems created by these "downsides," but nothing fundamental is transformed in the result. Why? Could the answer be because these approaches all follow one and the same antinature model?

The antinature model takes on too many specific forms to be easily pigeon holed with a concise definition. However, in the case of energy commodity production, the production of energy mainly as a feedstock to produce energy in another form — oil or natural gas to make electricity or renewable solar energy for the same purpose — rather than for their highest and best uses directly as energy sources in their own right is one of the most commonly seen indicators of an antinature path and agenda.

The real issue is firmly and finally to step off this antinature path and renew the knowledge-based, researched approach to extracting solutions that are innovative, economically attractive, environmentally appealing, and socially responsible.

The production, distribution, and supply of energy are keys to modern social and economic existence everywhere on this planet. There is no corner left on this earth where the solution to these problems can be purely local or not connected with the wider world. How problems arising in the production, distribution, and supply of energy are addressed should to be the heart and soul of solving the problem of raising the true standard of living everywhere, the notion crudely and incompletely estimated by the well-known index forming the ordinate axis in Figure 2-12, namely, the "per capita gross domestic product."

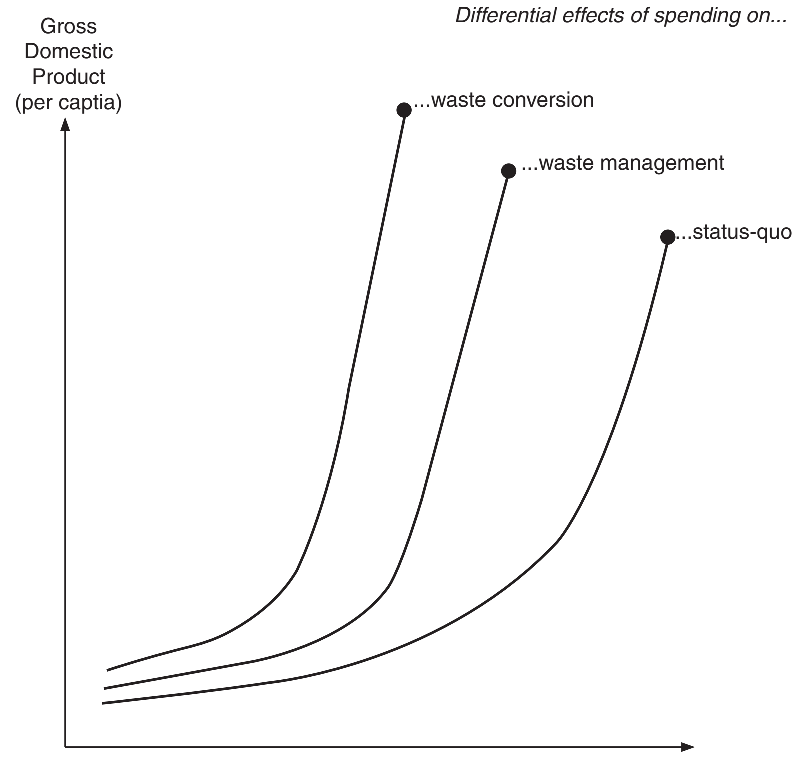


Figure 2-12. Deeds are but by intention, illustrating the longer-term effects on GDP of redirecting spending to socially useful ends.

Of course, spending of all kinds is incorporated in the per capita Gross Domestic Product (GDP), from prostitution, gambling, and tobacco consumption, to armaments production and export, to outer space research and medical research, etc. The idea underlying per capita GDP is that even if individuals, institutions, and corporations each have their own priorities, personal, private, or social, a meaningful statistic about the general state of social development as a whole can nevertheless be generated by averaging spending on all these wildly divergent agendas over the entire population. The legitimate question raised by this approach is what social progress other than mere arithmetic increases in spending can be registered by such a statistic? Mixing together privately directed and socially directed forms of spending on the assumption that their very direction, *i.e.*, the agendas informing the actual spending decisions, is a matter of indifference to economic science ensures that spending that is indeed harmful and informed by some private agenda is placed on par with spending that is socially beneficial and informed by an agenda that is not purely self-serving. Yet it is widely understood as a fact of life anywhere in this world that “good for some and bad for others is bad for everyone.” Allegedly “scientific”-minded indifference to this crucial distinction acts as a serious blind to understanding the fundamental issue of path. What the graph brings out is how crucial it is for collective well-being that social concerns are assigned their proper priority: any change in this direction, even just a focus on waste management, would being about an improvement compared to the status quo.

The pro-nature path is attained quite readily by taking a knowledge-based and researched approach. This means starting from the position that nothing that presently exists is the “last word” in technology. One must be ready fearlessly to work out arrangements based on what is actually found to work in nature, not on the basis that nature is broken and needs to be “fixed.” Nothing that originates from nature can pollute the natural environment if it is used and returned to nature without in the meantime being artificially reconstituted, deconstituted, or otherwise engineered outside the range within which it serves its customary natural purpose(s). Nothing that originates in nature is linear in its native state; if it is linearized and then returned to nature it will either pollute or cease to be useful. Natural constituents utilized according to their customary functions within nature are in fact time tested. This often ensures overcoming, by entirely avoiding, the costly burden of having to reengineer short-term solutions developed away from a nature-based path as they inevitably break down and start to fail.

What is not being suggested here is either any romanticized notion about nature as a mysterious force or mysterious treasure house, nor a theological or teleological assumption about the purposes of nature. It is an unavoidable starting point of all scientific investigation, into either social or natural-physical phenomena, that the universe at any moment is a function of space and time: $U = f(x, y, z; t)$. The actual space relevant to any proper specific investigation is not to be confused with, or held to be coequal with universal space, the "x, y, z" of U as just defined. A proper investigation depends on finding ways of investigating — perceiving and measuring — the portion of space that is relevant. The temporal component, however, is a different matter.

The temporal component of some finite process or system being observed or investigated, which can be defined as commencing at t_{initial} and ending at t_{final}, is one thing. However, the temporal component of processes in nature proper, so to speak, is very different, precisely because — frequently if not indeed customarily — t_{final} is simply unknown. All engineered processes can be handled very nicely according to the paradigm of the [t_{initial}, t_{final}] interval, whereas most natural processes can be investigated at best only according to an interval that looks something like [t_{initial} , $t_($finalbest – guess)], where the right-hand side of the interval cannot be closed. It could very well be that it closes at $t = \infty$. The one thing that is certain is that right-hand ")" can only be replaced with a "]" *after* a true t_{final} value is established, not before. From the standpoint of establishing scientific knowledge, it will not advance matters simply to assert and assume a t_{final} value and a right-hand "]" without some testing procedure and associated criteria that either establish a t_{final} value or demonstrate that t_{final} only appears or converges at $t = \infty$. The temporal component of a natural process is thus inherently intangible. Usually at this point, accusations are heard about "introducing matters of metaphysics and religion into science," etc. However, should acknowledging such intangibility have to be the end of further scientific discussion and investigation or could it not represent a new starting point?

Here it becomes important to clarify that what is being set forth is very far removed from the notions of intangibility presently attending the entire matter of natural gas pricing. These are quite concentrated, for example, in the specific area of risk management, specifically using derivatives (see Section 2.5.3):

> The asset that underlies a derivative can be a physical commodity (e.g., crude oil or wheat), foreign or domestic currencies, treasury bonds, company stock, indices representing the value of groups of securities or commodities, a service, *or even an intangible commodity such as a weather-related index (e.g., rainfall, heating degree days, or cooling degree days)*. What is critical is that the value of the underlying commodity or asset be unambiguous; *otherwise, the value of the derivative becomes ill-defined*." [Emphases added] (EIA, 2002)

The problem here lies with how intangible elements have been selected in the first place for inclusion among the components of the derivatives. The effective value of the derivative as a tradable instrument is entirely a function of time rendered in tangible form, *namely* the date(s) on which the underlying contract(s) must be completed. The intangibility of the "weather-related index" mentioned in the example earlier, however, is intended to provide the seller of the instrument with an "out," an escape from liability, in the event of a failure to perform the contract. Time as the essential intangible of processing of anything within nature, however, either adds or subtracts effective value. Issues of liability, much less distancing contracting parties from its reach, do not arise.

Viewed in the light that has been cast here, the time testedness of natural constituents and sources, before and without further chemically or industrially engineered processing, poses an interesting conundrum: are the best sources of energy locked in nature's newest products, or its oldest? The conventionally trained scientist's response would be that all of this is resolved by establishing the chemical structure: CH_4 from the living cow at around 1 KPa pressure or CH_4 from material that decayed eons ago, lying thousands of meters beneath the Earth's surface under confining pressures of thousands of KPa is also CH_4: end of story.

Consider, however, the implications of this claim:

- It is accepted that the entire planetary system and galaxies far beyond this planet and solar system are continuously undergoing change, and
- it is accepted that animal species have undergone and continue to undergo evolutionary change, speciation etc., but
- it is not acceptable to assume anything consequential might have happened at the molecular level over the billions of years that passed while some electron continued revolving about some nucleus.

The weight of this logic alone suggests that an important piece of the picture remains to be investigated further. Despite what has become commonly accepted, and which forms the basis of the entire energy extraction and processing sector, the energetic content of various well-defined molecular arrangements of carbon and hydrogen atoms may in the final analysis indeed *not* have been due after all, either mainly or entirely, to how these compounds as raw material were engineered (*i.e.*, processed or refined). Rather, these molecular remains of "dead" matter may have further transformed internally as raw material in the natural environment according to processes still not understood. Just because such material is no longer living organic material may not mean that it could not undergo further internal transformation on top of the transformations being imposed externally by continental drift and plate movement, climate change, etc. Within the field of petroleum engineering research, to dismiss any consideration of the possibility of subatomic intramolecular change in posthumous organic matter left undisturbed over geological periods of time cannot be considered a sign of wisdom. In the present context, it would even seem to serve as evidence for the conclusion that ongoing fanatical adherence to views that subsequent progress in actual research and discovery rendered unsustainable is a syndrome that cannot be attributed to any particular belief system, national or ethnic origin. All that is required to catch this disease is a strong will to remain ignorant.

As energy is in continual demand, solutions are needed on a continuous basis and that which can be shown to work in the long term can therefore take care of the short term as well. That is why a pro-nature-oriented scientific research effort could reasonably be expected to produce innovative breakthroughs not confined to managing pollution and waste but to conversion of so-called waste into useful new social products. The energy potential locked within many so-called "waste materials," researched and developed on a nature-friendly basis, could produce a tremendous improvement in the overall standard of living by rendering energy supplies affordable and available to broad sections of the population that are currently underserved and overcharged.

When it comes to problems of international economic development, the development of energy sources cannot be discussed seriously or honestly without addressing the critical issues of groundwater supply. This is the most fundamental metric of true living standards in the still overwhelmingly rural societies of Asia, Africa, and Latin America. The entire bias in favor of engineered solutions as "modern" and "sophisticated" has served to accelerate the departure from the pro-nature path. Before the arrival, starting in the 17th century (1600s), of colonizing powers from Europe and the United States in these regions, this path was the norm of many so-called "traditional" societies; it was the same story among the native peoples of North and Central America, who suffered an outright genocide that began a century earlier.

A powerful movement has spread far and wide in recent years in many of these countries to restore traditional technologies, renewed appropriately and sufficiently to deal adequately with the speed of information and development in the modern world but otherwise traditional in their fundamental approach, which was and had to be pro-nature by definition. This accounts for the success of such initiatives as the "Water University," which has developed in several drought-stricken regions of the Asian subcontinent to recapture and reapply ancient traditional knowledge — dismissed for the last century and more as "old wives' tales," folklore," "anecdotes but not science" etc. — of how to use the ambient natural conditions to conserve precious supplies of groundwater. The rapidity and scale of this success represent a most telling indictment as well of the signal failure of allegedly "advanced technological solutions" to lift more than a tiny elite from these societies into the ranks of "civilized" colonizers.

However, there is something yet more fundamental here than the confrontation between traditional societies and European colonization: the dependence of modern societal infrastructures on energy supplies is matched only by, and ultimately even underpinned by, the broader overall dependence of human social existence on the supply of water. The intimate connections between groundwater supplies and oil reservoirs were well understood in ancient times throughout Mesopotamia. One of the gravest indictments of the arrival of "advanced modern technology" to exploit the oil wealth of the region was the serious messing up of groundwater supplies that followed. The significance was multiplied greatly by the circumstance that these regions were literally at the edge of the westward-moving front of the central Asia desert. In the American southwest in particular, as activity in the civil courts serves to indicate, commercial natural gas development has enormously disrupted groundwater supplies and aquifers on the traditional lands of many native American tribes.

The corporate mantra of the true believers in "return on investment" as the final arbiter of the possible seems to be "if oil and water don't mix, forget the water!" Today, this stands increasingly exposed as a path imperiling the very future of humanity. The fact that one cannot sacrifice either the one or the other in the short term or the long term is increasingly well understood among all peoples everywhere. At the same time, Nature ultimately cannot accept the confines of any system of so-called private property. It has its own logic, which proceeds utterly oblivious to such inventions and conventions. That is why those following the corporate road must inevitably at some point take the antinature path — the path of "fixing" nature rather than learning its secrets and working with it, a path with a doubtful future even in the short term and no future whatsoever in the long term.

Earlier in this chapter, the question was posed: considered from an engineering standpoint, are the patterns, changes, and trends in the supply of and demand for so-called "nonrenewable" energy commodities consistent with their actual potential? Our best answer at the moment is that the role of "return on investment" — as the driver matching the supply of energy commodities to an asserted demand that, more often than not, has been manipulated by hidden hands — seems to have blocked humanity's access to the pro-nature path. This is the only path on which the actual potential for these commodities can be researched properly and the problems involved in releasing their potential solved in ways that will be innovative, economically attractive, environmentally appealing, and socially responsible.

In conclusion, natural gas has a key role to play as an energy source in the 21st century, and the demand, which will be driven by the development of new markets for natural gas, is growing steadily. The share of natural gas reserves located onshore, easy to produce and close to consumers, is decreasing while the share of natural gas reserves located offshore in challenging and hostile environments is growing. Technology is the driver, and current "stranded" gas resources may be developed using floating LNG plants, which could not be technically feasible just a few years ago. The major challenge of the gas industry is to be able to bring gas to the consumers economically. Substantial cost reductions achieved through technical progress have resulted in longer distance transports. LNG developments have also contributed to this evolution. In this case, gas transportation involves heavy and expensive infrastructures, which result in rather rigid ties between the producer and the consumer (Mokhatab and Purewal, 2006). Furthermore, a major part of transnational trade will be only feasible by long-distance gas pipelines. Natural gas transportation by pipelines is discussed in Chapter 3.

REFERENCES

Anadarko Petroleum, Current information about company activities available at http://www.anadarko.com (last accessed Sept. 14, 2005).

Aumann, R.J., Markets with a Continuum of Traders, *Econometrica* (1964).

Aumann, R.J., Subjectivity and correlation in randomized strategies, *J. Math. Econ.* (1974).

Aumann, R.J., and Branderburger, A., Correlated equilibrium as an expression of Bayesian rationality, *Econometrica* (1987).

Aumann, R.J., and Hart, S., "Handbook of Game Theory with Economic Applications," Elsevier (Vol. 1: 1992, Vol. 2: 1994, Vol. 3: 2002).

Bryce, R., "Royal Commission on Corporate Concentration," Study Report No. 16. The Irving Group, Queen's Printer, Ottawa (1978).

EIA, "Derivatives and Risk Management in the Petroleum, Natural Gas, and Electricity Industries," Energy Information Administration (EIA), U.S. Department of Energy, Washington, DC (Oct. 2002).

EIA, "Canada — Country Analysis Brief," Energy Information Administration (EIA), U.S. Department of Energy, Washington, DC (Feb. 2005).

EPA, "Petroleum Industry," EPA AP-42 Manual, Chapter 5.3. Environmental Protection Agency (EPA), Washington, DC (Jan. 1995).

EPA, "Economic Impact Analysis of the Oil and Natural Gas Production NESHAP and the Natural Gas Transmission and Storage NESHAP." Final Report (EPA-452/R-99-003), Chapter 2, Sections 2.2-3 on Natural Gas Production and Markets in the U.S. for the Period 1983–1993 and Chapter 4, Sections 1.1-3 on the Modeling of Market Adjustments to the Introduction of Deregulation of Natural Gas in the U.S., Environmental Protection Agency (EPA), Washington, DC (May 1999).

Fleay, B.J., "Climaxing Oil: How Will Transport Adapt?" Paper presented at the Chartered Institute of Transport in Australia National Symposium, Launceston (1998).

Islam, M.R., and Zatzman, G.M., "A New Energy Pricing Model," Proceeding of the 8th Mediterranean Petroleum Conference & Exhibition (MPC-2004), 210-226 Tripoli, Libya (2004).

Islam, M.R., and Zatzman, G.M., "A New Energy Pricing Model: Exploring the Evolving Relationship between International Reserve Currencies and Global Shifts in Access to and Control Over

Strategic Energy Resources," Paper presented at the 3rd international Conference on Energy Resource Development (ICERD-3), Kuwait (2005).

Izrael, Yu. A., *et al*., "Russian Federation Climate Change Country Study (Cooperative Agreement DE-FCO2-93PO10118), Final Report (Vol. 1 – Inventory of Technogenic GHG Emissions), Moscow, Russian Federal Service for Hydrometeorology and Environmental Monitoring (RFSHEM) (1997).

LIHEAP Clearinghouse, "An Overview and History of Gas Deregulation," Available at http://www.liheap.ncat.org/dereg/gasoview.htm (last accessed Sept. 14, 2005).

McQuaig, L., "It's the Crude, Dude: War, Big Oil, and the Fight for the Planet," Toronto-New York: Doubleday (2004).

Mokhatab, S., and Purewal, S., "Is LNG a competitive source of natural gas?" J. *Petroleum Sci. & Technol* **24**(2), 243–245 (2006).

Mullins, B.J., *et al*., "Atmospheric Emissions Survey of the Sour Gas Processing Industry," EPA-450/3-75-076, U.S. Environmental Protection Agency (EPA), Washington, DC (1975).

Natural Gas Organization: History of Regulation, at http://www.naturalgas.org/regulation/history.asp#earlydays (last accessed Sept. 14, 2005).

NRCan (Natural Resources Canada) Report, "Canadian Natural Gas: Review of 2003 and Outlook to 2020," Ottawa. Energy Policy Sector – Petroleum Resources Branch – Natural Gas Division, Ottawa, Canada (Dec. 2004).

Sen, C.T., "GTI's World LNG Source Book 2005," Gas Technology Institute (GTI), Des Plaines, Illinois (2005).

Udall, R., and Andrews, S., "Methane Madness: A Natural Gas Primer," Denver CO: Community Office for Resource Efficiency (2001).

US Department of Energy, Data about US Natural Gas Markets in 2005, spreadsheeted at http://tonto.eia.doe.gov/dnav/ng/ng_sum_lsum_dcu_nus_m.htm (2005).

CHAPTER 3

RAW GAS TRANSMISSION

3.1 INTRODUCTION

Natural gas is often found in places where there is no local market, such as in the many offshore fields around the world. For natural gas to be available to the market, it must be gathered, processed, and transported. Quite often, collected natural gas (raw gas) must be transported over a substantial distance in pipelines of different sizes. These pipelines vary in length between hundreds of feet to hundreds of miles, across undulating terrain with varying temperature conditions. Liquid condensation in pipelines commonly occurs because of the multicomponent nature of the transmitted natural gas and its associated phase behavior to the inevitable temperature and pressure changes that occur along the pipeline. Condensation subjects the raw gas transmission pipeline to two-phase, gas/condensate, flow transport. Hence, a better understanding of the flow characteristics is needed for the proper design and operation of pipelines. The problem of optimal design of such pipelines becomes accentuated for offshore gas fields, where space is limited and processing often is kept to a minimum; therefore, total production has to be transported via multiphase pipelines. These lines lie at the bottom of the ocean in horizontal and near-horizontal positions and may contain a three-phase mixture of hydrocarbon condensate, water (occurring naturally in the reservoir), and natural gas flowing through them.

Multiphase transportation technology has become increasingly important for developing marginal fields, where the trend is to economically transport unprocessed well fluids via existing infrastructures, maximizing the rate of return and minimizing both capital expenditure (CAPEX) and operational expenditure (OPEX) (Klemp, 1999). In fact, by transporting multiphase well fluid in a single pipeline, separate pipelines and receiving facilities for separate phases, costing both money and space, are eliminated, which reduces capital expenditure. However, phase separation and reinjection of water and gas save both capital expenditure and operating expenditure by reducing the size of the fluid transport/handling facilities and the maintenance required for the pipeline operation (Hill, 1997). Given the savings that can be available to the operators using multiphase technology, the market for multiphase flow transportation is an expanding one. Hence, it is necessary to predict multiphase flow behavior and other design variables of gas-condensate pipelines as accurately as possible so that pipelines and downstream processing plants may be designed optimally. This chapter covers all the important concepts of multiphase gas/condensate transmission from a fundamental perspective.

3.2 MULTIPHASE FLOW TERMINOLOGY

This section defines the variables commonly used to describe multiphase flow. For example, the general pressure drop equation for multiphase (two and three phase) flow is similar to that for single-phase flow except some of the variables are replaced with equivalent variables, which consider the effect of multiphase. The general pressure drop equation for multiphase flow is as follows (Brill and Beggs, 1991):

$$\left(\frac{dP}{dx}\right)_{\text{tot}} = \left(\frac{dP}{dx}\right)_{\text{ele}} + \left(\frac{dP}{dx}\right)_{\text{fri}} + \left(\frac{dP}{dx}\right)_{\text{acc}} \tag{3-1}$$

where

$$\left(\frac{dP}{dx}\right)_{\text{ele}} = \rho_{\text{tp}}\left(\frac{g}{g_{\text{c}}}\right)\sin\theta \tag{3-1-1}$$

$$\left(\frac{dP}{dx}\right)_{\text{fri}} = \frac{\rho_{\text{tp}}\, f_{\text{tp}} V_{\text{tp}}^{2}}{2 g_{\text{c}} D} \tag{3-1-2}$$

$$\left(\frac{dP}{dx}\right)_{\text{acc}} = \frac{\rho_{\text{tp}}\, f_{\text{tp}}}{g_{\text{c}}}\left(\frac{dV_{\text{tp}}}{dx}\right) \tag{3-1-3}$$

where $\frac{dP}{dx}$ is flow pressure gradient, x is pipe length, ρ is flow density, V is flow velocity, f is friction coefficient of flow, D is internal diameter of

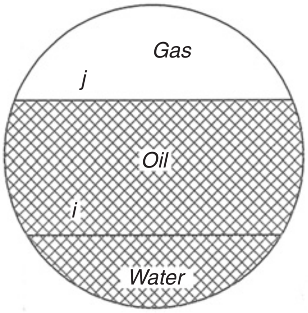


Figure 3-1. Three-phase flow pipe cross section (Taitel *et al.*, 1995).

pipeline, θ is inclination angle of pipeline, g is gravitational acceleration, and g_c is gravitational constant. The subscripts are "tot" for total, "ele" for elevation, "fri" for friction loss, "acc" for acceleration change terms, and "tp" for two- and/or three-phase flow.

Due to different flow mechanisms for different flow patterns, the multiphase (two/three phase) flow parameters used in the aforementioned pressure gradient equations should be defined separately. Definitions for commonly used multiphase variables are described based on Figure 3-1. The ideal flow of three fluids is considered in Figure 3-1: water, oil, and gas. It is assumed that the water is heavier than the oil and flows at the bottom, while the oil flows in the middle and the gas is the top layer.

3.2.1 Superficial Velocity

The superficial velocity is the velocity of one phase of a multiphase flow, assuming that the phase occupies the whole cross section of pipe by itself. It is defined for each phase as follows:

$$V_{\mathrm{SW}} = \frac{Q_{\mathrm{W}}}{A} \tag{3-2}$$

$$V_{\mathrm{SO}} = \frac{Q_{\mathrm{O}}}{A} \tag{3-3}$$

$$V_{\mathrm{SG}} = \frac{Q_{\mathrm{G}}}{A} \tag{3-4}$$

where

$$A = A_{\mathrm{W}} + A_{\mathrm{O}} + A_{\mathrm{G}} \tag{3-5}$$

The parameter A is the total cross-sectional area of pipe, Q is volumetric flow rate, V is velocity, and the subscripts are W for water, O for oil, G for gas, and S for superficial term.

3.2.2 Multiphase Flow Mixture Velocity

Mixture velocity is the sum of phase superficial velocities:

$$V_{\mathrm{M}} = V_{\mathrm{SW}} + V_{\mathrm{SO}} + V_{\mathrm{SG}} \tag{3-6}$$

where V_{M} is the multiphase mixture velocity.

3.2.3 Holdup

Holdup is the cross-sectional area, which is locally occupied by one of the phases of a multiphase flow, relative to the cross-sectional area of the pipe at the same local position.

For the liquid phase,

$$H_{\mathrm{L}} = \frac{A_{\mathrm{L}}}{A} = \frac{A_{\mathrm{W}} + A_{\mathrm{O}}}{A} = H_{\mathrm{W}} + H_{\mathrm{O}} \tag{3-7}$$

For the gas phase,

$$H_{\mathrm{G}} = \frac{A_{\mathrm{G}}}{A} \tag{3-8}$$

where the parameter H is the phase holdup and the subscripts are L for the liquid and G for the gas phase.

Although “holdup” can be defined as the fraction of the pipe volume occupied by a given phase, holdup is usually defined as the *in situ* liquid volume fraction, whereas the term “void fraction” is used for the *in situ* gas volume fraction (Hasan and Kabir, 2002).

3.2.4 Phase Velocity

Phase velocity (*in situ* velocity) is the velocity of a phase of a multiphase flow based on the area of the pipe occupied by that phase. It may also be defined for each phase as follows:

$$V_L = \frac{V_{SL}}{H_L} = \frac{V_{SW} + V_{SO}}{H_L} \tag{3-9}$$

$$V_G = \frac{V_{SG}}{H_G} \tag{3-10}$$

3.2.5 Slip

Slip is the term used to describe the flow condition that exists when the phases have different phase velocities. The slip velocity is defined as the difference between actual gas and liquid velocities, as follows:

$$V_S = V_G - V_L \tag{3-11}$$

The ratio between two-phase velocities is defined as the slip ratio. If there is no slip between the phases, $V_L = V_G$, and by applying the no-slip assumption to the liquid holdup definition, it can be shown that

$$H_{L,\text{no-slip}} = \lambda_L = \frac{V_{SL}}{V_M} \tag{3-12}$$

Investigators have observed that the no-slip assumption is not often applicable. For certain flow patterns in horizontal and upward inclined pipes, gas tends to flow faster than the liquid (positive slip). For some flow regimes in downward flow, liquid can flow faster than the gas (negative slip).

3.2.6 Multiphase Flow Density

Equations for two-phase gas/liquid density used by various investigators are as follows (Govier and Aziz, 1972):

$$\rho_S = \rho_L H_L + \rho_G H_G \tag{3-13}$$

$$\rho_n = \rho_L \lambda_L + \rho_G \lambda_G \tag{3-14}$$

$$\rho_k = \frac{\rho_L \lambda_L^2}{H_L} + \frac{\rho_G \lambda_G^2}{H_G} \tag{3-15}$$

Equation (3-13) is used by most investigators to determine the pressure gradient due to elevation change. Some correlations are based on the assumption of no slippage and therefore use Equation (3-14) for two-phase density. Equation (3-15) is used by some investigators to define the density used in the friction loss term and in the Reynolds number (Brill and Beggs, 1991). In the equations shown earlier, the total liquid density can be determined from the oil and water densities and flow rates if no slippage between these liquid phases is assumed:

$$\rho_L = \rho_O f_O + \rho_W f_W \tag{3-16}$$

where

$$f_O = \frac{Q_O}{Q_O + Q_W} = 1 - f_W \tag{3-17}$$

where the parameter *f* is the volume fraction of each phase.

3.3 MULTIPHASE FLOW REGIMES

Multiphase flow is a complex phenomenon that is difficult to understand, predict, and model. Common single-phase flow characteristics such as velocity profile, turbulence, and boundary layer are thus inappropriate for describing the nature of such flows. The flow structures are rather classified in flow regimes, whose precise characteristics depend on a number of parameters. Flow regimes vary depending on operating conditions, fluid properties, flow rates, and the orientation and geometry of the pipe through which the fluids flow. The transition between different flow regimes may be a gradual process. Due to the highly nonlinear nature of the forces that rule the flow regime transitions, the prediction is near impossible. In the laboratory, the flow regime may be studied by direct visual observation using a length of transparent piping (Wallis, 1969). However, the most utilized approach is to identify the actual flow regime from signal analysis of sensors whose fluctuations are related to the flow regime structure. This approach is generally based on average cross-sectional quantities, such as pressure drop or cross-sectional liquid holdup. Many studies have been documented using different sensors and different analysis techniques (Dukler and Hubbard, 1966; Jones and Zuber, 1975; Lin and Hanratty, 1987; Rajkovic *et al*., 1996; Soldati *et al*., 1996).

In order to obtain optimal design parameters and operating conditions, it is necessary to clearly understand two- and three-phase flow regimes and the boundaries between them, where the hydrodynamics of the flow, as well as the flow mechanisms, change significantly from one flow regime to another (Cheremisinoff, 1986). If an undesirable flow regime is not anticipated in the design, the resulting flow pattern can cause system pressure fluctuation and system vibration and even mechanical failures of piping components.

3.3.1 Two-Phase Flow Regimes

The description of two-phase flow can be simplified by classifying types of gas–liquid interfacial distribution and calling these “flow regimes” or “flow patterns.” The distribution of the fluid phases in space and time differs for the various flow regimes and is usually not under the control of the pipeline designer or operator (Amdal *et al*., 2001).

3.3.1.1 Horizontal Flow Regimes

Two-phase flow regimes for horizontal flow are shown in Figure 3-2. These horizontal flow regimes are defined as follows.

3.3.1.1.1 Stratified (Smooth and Wavy) Flow Stratified flow consists of two superposed layers of gas and liquid, formed by segregation under the influence of gravity. The gas–liquid interface is more or less curved and either smooth or rough because of capillary or gravity forces. The curvature of the interface increases with the velocity of the gas phase (Lagiere *et al*., 1984).

3.3.1.1.2 Intermittent (Slug and Elongated Bubble) Flow The intermittent flow regime is usually divided into two subregimes: plug or elongated bubble flow and slug flow. The elongated bubble flow regime can be considered as a limiting case of slug flow, where the liquid slug is free of entrained gas bubbles (Cindric *et al*., 1987). Although these flow regimes are quite similar to each other, their fluid dynamic characteristics are very different and greatly influence such quantities as pressure drop and

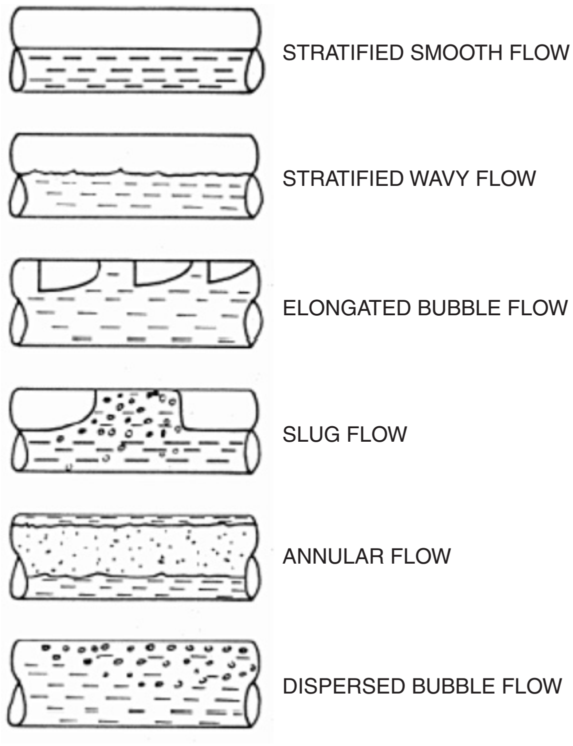


Figure 3-2. Horizontal two-phase flow regimes (Cindric *et al.*, 1987).

slug velocity (Bertola and Cafaro, 2001). Gas–liquid intermittent flow exists in the whole range of pipe inclinations and over a wide range of gas and liquid flow rates (VanHout *et al.*, 1992). It is characterized by an intrinsic unsteadiness due to regions in which the liquid slugs fill the whole pipeline cross section and regions in which the flow consists of a liquid layer and a gas layer. The presence of these slugs can often be troublesome in the practical applications (giving rise to sudden pressure pulses, causing large system vibration and surges in liquid and gas flow rates), and the prediction of the onset of slug flow is of considerable industrial importance.

3.3.1.1.3 Annular Flow During annular flow, the liquid phase flows largely as an annular film on the wall with gas flowing as a central core.

Some of the liquid is entrained as droplets in this gas core. The annular liquid film is thicker at the bottom than at the top of the pipe because of the effect of gravity and, except at very low liquid rates, the liquid film is covered with large waves.

3.3.1.1.4 Dispersed Bubble Flow At high liquid rates and low gas rates, the gas is dispersed as bubbles in a continuous liquid phase. The bubble density is higher toward the top of the pipeline, but there are bubbles throughout the cross section. Dispersed flow occurs only at high flow rates and high pressures. This type of flow, which entails high-pressure loss, is rarely encountered in flow lines (Lagiere *et al.*, 1984).

Note that raw gas pipelines usually have stratified smooth/wavy flow patterns. This arises because flow lines are designed to have appreciable velocities and the liquid content is usually quite low. Annular flow can also occur but this corresponds to high velocities, which are avoided to prevent erosion/corrosion, etc. In other words, raw gas lines are “sized” to be operated in stratified flow during normal operation.

3.3.1.2 Vertical Flow Regimes

Flow regimes frequently encountered in upward vertical two-phase flow are shown in Figure 3-3. These flow regimes tend to be somewhat more simpler than those in horizontal flow. This results from the symmetry in the flow induced by the gravitational force acting parallel to it. A brief description of the manner in which the fluids are distributed in the pipe for upward vertical two-phase flow is as follows. It is worth noting that vertical flows are not so common in raw gas systems (i.e., wells normally have some deviation and many risers are also inclined to some extent).

3.3.1.2.1 Bubble Flow The gas phase is distributed in the liquid phase as variable-size, deformable bubbles moving upward with zigzag motion. The wall of the pipe is always contacted by the liquid phase.

3.3.1.2.2 Slug Flow Most of the gas is in the form of large bullet-shaped bubbles that have a diameter almost reaching the pipe diameter. These bubbles are referred to as “Taylor bubbles,” move uniformly upward, and are separated by slugs of continuous liquid that bridge the pipe and contain small gas bubbles. Typically, the liquid in the film around the Taylor bubbles may move downward at low velocities, although the net

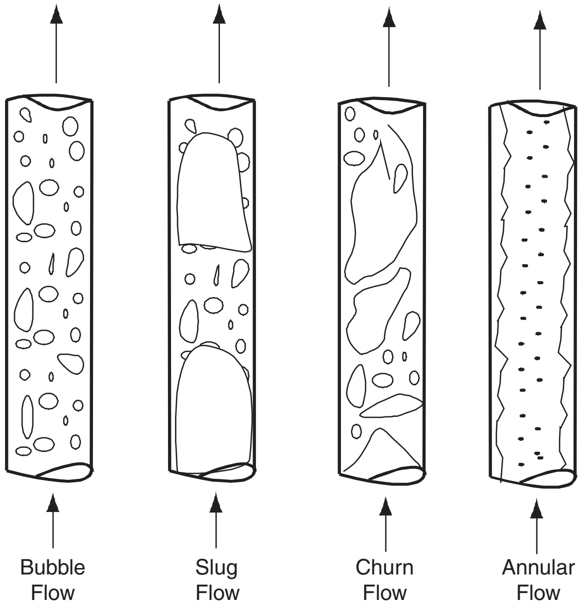


Figure 3-3. Upward vertical two-phase flow regimes (Shoham, 1982).

flow of liquid can be upward. The gas bubble velocity is greater than that of the liquid.

3.3.1.2.3 Churn Flow If a change from a continuous liquid phase to a continuous gas phase occurs, the continuity of the liquid in the slug between successive Taylor bubbles is destroyed repeatedly by a high local gas concentration in the slug. This oscillatory flow of the liquid is typical of churn flow. It may not occur in small-diameter pipes. The gas bubbles may join and liquid may be entrained in the bubbles.

3.3.1.2.4 Annular Flow Annular flow is characterized by the continuity of the gas phase in the pipe core. The liquid phase moves upward partly as a wavy film and partly in the form of drops entrained in the gas core.

Although downward vertical two-phase flow is less common than upward flow, it does occur in steam injection wells and down comer pipes from offshore production platforms. Hence a general vertical two-phase flow pattern is required that can be applied to all flow situations (Hasan, 1995). Reliable models for downward multiphase flow are currently unavailable and the design codes are deficient in this area.

3.3.1.3 Inclined Flow Regimes

The effect of pipeline inclination on the gas–liquid two-phase flow regimes is of a major interest in hilly terrain pipelines that consist almost entirely of uphill and downhill inclined sections. Pipe inclination angles have a very strong influence on flow pattern transitions. Generally, the flow regime in a near-horizontal pipe remains segregated for downward inclinations and changes to an intermittent flow regime for upward inclinations. An intermittent flow regime remains intermittent when tilted upward and tends to segregated flow pattern when inclined downward. The inclination should not significantly affect the distributed flow regime (Shoham, 1982; Scott *et al.*, 1987).

3.3.1.4 Flow Pattern Maps

The boundaries of different gas–liquid two-phase flow regimes have been determined experimentally and reported in the literature. The results of experimental studies are generally presented as a flow pattern map. The respective pattern may be represented as areas on a plot, the coordinates of which are the dimensional variables (i.e., superficial phase velocities) or dimensionless parameters containing these velocities (Collier and Thome, 1996). Although plots are useful in representing data, they are limited to the particular sets of conditions investigated and there is an obvious need to generalize flow pattern information so that it can be applied to any pair of fluids and any geometry. A more flexible method, which overcomes this difficulty, is to examine each transition individually and derive a criterion valid for that particular transition.

For horizontal flows, the classical map is that of Baker (1954), which is widely used in the petroleum industry. The original Baker diagram is shown in Figure 3-4. The map is based on air–water data at atmospheric pressure in 1-, 2-, and 4-inch pipes. The abscissa is the superficial mass velocity of the liquid phase (G_L) and the ordinate is the superficial mass velocity of the gas phase (G_G) and both coordinates have been corrected for physical properties of the respective phases. These parameters are

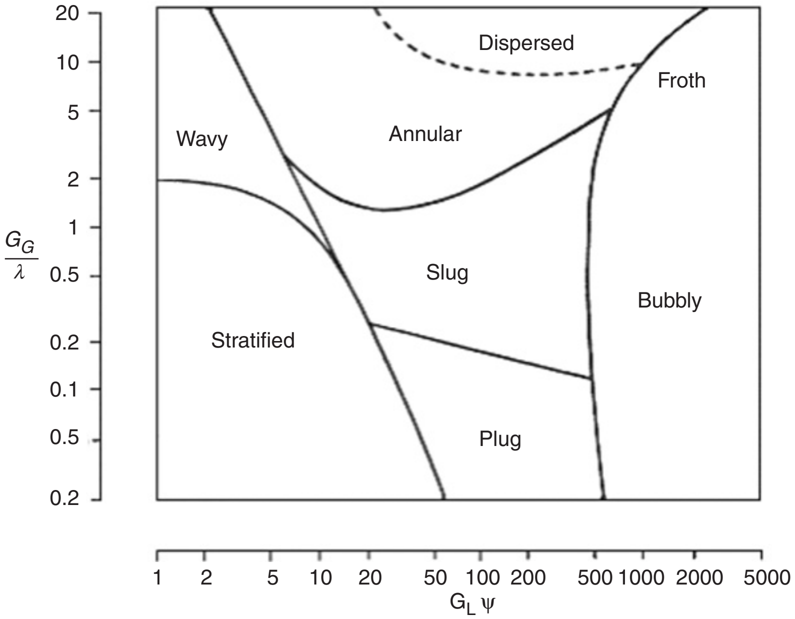


Figure 3-4. Flow pattern map for horizontal flow (Baker, 1954).

given by the following relationships:

$$G = \frac{\dot{m}}{A} \tag{3-18}$$

$$\lambda = \left[\left(\frac{\rho_G}{\rho_A}\right)\left(\frac{\rho_L}{\rho_W}\right)\right]^{0.5} \tag{3-19}$$

$$\Psi = \left(\frac{\sigma_W}{\sigma_L}\right)\left[\left(\frac{\mu_L}{\mu_W}\right)\left(\frac{\rho_W}{\rho_L}\right)^2\right]^{0.33} \tag{3-20}$$

where $\dot{m}$ is mass flow rate, lbm/hr; A is pipe cross section, ft^2; σ is surface tension, dynes/ft; and μ is viscosity, lb/ft.s. The subscripts A and W refer to the values of the physical properties for air and water, respectively, at atmospheric pressure and temperature.

For vertical gas–liquid two-phase flow, the flow pattern map of Hewitt and Roberts (1969) is shown in Figure 3-5. This map has been obtained

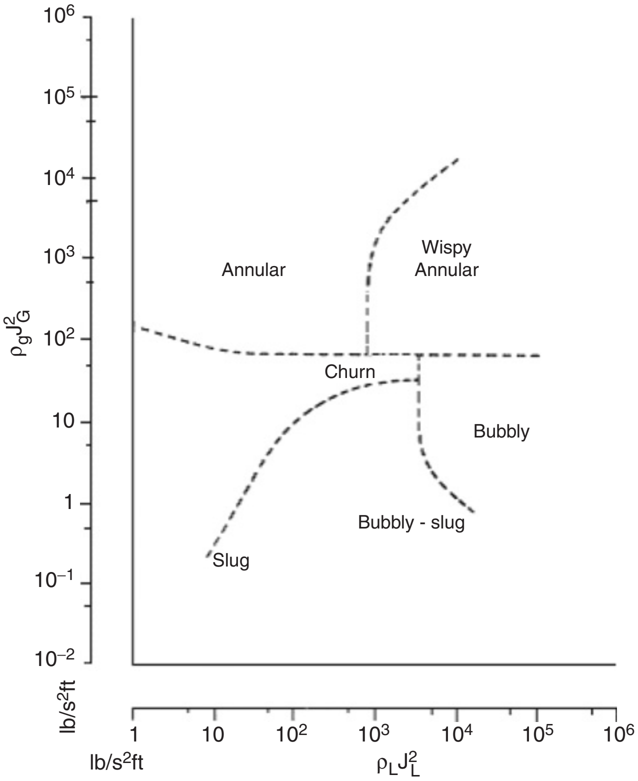


Figure 3-5. Flow pattern map for vertical flow (Hewitt and Roberts, 1969).

from observations on low-pressure air-water and high-pressure steam–water flow in small diameter (0.4–1.2 inch) vertical tubes. The axes are the superficial momentum fluxes of the liquid ($\rho_L \times J_L^2$) and vapor ($\rho_G \times J_G^2$) phases, respectively. These parameters can also be defined in terms of mass velocity (G) and the vapor quality (x) as follows:

$$\rho_L \times J_L^2 = \frac{[G(1-x)]^2}{\rho_L} \tag{3-21}$$

$$\rho_G \times J_G^2 = \frac{[Gx]^2}{\rho_G} \tag{3-22}$$

Note that a word of warning should be issued about the use of flow pattern maps. For example, the use of superficial phase velocities for the axes of the map restricts its application to one particular situation. Therefore, they should be used with discretion serving as a general guide to the likely pattern rather than a positive indication that the pattern actually exists for a given situation. There have been some attempts to evaluate the basic mechanisms of flow pattern transitions and thus to provide a mechanistic flow pattern map for estimating their occurrence (Taitel and Dukler, 1976; Barnea, 1987; Taitel *et al*., 1980, Taitel, 1990; Petalas and Aziz, 2000). In these transition models, the effects of system parameters are incorporated; hence they can be applied over a range of conditions. However, most of them are somewhat complex and require the use of a predetermined sequence to determine the dominant flow pattern (Hasan and Kabir, 2002).

3.3.2 Three-Phase Flow Regimes

The main difference between two-phase (gas/liquid) flows and three-phase (gas/liquid/liquid) flows is the behavior of the liquid phases, where in three-phase systems the presence of two liquids gives rise to a rich variety of flow patterns (Hall, 1997). Basically, depending on the local conditions, the liquid phases appear in a separated or dispersed form (Brauner and Maron, 1992; Brauner, 2001). In the case of separated flow, distinct layers of oil and water can be discerned, although there may be some interentrainment of one liquid phase into the other. In dispersed flow, one liquid phase is completely dispersed as droplets in the other, resulting in two possible situations, namely an oil continuous phase and a water continuous phase (Chen and Guo, 1999). The transition from one liquid continuous phase to the other is known as phase inversion. If the liquid phases are interdispersed, then prediction of phase inversion is an important item. For this purpose, Decarre and Fabre (1997) developed a phase inversion model that can be used to determine which of two liquids is continuous.

Due to the many possible transport properties of three-phase fluid mixtures, the quantification of three-phase flow pattern boundaries is a difficult and challenging task. Acikgoz *et al*. (1992) observed a very complex array of flow patterns and described 10 different flow regimes. In their work, the pipe diameter was only 0.748 inch and stratification was seldom achieved. In contrast, the Lee *et al*. (1993) experiments were carried out in a 4-inch-diameter pipe. They observed and classified seven flow patterns, which were similar to those of two-phase flows: (1) smooth stratified, (2) wavy stratified, (3) rolling wave, (4) plug flow, (5) slug flow, (6) pseudo slug, and (7) annular flow. The first three flow patterns can be classified as stratified flow regime and they noted that the oil and the water are generally segregated, with water flowing as a liquid layer at the bottom of the pipe and oil flowing on top. Even for plug flow, the water remained at the bottom because agitation of the liquids was not sufficient to mix the oil and water phases. Note that turbulence, which naturally exists in a pipeline, can be sufficient to provide adequate mixing of the water and oil phase. However, the minimum natural turbulent energy for adequate mixing depends on the oil and water flow rates, pipe diameter and inclination, water concentration, viscosity, density, and interfacial tension. Dahl *et al*. (2001) provided more detailed information on the prediction methods that can be used to determine whether a water-in-oil mixture in the pipe is homogeneous or not.

3.4 CALCULATING MULTIPHASE FLOW PRESSURE GRADIENTS

The hydraulic design of a multiphase flow pipeline is a two-step process. The first step is the determination of the multiphase flow regimes because many pressure drop calculation methods rely on the type of flow regime present in the pipe. The second step is the calculation of flow parameters, such as pressure drop and liquid holdup, to size pipelines and field processing equipment, such as slug catchers.

3.4.1 Steady-State Two-Phase Flow

The techniques used most commonly in the design of a two-phase flow pipeline can be classified into three categories: single-phase flow approaches, homogeneous flow approaches, and mechanistic models. Within each of these groups are subcategories that are based on the general characteristics of the models used to perform design calculations.

3.4.1.1 Single-Phase Flow Approaches

In this method, the two-phase flow is assumed to be a single-phase flow having pseudo-properties arrived at by suitably weighting the properties of the individual phases. These approaches basically rely solely on the well-established design equations for single-phase gas flow in pipes. Two-phase flow is treated as a simple extension by use of a multiplier: a safety factor to account for the higher pressure drop generally encountered in two-phase flow. This heuristic approach was widely used and generally resulted in inaccurate pipeline design (Dukler, 1969; Oranje, 1973; Schewikert, 1986).

In the past, single-phase flow approaches were used commonly for the design of wet gas pipelines. When the amount of the condensed liquid is negligibly small, the use of such methods could at best prevent underdesign, but more often than not, the quantity of the condensed liquid is significant enough that the single-phase flow approach grossly overpredicted the pressure drop (Ullah, 1987). Hence, this method is not covered in depth here but can be reviewed elsewhere (Uhl, 1965).

3.4.1.2 Homogeneous Flow Approaches

The inadequacy of the single-phase flow approaches spurred researchers to develop better design procedures and predictive models for two-phase flow systems. This effort led to the development of homogeneous flow approaches to describe these rather complex flows. The homogeneous approach, also known as the friction factor model, is similar to that of the single-phase flow approach except that mixture fluid properties are used in determination of the friction factor. Therefore, the appropriate definitions of the fluid properties are critical to the accuracy of the model. The mixture properties are expressed empirically as a function of the gas and liquid properties, as well as their respective holdups. Many of these correlations are based on flow regime correlations that determine the two-phase (gas/liquid) flow friction factor, which is then used to estimate pressure drop. While some of the correlations predict pressure drop reasonably well, their range of applicability is generally limited, making their use as a scale up tool marginal. This limitation is understandable because the database used in developing these correlations is usually limited and based on laboratory-scale experiments. Extrapolation of these data sets to larger lines and hydrocarbon systems is questionable at best (Cindric *et al.*, 1987; Mokhatab, 2002a,b,c; Mokhatab and Bonizzi, 2006). Blind application of these correlations can lead to results, which are marginal (Holt *et al.*, 1995), resulting in two-phase equipment overdesign leading to unnecessary costs or even failure (Holt *et al.*, 1999). However, many of these correlations have been used for lack of other design means and two of which are outlined in the following section. In-depth comparative analyses of the available homogeneous flow approaches have been reported by Brill and Beggs (1991) and Collier and Thome (1996).

3.4.1.2.1 Lockhart and Martinelli Method This method was developed by correlating experimental data generated in horizontal isothermal two-phase flow of two-component systems (air–oil and air–water) at low pressures (close to atmospheric) in a 1-inch-diameter pipe. Lockhart and Martinelli (1949) separated data into four sets dependent on whether the phases flowed in laminar or turbulent flow, if the phases were flowing alone in the same pipe. In this method, a definite portion of the flow area is assigned to each phase and it is assumed that the single-phase pressure drop equations can be used independently for each phase. The two-phase frictional pressure drop is calculated by multiplying by a correction factor for each phase, as follows:

$$\frac{dP}{dx} = \varphi_G^2 \left(\frac{dP}{dx}\right)_G = \varphi_L^2 \left(\frac{dP}{dx}\right)_L \tag{3-23}$$

where

$$\left(\frac{dP}{dx}\right)_G = \left(\frac{f_G \rho_G V_{SG}^2}{2 g_C D}\right) \tag{3-23-1}$$

$$\left(\frac{dP}{dx}\right)_L = \left(\frac{f_L \rho_L V_{SL}^2}{2 g_C D}\right) \tag{3-23-2}$$

The friction factors f_G and f_L are determined from the Moody (1944) diagram using the following values of the Reynolds number:

$$N_{Re,G} = \frac{\rho_G V_{SG} D}{\mu_G} \tag{3-24}$$

$$N_{Re,L} = \frac{\rho_L V_{SL} D}{\mu_L} \tag{3-25}$$

The two-phase flow correction factors (φ_G, φ_L) are determined from the relationships of Equations (3-26) and (3-27) (Lockhart and

Martinelli, 1949; Chisholm and Sutherland, 1969):

$$\varphi_{\mathrm{G}}^{2} = 1 + CX + CX^{2} \tag{3-26}$$

$$\varphi_{\mathrm{L}}^{2} = 1 + \frac{C}{X} + \frac{1}{X^{2}} \tag{3-27}$$

where

$$X = \left[\frac{\left(\frac{dP}{dx}\right)_{\mathrm{L}}}{\left(\frac{dP}{dx}\right)_{\mathrm{G}}} \right]^{0.5} \tag{3-28}$$

where parameter C has the values shown in Table 3-1. Note that the laminar flow regime for a phase occurs when the Reynolds number for that phase is less than 1000.

In this method, the correlation between liquid holdup and Martinelli parameter, X, is independent of the flow regime and can be expressed as follows (Collier and Thome, 1996):

$$H_{\mathrm{L}}^{-2} = 1 + \frac{20}{X} + \frac{1}{X^{2}} \tag{3-29}$$

The fluid acceleration pressure drop was ignored in this method. However, the extension of the work covering the estimation of the accelerative term was done by Martinelli and Nelson (1948). Although different modifications of this method have been proposed, the original method is believed to be generally the most reliable (Collier and Thome, 1996).

Table 3-1
'*C*' Parameter

Liquid phase	Gas phase	*C*
Turbulent	Turbulent	20
Laminar	Turbulent	12
Turbulent	Laminar	10
Laminar	Laminar	5

3.4.1.2.2 Beggs and Brill Method This method was developed from 584 experimental data sets generated on a laboratory-scale test facility using an air–water system. The facility consisted of 90 feet of 1- or 1.5-inch-diameter acrylic (smooth) pipe, which could be inclined at any angle (Beggs and Brill, 1973). The pipe angle was varied between horizontal to vertical and the liquid holdup and the pressure were measured. For each pipe size in the horizontal position, flow rates of two phases were varied to achieve all flow regimes. Beggs and Brill (1973) developed correlations for the liquid holdup for each of three horizontal flow regimes and then corrected these for the pipe inclination/angle.

The following parameters are used for determination of all horizontal flow regimes (Brill and Beggs, 1991):

$$N_{\mathrm{Fr}} = \frac{V_{\mathrm{M}}^2}{gD} \tag{3-30}$$

$$L_1 = 316\,\lambda_{\mathrm{L}}^{0.302} \tag{3-31}$$

$$L_2 = 0.0009252\,\lambda_{\mathrm{L}}^{-2.4684} \tag{3-32}$$

$$L_3 = 0.10\,\lambda_{\mathrm{L}}^{-1.4516} \tag{3-33}$$

$$L_4 = 0.5\,\lambda_{\mathrm{L}}^{-6.738} \tag{3-34}$$

The flow regimes limits are

Segregated

$$N_{\mathrm{Fr}} < L_1 \quad \text{for} \quad \lambda_{\mathrm{L}} < 0.01$$

$$N_{\mathrm{Fr}} < L_2 \quad \text{for} \quad \lambda_{\mathrm{L}} \geq 0.01$$

Transition

$$L_2 \leq N_{\mathrm{Fr}} \leq L_3 \quad \text{for} \quad \lambda_{\mathrm{L}} \geq 0.01$$

Intermittent

$$L_3 \leq N_{\mathrm{Fr}} \leq L_1 \quad \text{for} \quad 0.01 < \lambda_{\mathrm{L}} \leq 0.4$$

$$L_3 < N_{\mathrm{Fr}} \leq L_4 \quad \text{for} \quad \lambda_{\mathrm{L}} \geq 0.4$$

Distributed

$$N_{\mathrm{Fr}} \geq L_1 \quad \text{for} \quad \lambda_{\mathrm{L}} < 0.4$$

$$N_{\mathrm{Fr}} > L_4 \quad \text{for} \quad \lambda_{\mathrm{L}} \geq 0.4$$

Also, the horizontal liquid holdup, $H_{\mathrm{L}}(0)$, is calculated using the following equation:

$$H_{\mathrm{L}}(0) = \frac{\mathrm{a}\lambda_{\mathrm{L}}^{\mathrm{b}}}{N_{\mathrm{Fr}}^{\mathrm{c}}} \tag{3-35}$$

where parameters a, b, and c are determined for each flow regime and are given in Table 3-2. With the constraint that $H_{\mathrm{L}}(0) \geq \lambda_{\mathrm{L}}$.

When the flow is in the transition region, the liquid holdup is calculated by interpolating between the segregated and the intermittent flow regimes as follows:

$$H_{\mathrm{L}}(0)_{\mathrm{transition}} = AH_{\mathrm{L}}(0)_{\mathrm{segregated}} + (1 - A)H_{\mathrm{L}}(0)_{\mathrm{intermittent}} \tag{3-36}$$

where

$$A = (L_3 - N_{\mathrm{Fr}})/(L_3 - L_2) \tag{3-37}$$

The amount of liquid holdup in an inclined pipe, $H_{\mathrm{L}}(\theta)$, is determined by multiplying an inclination factor (Ψ) by the calculated liquid holdup for the horizontal conditions:

$$H_{\mathrm{L}}(\theta) = H_{\mathrm{L}}(0) \times \Psi \tag{3-38}$$

Table 3-2
a, b, c Parameters

Flow regime	a	b	c
Segregated	0.98	0.4846	0.0868
Intermittent	0.845	0.5351	0.0173
Distributed	1.065	0.5824	0.0609

where

$$\Psi = 1 + \alpha[\sin(1.8\theta) - 0.333 \sin^3(1.8\theta)] \tag{3-38-1}$$

where θ is the pipe angle and α is calculated as follows:

$$\alpha = (1 - \lambda_\mathrm{L})\,\mathrm{Ln}[d\lambda_\mathrm{L}^\mathrm{e} N_\mathrm{LV}^\mathrm{f} N_\mathrm{Fr}^\mathrm{g}] \tag{3-38-2}$$

where

$$N_\mathrm{LV} = V_\mathrm{SL}\left(\frac{\rho_\mathrm{L}}{g\sigma}\right)^{0.25} \tag{3-38-3}$$

The equation parameters are determined from each flow regime using the numbers from Table 3-3.

The two-phase pressure gradient due to pipeline elevation can be determined as follows:

$$\left(\frac{dP}{dx}\right)_\mathrm{ele} = \frac{g}{g_\mathrm{C}}\{\rho_\mathrm{L} H_\mathrm{L}(\theta) + \rho_\mathrm{G}[1 - H_\mathrm{L}(\theta)]\} \tag{3-39}$$

Also, the two-phase frictional pressure drop is calculated as

$$\left(\frac{dP}{dx}\right)_\mathrm{fri} = \frac{f_\mathrm{tp}\rho_\mathrm{n} V_\mathrm{M}^2}{2g_\mathrm{c}D} \tag{3-40}$$

where

$$f_\mathrm{tp} = f_\mathrm{n}\exp(\beta) \tag{3-40-1}$$

Table 3-3
d, e, f, and g Parameters

Flow regime	d	e	f	g
Segregated uphill	0.011	−3.768	3.539	−1.614
Intermittent uphill	2.96	0.305	−0.4473	0.0978
Distributed uphill		$\alpha = 0$, $\Psi = 1$		
All flow regimes downhill	4.7	−0.3692	0.1244	−0.5056

where f_n is the no-slip friction factor determined from the smooth pipe curve of the Moody diagram using no-slip viscosity and the density in calculating the two-phase flow Reynolds number. In other words,

$$f_\mathrm{n} = \frac{1}{\left[2\log\left(\dfrac{N_\mathrm{Re,n}}{4.5223\log\ N_\mathrm{Re,n} - 3.8215}\right)\right]^2} \tag{3-40-1-1}$$

where

$$N_\mathrm{Re,n} = \frac{\rho_\mathrm{n} V_\mathrm{M} D}{\mu_\mathrm{n}} = \frac{[(\rho_\mathrm{L}\lambda_\mathrm{L} + \rho_\mathrm{G}\,(1-\lambda_\mathrm{L}))\,V_\mathrm{M} D]}{[\mu_\mathrm{L}\lambda_\mathrm{L} + \mu_\mathrm{G}\,(1-\lambda_\mathrm{L})]} \tag{3-40-1-2}$$

The exponent β is given by

$$\beta = [\mathrm{LnY}]/\{-0.0523 + 3.182\,\mathrm{LnY} - 0.8725(\mathrm{LnY})^2 + 0.01853(\mathrm{LnY})^4\} \tag{3-40-2}$$

where

$$Y = \frac{\lambda_\mathrm{L}}{[H_\mathrm{L}(\theta)]^2} \tag{3-40-3}$$

The pressure drop due to acceleration is only significant in gas transmission pipelines at high gas flow rates. However, it can be included for completeness as

$$\left(\frac{dP}{dx}\right)_\mathrm{acc} = \left[\frac{\rho_\mathrm{S} V_\mathrm{M} V_\mathrm{SG}}{g_\mathrm{c} P}\right]\left(\frac{dP}{dx}\right)_\mathrm{tot} \tag{3-41}$$

The Beggs and Brill (1973) method can be used for horizontal and vertical pipelines, although its wide acceptance is mainly due to its usefulness for inclined pipe pressure drop calculation (Brill and Beggs, 1991). However, the proposed approach has limited applications dictated by the database on which the correlations are derived.

Note that the Beggs and Brill (1973) correlation is not usually employed for the design of wet gas pipelines. Specifically the holdup characteristic (holdup versus flow rate) is not well predicted by this method. This makes design difficult because one is unable to reliably quantify the retention of liquid in the line during turndown conditions.

3.4.1.3 Mechanistic Models

While it is indeed remarkable that some of the present correlations can adequately handle noncondensing two-phase flow, they give erroneous results for gas-condensate flow in large-diameter lines at high pressures (Osman and El-Feky, 1985; Battara *et al.*, 1985; Mokhatab, 2002a,b,c; Mokhatab and Bonizzi, 2006). This shortcoming of existing methods has led to the development of mechanistic models based on fundamental laws and thus can offer more accurate modeling of the pipe geometric and fluid property variations. All of these models predict a stable flow pattern under the specified conditions and then use momentum balance equations to calculate liquid holdup, pressure drop, and other two-phase flow parameters with a greater degree of confidence than that possible by purely empirical correlations (Hasan and Kabir, 1986, 1992, 1999; Collier and Thome, 1996; Holt *et al.*, 1999). The mechanistic models presented in the literature are either incomplete in that they only consider flow pattern determination or are limited in their applicability to only some pipe inclinations or small-diameter, low-pressure two-phase flow lines (Taitel and Dukler, 1976; Barnea, 1987; Xiao *et al.*, 1990; Ansari *et al.*, 1994, Taitel *et al.*, 1995; Wilkens, 1997). However, new mechanistic models presented by Petalas and Aziz (2000) and Zhang et al. (2003) have proven to be more robust than previous models, although further investigations and testing of these models are needed with high-quality field and laboratory data in larger diameter, high-pressure systems. Readers are referred to the original references for a detailed treatment of these models.

3.4.2 Steady-State Three-Phase Flow

Compared to numerous investigations of two-phase flow in the literature, there are only limited works on three-phase flow of gas–liquid–liquid mixtures. In fact, the complex nature of such flows makes prediction very difficult. In an early study, Tek (1961) treated the two immiscible liquid phases as a single fluid with mixture properties, thus a two-phase flow correlation could be used for pressure loss calculations. Studies by Pan (1996) have shown that the classical two-phase flow correlations for gas–liquid two-phase flow can be used as the basis in the determination of three-phase flow parameters; however, the generality of such empirical approaches is obviously questionable. Hence, an appropriate model is required to describe the flow of one gas and two liquid phases.

One of the most fundamental approaches used to model such systems is the two-fluid model, where the presented approach can be used by combining the two liquid phases as one pseudo-liquid phase and modeling the three-phase flow as a two-phase flow. However, for more accurate results, three-fluid models should be used to account for the effect of liquid–liquid interactions on flow characteristics, especially at low flow rates. Several three fluid models were found in the literature. All of these models are developed from the three-phase momentum equations with few changes from one model to the other. The most obvious model has been developed by Barnea and Taitel (1996); however, such a model introduces much additional complexity and demands much more in computer resources compared with the two-fluid model for two-phase flow (Bonizzi and Issa, 2003).

3.4.3 Transient Multiphase Flow

Transient multiphase flow in pipelines can occur due to changes in inlet flow rates, outlet pressure, opening or closing of valves, blowdown, ramp-up, and pigging. In each of these cases, detailed information of the flow behavior is necessary for the designer and the operator of the system to construct and operate the pipeline economically and safely. The steady-state pipeline design tools are not sufficient to adequately design and confirm the operational flexibility of multiphase pipelines. Therefore, a model for predicting the overall flow behavior in terms of pressure, liquid holdup, and flow rate distributions for these different transient conditions would be very useful. Some current research challenges in modeling transient flow relate to an understanding and formulation of basic flow models for oil–water–gas flow and to numerical methods applicable for the solution of transient multiphase flows. These methods are available to some extent in multiphase fluid dynamics simulation codes, which are the key design tools for multiphase flow pipeline calculations.

Transient multiphase flow is traditionally modeled by one-dimensional averaged conservation laws, yielding a set of partial differential equations. In this section, two models of particular industrial interest are described.

- The two-fluid model (TFM), consisting of a separate momentum equation for each phase.
- The drift-flux model (DFM), consisting of a momentum equation and an algebraic slip relation for the phase velocities.

The TFM is structurally simpler, but involves an extra differential equation when compared to the DFM. They do yield somewhat different transient results, although the differences are often small (Masella *et al.*, 1998).

The following major assumptions have been made in the formulation of the differential equations.

1. Two immiscible liquid phases (oil and water) that are assumed to be a single fluid with mixture properties.
2. Flow is one dimensional in the axial direction of the pipeline.
3. Flow temperature is constant at wall, and no mass transfer occurs between gas and liquid phases. Note that most commercial codes allow phase change.
4. The physical properties of multiphase flow are determined at the average temperature and pressure of flow in each segment of the pipeline.

3.4.3.1 Two-Fluid Model

The TFM is governed by a set of four partial differential equations, the first two of which express mass conservation for gas and liquid phases, respectively,

$$\frac{\partial}{\partial t}[\rho_G H_G] + \frac{\partial}{\partial x}[\rho_G H_G V_G] = 0 \qquad (3\text{-}42)$$

$$\frac{\partial}{\partial t}[\rho_L H_L] + \frac{\partial}{\partial x}[\rho_L H_L V_L] = 0 \qquad (3\text{-}43)$$

The last two equations represent momentum balance for the gas and liquid phases, respectively,

$$\frac{\partial}{\partial t}[\rho_G H_G V_G] + \frac{\partial}{\partial x}\left[\rho_G H_G V_G^2 + H_G \Delta P_G\right] + H_G \frac{\partial P}{\partial x} = \tau_G + \tau_i - \rho_G g \sin\theta \qquad (3\text{-}44)$$

$$\frac{\partial}{\partial t}[\rho_L H_L V_L] + \frac{\partial}{\partial x}\left[\rho_L H_L V_L^2 + H_L \Delta P_L\right] + H_L \frac{\partial P}{\partial x} = \tau_L - \tau_i - \rho_L g \sin\theta \qquad (3\text{-}45)$$

In Equations (3-44) and (3-45), parameter P denotes the interface pressure, whereas V_k, ρ_k, and H_k are the velocity, the density, and the volume fraction of phase $k \in \{G, L\}$, respectively. The variables τ_i and τ_k are the interfacial and wall momentum transfer terms. The quantities ΔP_G and ΔP_L correspond to the static head around the interface (De Henau and Raithby, 1995), defined as follows:

$$\Delta P_G = P_G - P = -\rho_G \left[\frac{1}{2} \cos\left(\frac{\omega}{2}\right) + \frac{1}{3\pi H_G} \sin^3\left(\frac{\omega}{2}\right) \right] gD \cos\theta \tag{3-46}$$

$$\Delta P_L = P_L - P = -\rho_L \left[\frac{1}{2} \cos\left(\frac{\omega}{2}\right) + \frac{1}{3\pi H_L} \sin^3\left(\frac{\omega}{2}\right) \right] gD \cos\theta \tag{3-47}$$

where ω is the wetted angle.

The detailed description of the solution algorithm for these equations is based on a finite volume method (Masella *et al.*, 1998).

One major limitation of this type of model is the treatment of the interfacial coupling. While this is relatively easy for separated flows (stratified and annular), this treatment is intrinsically flawed for intermittent flows. Another drawback is that propagation phenomena, especially pressure waves, tend not to account for satisfactorily (King, 1998).

3.4.3.2 Drift Flux Model

The DFM is derived from the two-fluid model by neglecting the static head terms ΔP_G and ΔP_L in Equations (3-46) and (3-47) and replacing the two momentum equations by their sum. The main advantages of this three-equation model are as follows.

- The equations are in conservative form, which makes their solution by finite volume methods less onerous.
- The interfacial shear term, τ_i, is cancelled out in the momentum equations, although it appears in an additional algebraic relation called the slip law.
- The model is well posed and does not exhibit a complex characteristic.

Adding Equations (3-44) and (3-45) together yields

$$\frac{\partial}{\partial t}\left[\rho_{G}H_{G}V_{G}+\rho_{L}H_{L}V_{L}\right]+\frac{\partial}{\partial x}\left[\rho_{G}H_{G}V_{G}^{2}+\rho_{L}H_{L}V_{L}^{2}+P\right] = \tau_{G}+\tau_{L}-(\rho_{G}H_{G}+\rho_{L}H_{L})\sin\theta \qquad (3\text{-}48)$$

The interfacial exchange term, τ_i, is no longer present in the aforementioned equation. This leads in the DFM to a new model that consists of three partial differential equations, i.e., Equations (3-42), (3-43), and (3-48). Additional model numerical solution details can be found in Faille and Heintze (1996).

Note that the drift flux approach is best applied to closely coupled flows such as bubbly flow. Its application to stratified flows is, at best, artificial (Banerjee, 1986).

3.5 MULTIPHASE FLOW IN GAS/CONDENSATE PIPELINES

Gas/condensate flow is a multiphase flow phenomenon commonly encountered in raw gas transportation. However, the multiphase flow that takes place in gascondensate transmission lines differs in certain respects from the general multiphase flow in pipelines. In fact, in gas/condensate flow systems, there is always interphase mass transfer from the gas phase to the liquid phase because of the temperature and pressure variations, which leads to compositional changes and associated fluid property changes (Ayala and Adewumi, 2003). In addition, the amount of liquid in such systems is assumed to be small, and the gas flow rate gives a sufficiently high Reynolds number that the fluid flow regime for a nearly horizontal pipe can be expected to be annular-mist flow and/or stratified flow (Boriyantoro and Adewumi, 1994; Asante, 2002). For other inclined cases, even with small quantities of liquids, slug-type regimes may be developed if liquids start accumulating at the pipe lower section.

In order to achieve optimal design of gas/condensate pipelines and downstream processing facilities, one needs a description of the relative amount of condensate and the flow regime taking place along the pipelines, where fluid flowing in pipelines may traverse the fluid phase envelope such that the fluid phase changes from single phase to two phase or vice versa. Hence, compositional singlephase/multiphase hydrodynamic modeling, which couples the hydrodynamic model with the natural gas-phase behavior model, is necessary to predict fluid dynamic behavior in gas/condensate transmission lines. The hydrodynamic model is required to obtain flow parameters along the pipeline, and the phase behavior model is required for determining the phase condition at any point in the pipe, the mass transfer between the flowing phases, and the fluid properties.

Despite the importance of gas flow with low liquid loading for the operation of gas pipelines, few attempts have been made to study flow parameters in gas/condensate transmission lines. While the single-phase flow approaches have been applied previously to gas/condensate systems, only a few attempts have been reported for the use of the two-fluid model for this purpose. Some of them have attempted to make basic assumptions (e.g., no mass transfer between gas and liquid phases) in their formulation and others simply assume one flow regime for the entire pipe length. However, a compositional hydrodynamic model that describes the steady-state behavior of multiphase flow in gas/condensate pipelines has been presented by Ayala and Adewumi (2003). The model couples a phase behavior model, based on the Peng and Robinson (1976) equation of state, and a hydrodynamic model, based on the two-fluid model. The proposed model is a numerical approach, which can be used as an appropriate tool for engineering design of multiphase pipelines transporting gas and condensate. However, the complexity of this model precludes further discussion here.

Note that the presence of liquid (condensates), in addition to reducing deliverability, creates several operational problems in gascondensate transmission lines. Periodic removal of the liquid from the pipeline is thus desirable. To remove liquid accumulation in the lower portions of pipeline, pigging operations are performed. These operations keep the pipeline free of liquid, reducing the overall pressure drop increasing pipeline flow efficiency. However, the pigging process associates with transient flow behavior in the pipeline. Thus, it is imperative to have a means of predicting transient behavior encountered in multiphase, gas/gas-condensate pipelines. Until recently, most available commercial codes were based on the two-fluid model; however, the model needs many modifications to be suitable for simulating multiphase transient flow in gas/gas-condensate transmission lines. For example, the liquid and gas continuity equations need to be modified to account for the mass transfer between phases. So far, several codes have been reported for this purpose, where three main commercial transient codes are OLGA (Bendiksen *et al.*, 1987, 1991), PROFES (Black *et al.*, 1990), and TACITE (Pauchon *et al.*, 1993, 1994). Detailed discussion of these codes is beyond the scope of this book; readers are referred to the original papers for further information.

3.6 TEMPERATURE PROFILE OF MULTIPHASE PIPELINES

Predicting the flow temperature and pressure changes has become increasingly important for use in both the design and the operation of flow transmission pipelines. It is therefore imperative to develop appropriate methods capable of predicting these parameters for multiphase pipelines (Mokhatab, 2006a). A simplified flowchart of a suitable computing algorithm is shown in Figure 3-6. This algorithm calculates pressure and temperature along the pipeline by iteratively converging on pressure and temperature for each sequential "segment" of the pipeline. The algorithm converges on temperature in the outer loop and pressure in the inner loop because robustness and computational speed are obtained when converging on the least sensitive variable first (Brill and Beggs, 1991).

The pipeline segment length should be chosen such that the fluid properties do not change significantly in the segment. More segments are recommended for accurate calculations for a system where fluid properties can change drastically over short distances. Often, best results are obtained when separate segments (with the maximum segment length less than about 10% of total line length) are used for up, down, and horizontal segments of the pipeline (Brill and Beggs, 1991).

Prediction of the pipeline temperature profile can be accomplished by coupling the pressure gradient and enthalpy gradient equations as follows (Brill and Beggs, 1991).

$$\Delta H = \frac{-V_{\mathrm{M}} V_{\mathrm{SG}} \Delta P}{(778)(32.17)(\overline{P})} + \frac{\Delta Z}{778} - \frac{U \pi D(\overline{T} - T_{\mathrm{a}}) \Delta L}{3600(\dot{M}_{\mathrm{M}})} \tag{3-49}$$

where ΔH is enthalpy change in the calculation segment, Btu/$\mathrm{lb_m}$; V_{M} is velocity of the fluid, ft/sec; V_{SG} is superficial gas velocity, ft/sec; ΔP is estimated change in pressure, psi; $\overline{P}$ is average pressure in calculation segment, psia; ΔZ is change in elevation, ft; U is overall heat transfer coefficient, Btu/hr-ft^2-°F; D is reference diameter on which U is based, ft; T is estimated average temperature in calculation segment, °F; T_{a} is ambient temperature, °F; ΔL is change in segment length, ft; and $\dot{M}_{\mathrm{M}}$ is gas–liquid mixture mass flow rate, $\mathrm{lb_m}$/s.

As can be seen from Equation (3-49), temperature and pressure are mutually dependent variables so that generating a very precise temperature profile requires numerous iterative calculations. The temperature

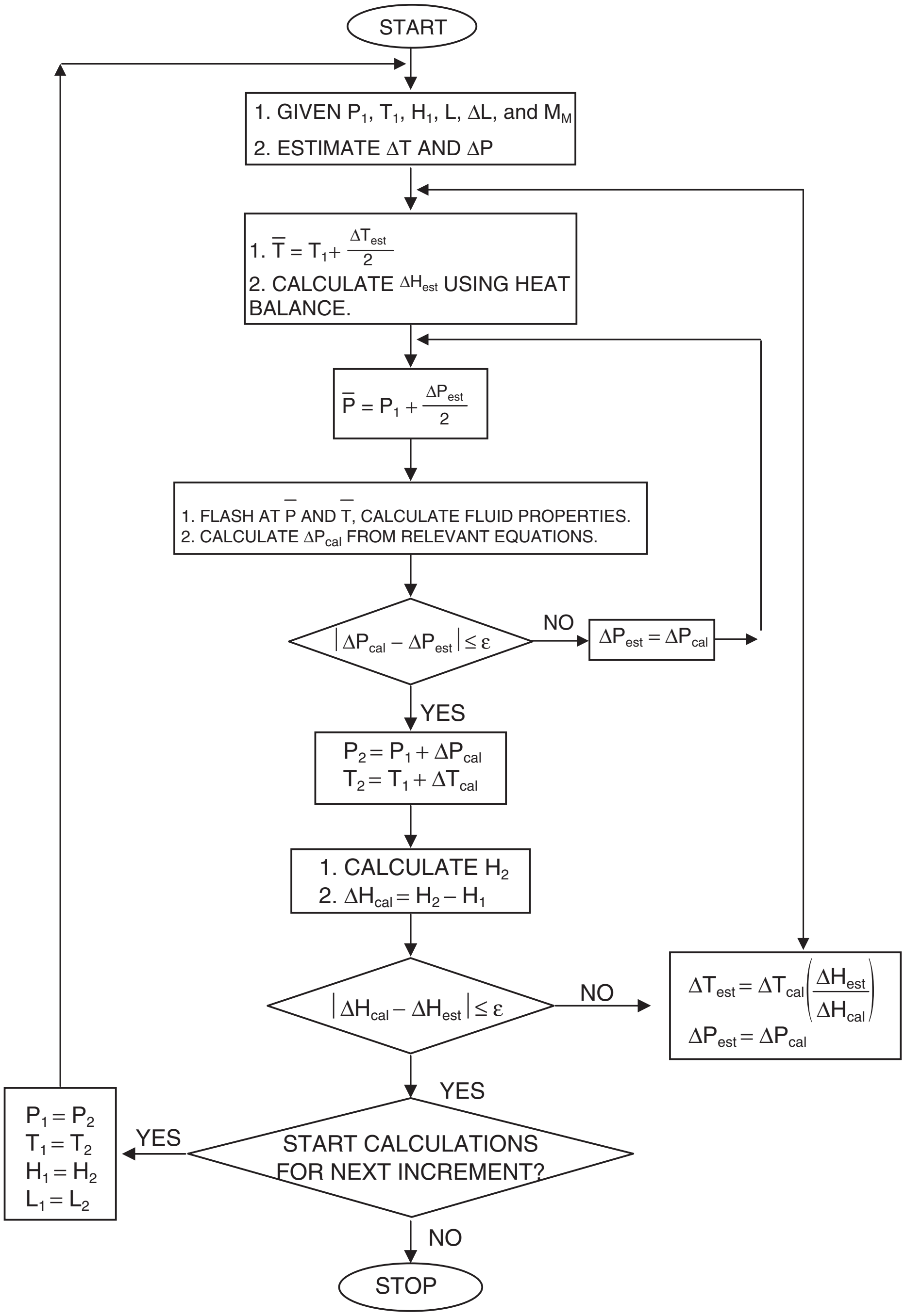


Figure 3-6. Pressure and temperature calculation procedure (Brill and Beggs, 1991).

and pressure of each pipe segment are calculated using a double-nested procedure in which for every downstream pressure iteration, convergence is obtained for the downstream temperature by property values evaluated at the average temperature and pressure of that section of pipe. Experience shows that it is essential to use a good pressure-drop model to assess the predominant parameter, i.e., pressure.

The overall heat transfer coefficient (U) can be determined by a combination of several coefficients, which depend on the method of heat transfer and pipe configuration. Figure 3-7 is a cross section of a pipe, including each "layer" through which heat must pass to be transferred from the fluid to the surroundings or vice versa. This series of layers has an overall resistance to heat transfer made up of the resistance of each layer.

In general, the overall heat transfer coefficient for a pipeline is the reciprocal of the sums of the individual resistances to heat transfer, where each resistance definition is given in Table 3-4. The individual resistances are calculated from the given equations in Table 3-5.

Since fluid properties are key inputs into calculations such as pressure drop and heat transfer, the overall simulation accuracy depends on accurate property predictions of the flowing phases. Most of the required physical and thermodynamic properties of the fluids are derived from the equation of state. However, empirical correlations are used for the

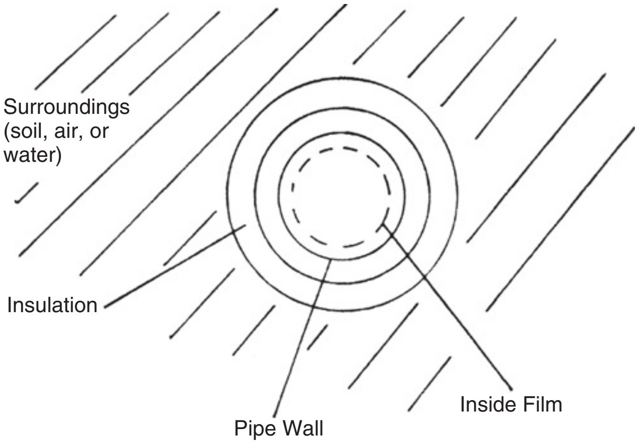


Figure 3-7. Cross section of a pipe showing resistance layers.

Table 3-4
Types of Resistance Layers for a Pipe

Resistance	Due to
$R_{\text{inside, film}}$	Boundary layer on the inside of the pipe
R_{pipe}	Material from which the pipe is made
$R_{\text{insulation}}$	Insulation (up to five concentric layers)
R_{surr}	Surroundings (soil, air, water)

calculation of viscosity and surface tension. When the physical properties of the fluids are calculated for two-phase flow, the physical properties of the gas and liquid mixture can be calculated by taking the mole fraction of these components into account. Similarly, when water is present in the system, the properties of oil and water are combined into those of a pseudo-liquid phase. To produce the phase split for a given composition, pressure, and temperature, an equilibrium flash calculation utilizing an appropriate equation of state must be used. A simple and stable three-phase flash calculation with significant accuracy for pipeline calculation has been described by Mokhatab (2003) (see Appendix 1).

3.7 VELOCITY CRITERIA FOR SIZING MULTIPHASE PIPELINES

Nowadays line sizing of raw gas lines (i.e., wellhead to platform) is performed using integral reservoir deliverability conditions. However, the procedure is to specify the line size and predict the production profile. The maximum velocities are then checked to ensure that erosion–corrosion problems do not occur.

3.7.1 Erosion Criteria

The primary sizing criterion for multiphase flow pipelines is erosional velocity, which should be kept below the known fluid erosional velocity. However, the sizing of pipelines for multiphase flow is significantly more complex than for single-phase flow because the resultant erosional conditions are totally dependent on the specific flow regimes. If sand

Table 3-5
Heat Transfer Resistances for Pipes (Hein, 1984; Brill and Beggs, 1991)

$R_{\text{inside, film}} = D/[0.027\ K_f\ Re^{0.8}\ Pr^{0.33}]$	For stationary surroundings like soil,
D is reference diameter on which U is based, inch	
K_f is thermal conductivity of the fluid, Btu/hr-ft-°F	$R_{\text{surr}} = \dfrac{D \ln\left[(2D' + (4D'^2 - D_t^2)^{0.5})/D_t\right]}{24k_{\text{surr}}}$
Re is Reynolds number $= \dfrac{(\rho_L \lambda_L + \rho_G \lambda_G)(V_{SL} + V_{SG})(D_i)(124.016)}{\lambda_L \mu_L + \lambda_G \mu_G}$	D is reference diameter on which U is based, inch
	k_{surr} is soil thermal conductivity, Btu/hr-ft-°F
	D' is depth from top of soil to pipe centerline, ft
	D_t is diameter of the pipe plus insulation, ft
ρ_L is density of the liquid, lb/ft^3	For fluid surroundings like air or water,
λ_L is no-slip liquid holdup	
ρ_G is density of the gas, lb/ft^3	$R_{\text{surr}} = \dfrac{D}{12k_{\text{surr}} 10^{[0.26694(\log \text{Re}_{\text{surr}})^{1.3681}]}}$
λ_G is no-slip gas holdup $= 1 - \lambda_L$	
V_{SL} is liquid superficial velocity, ft/sec	
V_{SG} is gas superficial velocity, ft/sec	D is reference diameter on which U is based, inch
D_i is inside diameter of the pipe, inch	k_{surr} is thermal conductivity of the surroundings, Btu/hr-ft-°F
μ_L is liquid viscosity, cp	

Continued

Table 3-5
Heat Transfer Resistances for Pipes (Hein, 1984; Brill and Beggs, 1991)—cont'd

μ_G is gas viscosity, cp

Pr is Prandtl number $= 2.42(\mu_L\lambda_L + \mu_G\lambda_G)(C_{pL}\lambda_L + C_{p,G}\lambda_G)/K_f$

C_p is specific heat at constant pressure, Btu/lb-°F

$$R_{pipe} = \frac{D\ln(D_o/D_i)}{24k_p}$$

k_p is thermal conductivity of the pipe, Btu/hr-ft-°F

D_o is outside diameter of the pipe, ft

D_i is inside diameter of the pipe, ft

$$R_{insulation} = \frac{D\sum_{j=1}^{n}\frac{1}{k_j}\ln(D_j/D_j - 1)}{24}$$

D is reference diameter on which U is based, inch

k_j is conductivity of jth layer of insulation, Btu/hr-ft-°F

D_j is outer diameter of the jth layer, ft

D_{j-1} is inner diameter of the jth layer, ft

Re_{surr} is Reynolds number for the surroundings $= 0.0344\rho_{surr}\ V_{surr}\ D_t/\mu_{surr}$

ρ_{surr} is density of ambient fluid, lb/ft^3

V_{surr} is velocity of ambient fluid, ft/hr

D_t is total diameter of pipe plus insulation, inch

μ_{surr} is viscosity of surrounding fluid, cp production is considered, the situation is even more complex. In general, when considering erosion criteria, the flow velocity must be limited to the following conditions (Mokhatab, 2004).

For duplex stainless steel or alloy material:

- 328 ft/s in multiphase lines in stratified flow regimes (213 ft/s for 13% Cr. material).
- 66 ft/s in multiphase lines in annular, bubble, or hydrodynamic slug flow regime.
- 230 ft/s in multiphase lines in mist flow regimes.

For carbon steel material:

- For continuous injection of corrosion inhibitor, the inhibitor film ensures a lubricating effect that shifts the erosion velocity limit. The corrosion inhibitor erosion velocity limit will be calculated taking into account the inhibitor film wall shear stress (Smart, 2001).
- For injection of corrosion inhibitor, use Equation (3-50) to determine the erosion velocity (API RP 14E, 1991):

$$V_e = \frac{C}{(\rho_M)^{0.5}} \tag{3-50}$$

where V_e is erosional velocity, ft/s; C is empirical constant; and ρ_M is gas/liquid mixture density at flowing conditions, lb/ft^3.

Industry experience to date indicates that for solid-free fluids, a C factor of 100 for continuous service and a C factor of 125 for intermittent service are conservative. However, in the latest API RP 14E (1991), higher C values of 150 to 200 may be used when corrosion is controlled by inhibition or by employing corrosion-resistant alloys.

The aforementioned simple criterion is specified for clean service (noncorrosive and sand free), and the limits should be reduced if sand or corrosive conditions are present. The lower limit of the flow velocity range should be the velocity that keeps erosive matter in suspension, and the upper limit should be when erosion–corrosion and cavitations attacks are minimal. However, no guidelines are provided for these reductions (Salama, 2000; McLaury and Shirazi, 2000).

3.7.2 Corrosion Criteria

The following conditions for corrosion criteria should be considered.

- For corrosion-resistant material, no limitation of flowing velocity up to 328 ft/s and no requirement for corrosion allowance.
- For noncorrosion-resistant material, in a corrosive fluid service, or when a corrosion inhibitor is injected, corrosion allowances for a design service life are required.

Note that it is rarely economical for pipelines to use corrosion-resistant alloys, thus corrosion inhibitors and internal cladding are used as cost-effective alternatives. The injected inhibitor forms a protective layer on the internal surface of the pipe, and the continuity and the quality of this layer are the controlling factors for effective corrosion protection. However, for some specific flow patterns in multiphase flows, such as slug flow, the corrosion may be very severe due to removal of the corrosion product and inhibitor layers by turbulence and high shear stress (Kang *et al.*, 1996). However, the slug can also help by distributing the corrosion inhibitor and preventing top-of-line corrosion. Determination of the corrosion allowance for deepwater pipelines should be made using the corrosion-inhibitor availability model that reflects actual performance and realistic inhibitor availability rather than arbitrary inhibitor effectiveness criteria (Smart, 2001).

After selecting the appropriate outside diameter for a pipe, it is necessary to determine a suitable wall thickness that can withstand the internal pressure. For a given outside pipe diameter, the pipe wall thickness is calculated to withstand different loads (e.g., lateral loading, external pressure collapse, installation stresses). Typical design practice limits hoop stress and differential pressure and limits the equivalent stress for design against combined loads (API RP 14E, 1991; Bay, 2001). This practice has proved to be very safe in general, except when external impact loads are critical to the integrity of the pipeline (Sotberg and Bruschi, 1992). Nevertheless, this practice has been used by the pipeline industry for decades with little change, despite significant improvements and developments in pipeline technology (Verley *et al.*, 1994). Additional design details of wall thickness are reviewed and compared with industry practice by Bay (2001).

3.8 MULTIPHASE FLOW ASSURANCE

In thermal-hydraulic design of multiphase flow transmission systems, the system designer is faced with several challenges associated with multiphase flow, which can significantly change design requirements. The goal of any pipeline designer is to secure "flow assurance," i.e., the transmission system must operate in a safe, efficient, and reliable manner throughout the design life. Failure to do so has significant economic consequences, particularly for offshore gas production systems. "Flow assurance" covers the whole range of possible flow problems in pipelines, including both multiphase flow and fluid-related effects such as gas hydrate formation, wax and asphaltene deposition on walls, corrosion, erosion, scaling, emulsions, foaming, and severe slugging. However, the major flow assurance issues for raw gas transmission lines are hydrates, corrosion, wax, and severe slugging, which are discussed individually in different sections later. The avoidance or remediation of these problems is the key aspect of flow assurance that enables the design engineer to optimize the production system and to develop safe and cost-effective operating strategies for the range of expected conditions, including start-up, shutdown, and turndown scenarios. However, as production systems go deeper and deeper, flow assurance becomes a major issue for offshore production and transportation systems, where traditional approaches are inappropriate for deepwater development systems due to extreme distances, depths, temperature, or economic constraints (Wilkens, 2002).

3.8.1 Gas Hydrates

A gas hydrate is an ice-like crystalline solid called a clathrate, which occurs when water molecules form a cage-like structure around smaller guest molecules. The most common guest molecules are methane, ethane, propane, isobutane, normal butane, nitrogen, carbon dioxide, and hydrogen sulfide, of which methane occurs most abundantly in natural hydrates. Several different hydrate structures are known. The two most common are structure I and structure II. Type I forms with smaller gas molecules such as methane, ethane, hydrogen sulfide, and carbon dioxide, whereas structure II is a diamond lattice, formed by large molecules such as propane and isobutane (Sloan, 1998). However, nitrogen, a relatively small molecule, also forms a type II hydrate (Carroll, 2003). In addition, in the presence of free water, the temperature and pressure can also govern the type of hydrate structure, where the hydrate structure may change from structure II hydrate at low temperatures and pressures to structure I hydrate at high pressures and temperatures (GPSA, 1998). It should be noted that *n*-butane does form a hydrate, but is very unstable (Ng and Robinson, 1976; Kumar, 1987). However, it will form a stabilized hydrate in the presence of small "help" gases such as methane or nitrogen (Edmonds *et al.*, 1998). It has been assumed that normal paraffin molecules larger than *n*-butane are nonhydrate formers (Kumar, 1987). Furthermore, the existence of a new hydrate structure, type H, has been described by Ripmeester *et al.* (1987). Some isoparaffins and cycloalkanes larger than pentane are known to form structure H hydrates (Mehta and Sloan, 1996). However, little is known about type H structures.

While many factors influence hydrate formation, the two major conditions that promote hydrate formations are the gas being (1) at the appropriate temperature and pressure and (2) at or below its water dew point (Sloan, 1998). Note that free water is not necessary for hydrate formation, but it certainly enhances hydrate formation (Carroll, 2003). Other factors that affect, but are not necessary for, hydrate formation include turbulence, nucleation sites, surface for crystal formation, agglomeration, and the salinity of the system. The exact temperature and pressure at which hydrates form depend on the composition of the gas and the water. For any particular composition of gas at a given pressure, there is a temperature below which hydrates will form and above which hydrates will not form. As the pressure increases, the hydrate formation temperature also increases. As a general rule, when the pressure of the gas stream increases or as the gas becomes colder, the tendency to form hydrates increases (GPSA, 1998). Hence, many gas-handling systems are at significant risk of forming hydrate plugs during shut-in and subsequent start-up (Hunt, 1996; Ramachandran *et al.*, 2000; Wilkens, 2002).

Although gas hydrates may be of potential benefit both as a important source of hydrocarbon energy and as a means of storing and transmitting natural gas, they represent a severe operational problem, as the hydrate crystals may deposit on the pipe wall and accumulate as large plugs that can completely block pipelines, shutting in production (Sloan, 1998, GPSA, 1998, Edmonds *et al.*, 1998). Acceleration of these plugs due to a pressure gradient can also cause considerable damage to production facilities. In addition, the remediation of hydrate blockages can present significant technical difficulties with major cost implications (Mehta *et al*., 2001). Because of these problems, methods of preventing hydrate solids development in gas production systems have been of considerable interest for a number of years (Bufton, 2003).

3.8.1.1 Hydrate Locus for Natural Gas Components

The thermodynamic stability of hydrates, with respect to temperature and pressure, may be represented by the hydrate curve. The hydrate curve represents the thermodynamic boundary between hydrate stability and dissociation. Conditions to the left of the curve represent situations in which hydrates are stable and "can" form. Operating under such conditions does not necessarily mean that hydrates will form, only that they are possible. Figure 3-8 shows the hydrate locus for natural gas components. The extension to mixtures is not obvious from this diagram. The hydrate curve for multicomponent gaseous mixtures may be generated by a series of laboratory experiments or, more commonly, is predicted using thermodynamic software based on the composition of the hydrocarbon and aqueous phases in the system.

The thermodynamic understanding of hydrates indicates the conditions of temperature, pressure, and composition that hydrates are stable. However, it does not indicate when hydrates will form and, more importantly, whether they will cause blockages in the system.

3.8.1.2 Prediction of Hydrate Formation Conditions

There are numerous methods available for predicting hydrate formation conditions. Three popular methods for rapid estimation of hydrate formation conditions are discussed. A detailed discussion of other methods that are perhaps beyond the scope of the present discussion can be found in publications by Kumar (1987), Sloan (1998), and Carroll (2003).

3.8.1.2.1 *K*-Factor Method This method was developed originally by Carson and Katz (1942), although additional data and charts have been reproduced since then. In this method, the hydrate temperature can be predicted using vapor–solid (hydrate) equilibrium constants. The basic

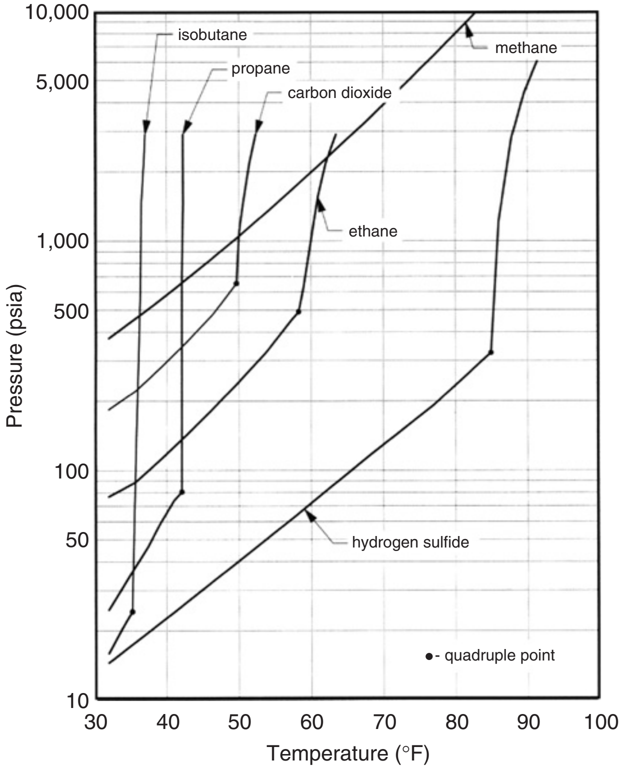


Figure 3-8. Hydrate-forming conditions for natural gas components (Carroll, 2003).

equation for this prediction is (Carson and Katz, 1942)

$$\sum_{i=1}^{n}\left(\frac{y_{\mathrm{i}}}{K_{\mathrm{i}}}\right) = 1.0 \tag{3-51}$$

where y_{i} is mole fraction of component i in gas on a water-free basis, K_{i} is vapor–solid equilibrium constant for component i, and n is number of components.

The calculation is iterative, and the incipient solid formation point will determine when the aforementioned equation is satisfied. This procedure is akin to a dew point calculation for multicomponent gas mixture.

The vapor–solid equilibrium constant is determined experimentally and is defined as the ratio of the mole fraction of the hydrocarbon component in gas on a water-free basis to the mole fraction of the hydrocarbon component in the solid on a water-free basis (Carson and Katz, 1942):

$$K_{\mathrm{i}} = \left(\frac{y_{\mathrm{i}}}{x_{\mathrm{i}}}\right) \tag{3-52}$$

where x_{i} is mole fraction of component *i* in solid on a water-free basis.

Figures 3-9 through 3-13 provide the vapor–solid equilibrium constants at various temperatures and pressures. For nitrogen and components heavier than butane, the equilibrium constant is taken as infinity. It should be stressed that in the original method of Carson and Katz (1942) it is assumed that nitrogen is a nonhydrate former and that *n*-butane, if present in mole fractions less than 5%, has the same equilibrium constant as ethane. Theoretically, this assumption is not correct, but from a practical viewpoint, even using an equilibrium constant equal to infinity for both nitrogen and *n*-butane, provides acceptable engineering results (Campbell, 1992).

The Carson and Katz (1942) method gives reasonable results for sweet natural gases and has been proven to be appropriate up to about 1000 psia (GPSA, 1998). However, Mann *et al*. (1989) presented new *K*-value charts that cover a wide range of pressures and temperatures. These charts can be an alternate to the Carson and Katz (1942) *K*-value charts, which are not a function of structure or composition.

3.8.1.2.2 Baillie and Wichert Method The method presented by Baillie and Wichert (1987) is a chart method (Figure 3-14) that permits estimation of hydrate formation temperatures at pressure in the range of 100 to 4000 psia for natural gas containing up to 50% hydrogen sulfide and up to 10% propane (Carroll, 2003). The method may not apply to a sweet gas mixture containing CO_2, but is considered fairly accurate if the CO_2 is less than about 5 mol% (Carroll, 2004).

3.8.1.2.3 Gas Gravity Method Until now, several other methods have been proposed for predicting hydrate-forming conditions in natural gas systems. The most reliable of these requires a gas analysis. However, if the gas composition is not known, even the previous methods cannot be used to predict the hydrate formation conditions, and the Katz (1945)

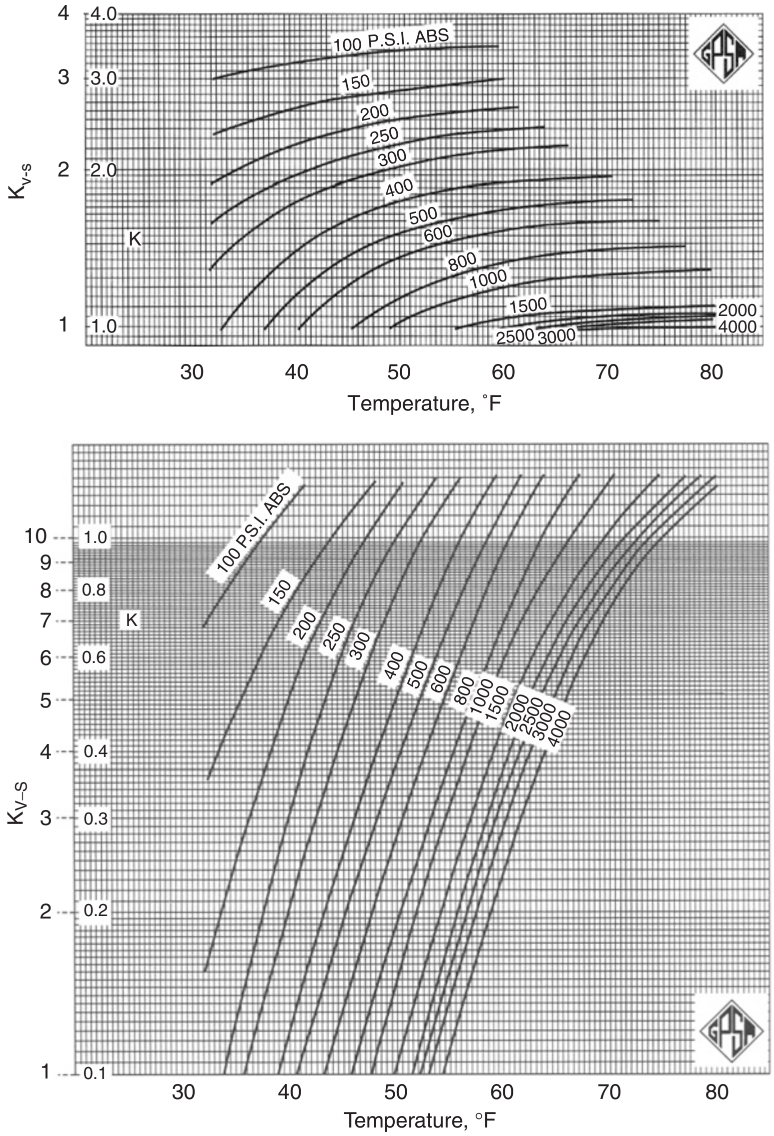


Figure 3-9. Vapor–solid equilibrium constants for methane (top) and ethane (bottom) (GPSA, 1998).

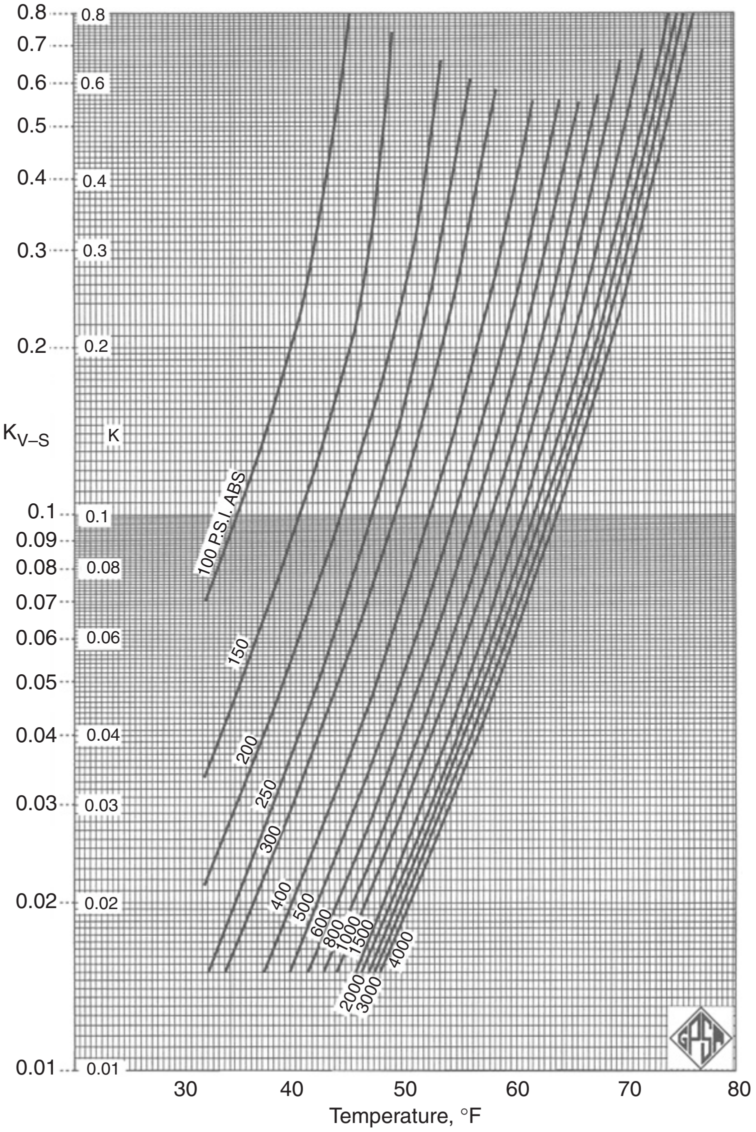


Figure 3-10. Vapor–solid equilibrium constant for propane (GPSA, 1998).

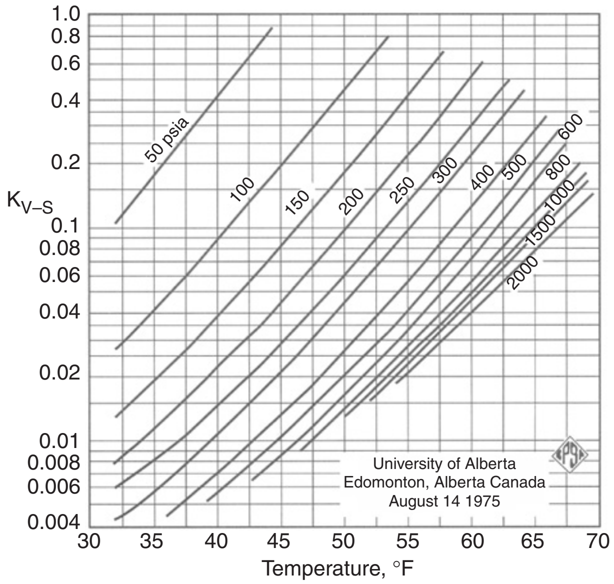


Figure 3-11. Vapor–solid equilibrium constants for isobutane (GPSA, 1998).

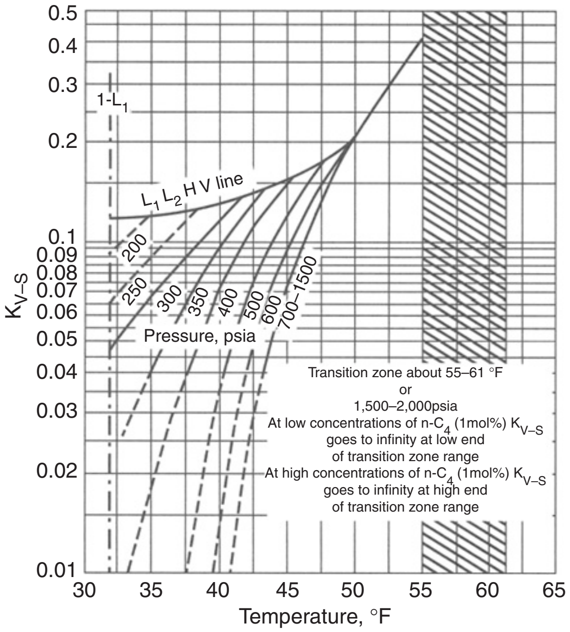


Figure 3-12. Vapor–solid equilibrium constants for *n*-butane (GPSA, 1998).

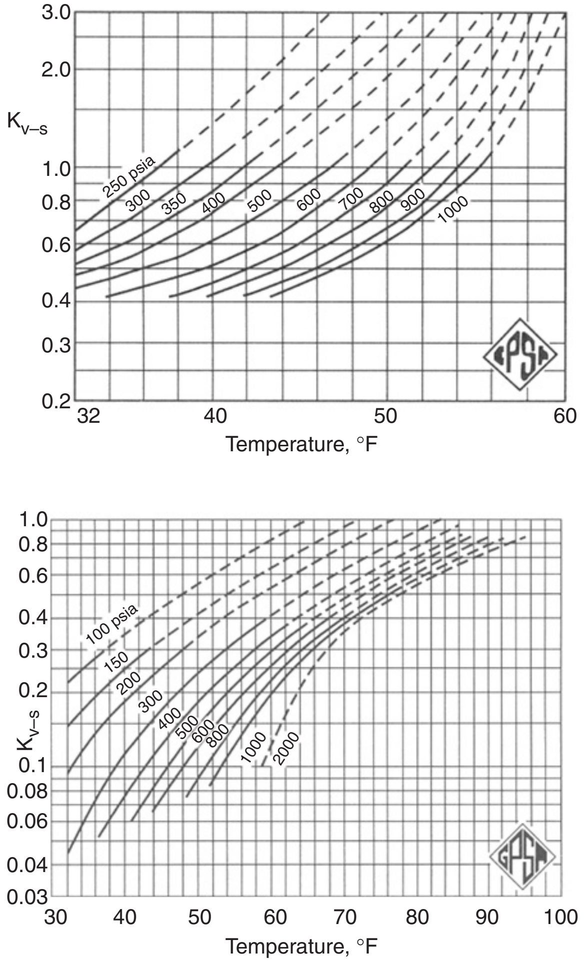


Figure 3-13. Vapor–solid equilibrium constants for carbon dioxide (top) and for hydrogen sulfide (bottom) (GPSA, 1998).

gravity chart (Figure 3-15) can be used to predict the approximate pressure and temperature for hydrate formation, provided hydrates exist in the pressure–temperature region above the appropriate gravity curve. Therefore, as a first step to predict hydrate formation temperature, one can develop an appropriate equation representing the Katz (1945) gravity chart. Such a correlation uses two coefficients that correlate temperature,

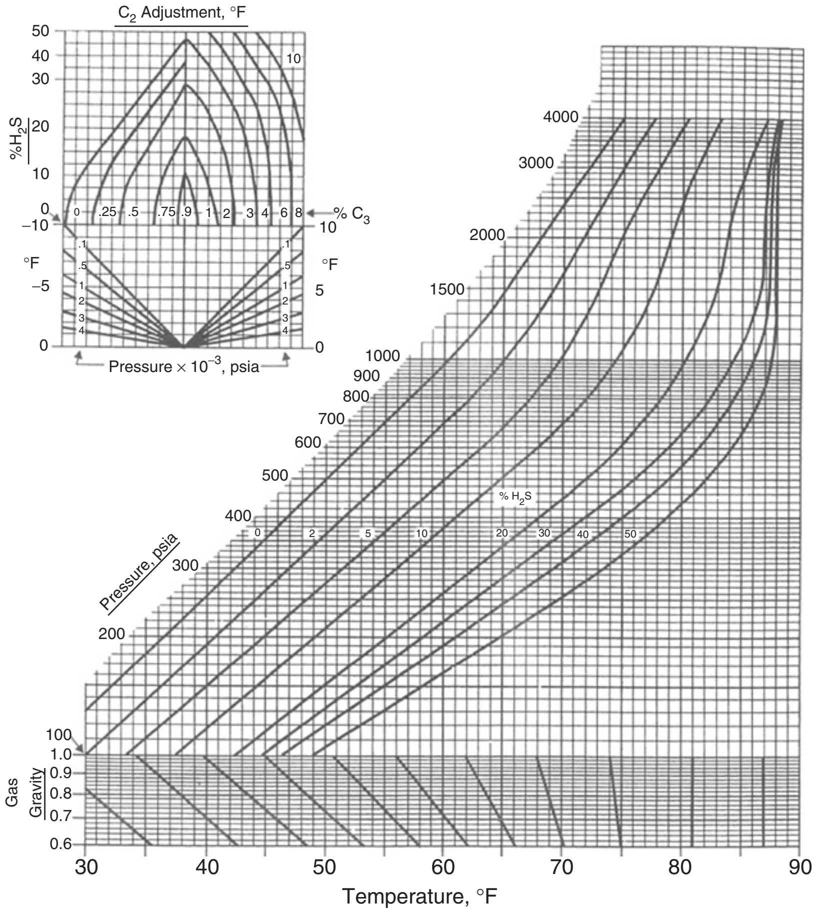


Figure 3-14. Baillie and Wichert chart for estimating hydrate formation conditions (GPSA, 1998).

pressure, and specific gravity of gas (Towler and Mokhatab, 2005):

$$T_h = 13.47\ln(P) + 34.27\ln(\mathrm{SG}) - 1.675\,[\ln(P) \times \ln(\mathrm{SG})] - 20.35 \tag{3-53}$$

where T is gas flow temperature, °F; P is gas flow pressure, psia; and SG is specific gravity of gas (air = 1.0).

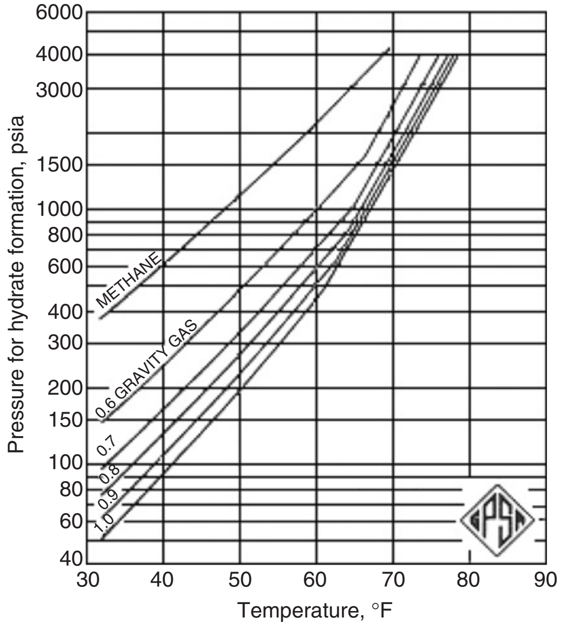


Figure 3-15. Katz's gravity chart for predicting hydrate formation conditions (GPSA, 1998).

Note that while Equation (3-53) is based on the GPSA chart, it is only accurate up to 65°F. Beyond that it overestimates the temperature slightly.

The Katz (1945) gravity chart was generated from a limited amount of experimental data and a more substantial amount of calculations based on the *K*-value method. The components used for the construction of this chart are methane, ethane, propane, butane, and normal pentane; therefore, using this chart for compositions other than those used to derive these curves will produce erroneous results (Sloan, 1998). In fact, this method is an appropriate method of estimating hydrate formation conditions for sweet natural gas mixtures. However, the Baillie and Wichert (1987) method is better than this chart when applied to sweet gas because of the inclusion of a correction factor for propane (Carroll, 2003).

3.8.1.2.4 Commercial Software Programs Hydrate formation conditions based on fluid compositions are normally predicted with the commercially available software programs. These programs are generally quite good and so simple to use that they often require less time than the simplified methods presented. The bases of these computer programs are the statistical thermodynamic models, which use a predictive algorithm with additional experimental data included to modify or "tune" the mathematical predictions. Most commercially available softwares use algorithms developed by D.B. Robinson and Associates (EQUIPHASE) and by Infochem Computer Services Ltd (MULTIFLASH). A detailed discussion of the accuracy of these programs and other hydrate software programs can be found in Sloan (1998) and Carroll (2003).

3.8.1.3 Hydrate Prevention Techniques

The multiphase fluid produced at the wellhead will normally be at a high pressure and a moderate temperature. As the fluid flows through the pipelines, it becomes colder, which means such pipelines could experience hydrates at some point in their operating envelope. For this reason, the hydrate formation in gas transmission pipelines should be prevented effectively and economically to guarantee that the pipelines operate normally. Control of hydrates relies on keeping the system conditions out of the region in which hydrates are stable. It may be possible to keep the fluid warmer than the hydrate formation temperature (with the inclusion of a suitable margin for safety) or operate at a pressure less than the hydrate formation pressure.

Although there are several methods to avoid hydrate formation, depending on the possible location of a gas hydrate plug, some of the techniques to remediate onshore hydrates (e.g., installation of line heaters and line depressurization) may not be practical in long and high pressure, subsea gas transmission pipelines (Son and Wallace, 2000; Wilkens, 2002).

There are few methods of preventing hydrate formation in offshore transmission systems. The permanent solution is removal of water prior to pipeline transportation, using a large offshore dehydration plant that is not often the most cost-effective solution. In general, two methods are applicable at the well site, namely thermal and chemical. These techniques are discussed in the following sections.

3.8.1.3.1 Thermal Methods Thermal methods use either conservation or introduction of heat in order to maintain the flowing mixture outside the hydrate formation range. Heat conservation is common practice and is accomplished through insulation (Oram, 1995; Hunt, 1996). This method can be feasible for some subsea applications depending on the fluid being transported, the tieback distance, and topside capabilities of the host platform. The design of such conservation systems typically seeks a balance among the high cost of the insulation, the intended operability of the system, and the acceptable risk level (Fidel-Dufour and Herri, 2002).

A number of different concepts are available for introducing additional heat to a pipeline. The simplest is an external hot-water jacket, either for a pipe-in-pipe system or for a bundle. Other methods use either conductive or inductive heat tracing (Hunt, 1996). There is concern over the reliability of conductive systems. An electrical resistance heating system may be desirable for long offset systems, where available insulation is insufficient, or for shut-in conditions (Langner *et al*., 1999). The ability to heat during production depends on the specific electrical heating implementation. Such systems provide environmentally friendly fluid temperature control without flaring for pipeline depressurization. The effect is also an increase in production as there is no time lost by unnecessary depressurization, pigging, heating-medium circulation, or removal of hydrate blockage. However, it is still difficult to persuade operators to install an acting heating system.

3.8.1.3.2 Chemical Inhibition An alternative to the thermal processes is chemical inhibition. Chemical inhibitors are injected at the wellhead and prevent hydrate formation by depressing the hydrate temperature below that of the pipeline operating temperature. Chemical injection systems for subsea lines have a rather high capital expenditure price tag associated with them, in addition to the often high operating cost of chemical treatment (Covington *et al*., 1999; Sloan, 2000; Frostman *et al*., 2003). However, hydrate inhibition using chemical inhibitors is still the most widely used method for unprocessed gas streams, and the development of alternative, cost-effective, and environmentally acceptable hydrate inhibitors is a technological challenge for the gas production industry.

3.8.1.3.2.1 Types of Inhibitors Traditionally, the most common chemical additives used to control hydrates in gas production systems have been methanol, ethylene glycol, or triethylene glycol at a high enough concentration (Sloan, 1991). These chemicals are called “thermodynamic inhibitors” and have the effect of shifting the hydrate curve to the left, which causes the hydrate stability point to be displaced to a lower temperature and/or a higher pressure. Increasing salt content in the produced brine (by injecting electrolyte solutions such as sodium chloride, calcium chloride, and potassium chloride) can also provide some hydrate temperature suppression, but normally this alone is not sufficient to avoid hydrates in the subsea gas production systems. In some cases, blended inhibitors of methanol/glycols and electrolyte(s) are preferred for subsea applications (Zuo and Zhang, 1999).

The inhibitor selection process often involves comparison of many factors, including capital/operating cost, physical properties, safety, corrosion inhibition, and gas dehydration capacity. However, a primary factor in the selection process is whether the spent chemical will be recovered, regenerated, and reinjected. Usually methanol is not regenerated because its use is intermittent (i.e., during start-up or shutdown). However, when it is injected continuously, as is often observed in gas systems then it is sometimes regenerated. However, losses to the vapor phase can be prohibitive, in which case operators select monoethylene glycol (GPSA, 1998). Often when applying this inhibitor, there is a significant expense associated with the cost of "lost" methanol. However, because methanol has lower viscosity and lower surface tension it makes for an effective separation from the gas phase at cryogenic conditions (below –13°F) and is usually preferred (Esteban *et al*., 2000). In many cases, hydrate plug formation is prevented through the addition of glycols (usually ethylene glycol because of its lower cost, lower viscosity, and lower solubility in liquid hydrocarbons, and lower vapor pressure giving reduced gas-phase losses) to depress the hydrate formation temperature (Kohl, 1985). However, in order to be effective, glycols must be added at rates of up to 100% of the weight of water. Because glycols are expensive inhibitors, there is a definite need for extra, costly, and space-consuming onshore or offshore plants for their regeneration. Therefore, it would be useful to develop new hydrate depressants, which can be used at much lower concentrations and therefore much lower cost. Two new types of low-dosage inhibitors have been developed, that will enable the subsea gas transmission pipelines to handle increased gas volumes without additional glycol injection or extra glycol recovery units (Bloys *et al*., 1995; Lederhos *et al*., 1996; Palermo *et al*., 2000). These new hydrate inhibitors can lead to very substantial cost savings, not only for the reduced cost of the new inhibitor, but also in the size of the injection, pumping, and storage facilities (Mehta *et al*., 2002; Frostman *et al*., 2003). These new hydrate inhibitors, called "low-dosage hydrate inhibitors" (LDHIs), form the basis of a technique that does not operate by changing the thermodynamic conditions of the system. In fact, LDHIs act at the early stages of hydrate formation by modifying the rheological properties of the system (Sinquin *et al*., 2004). There are two types of LDHIs: “kinetic hydrate inhibitors” (KHIs), and “antiagglomerants” (AAs). Most commercial kinetic inhibitors are high molecular weight polymeric chemicals (i.e., poly[*N*-vinyl pyrrolidone] or poly[vinylmethylacetamide/vinylcaprolactam]), which are effective at concentrations typically 10 to 100 times less than thermodynamic inhibitors concentrations. KHIs may prevent crystal nucleation or growth during a sufficient delay compared to the residence time in the pipeline. The deeper a system operates in the hydrate region, the shorter the time during which kinetic hydrate inhibitors can delay hydrate formation. The achievable delays range between weeks if the pipeline operates less than 42°F in the hydrate region to hours if the pipeline operates at 50°F in the hydrate region (Mokhatab *et al*., 2006b). Kinetic inhibitors are relatively insensitive to the hydrocarbon phase and may therefore turn out to be applicable to a wide range of hydrocarbon systems. However, the industrial application of kinetic inhibitors depends on the repeatability of multiphase pipeline testing results among laboratory, pilot plant and field, and the transferability among different plants (Kelland *et al*., 2000).

In contrast to other types of inhibitors, AAs, which are surface-active chemicals (i.e., alkyl aromatic sulphonates or alkylphenylethoxylates), do not prevent the formation of hydrate crystals but keep the particles small and well dispersed so that fluid viscosity remains low, allowing the hydrates to be transported along with the produced fluids. AAs performance is relatively independent of time. In addition, AAs appear to be effective at more extreme conditions than KHIs, which makes these products of interest to operators looking for cost-effective hydrate control in deepwater fields. These additives are currently applied in the Gulf of Mexico, the North Sea, and west Africa (Mehta *et al*., 2002; Frostman, 2003). However, they have mainly limitations in terms of water cut, where they require a continuous oil phase and therefore are only applicable at lower water cuts. The maximum water cut is expected to be between 40 and 50%. This limitation is caused by the rheological properties of suspensions with high solid fraction and may depend on flow regime conditions (Kelland *et al*., 1995; Frostman, 2000; Mehta *et al*., 2002).

As stated earlier, the choice between inhibitor alternatives must be based on physical limitations as well as economics. However, operating conditions may also limit the number of available choices. For example, in a project carried out by Baker Petrolite it was shown that under severe conditions, the required dosage of an antiagglomerator unlike thermodynamic and kinetic inhibitors does not increase as the degree of subcooling increases. Therefore, this method of treatment would be a cost-effective solution for the control of gas hydrates (Ramachandran *et al.*, 2000).

3.8.1.3.2.2 Prediction of Inhibitor Requirements The inhibitor must be present in a minimum concentration to avoid hydrate formation. Accurate prediction of this minimum inhibitor concentration is required for cost-effective design and operation of multiphase pipelines. Various empirical methods, charts, and computer programs have been developed for this purpose, including the venerable Hammerschmidt's (1934) empirical equation, which is a relatively simple method that has been used to calculate the amount of inhibitor required in the water phase to lower the hydrate formation temperature,

$$\Delta T = \frac{KW}{100(\mathrm{MW}) - (\mathrm{MW})W} \tag{3-54}$$

where ΔT is depression of hydrate formation temperature, °F; MW is molecular weight of inhibitor; W is weight percentage of inhibitor in final water phase; and K is constant, depending on the type of inhibitor.

Experimentally determined values of K and molecular weights of inhibitors are given in Table 3-6. To use this equation, the hydrate formation temperature in the gas without the inhibitor being present must be known. In fact, Equation (3-54) only predicts the deviation from the hydrate formation temperature without an inhibitor present.

The Hammerschmidt equation is limited to inhibitor concentrations of about 20–25 wt% for methanol and 60–70 wt% for glycols (GPSA, 1998). However, for higher methanol concentrations, Nielsen and Bucklin (1983) have recommended the following equation:

$$\Delta T = -129.6 \ln(X_{\mathrm{H_2O}}) \tag{3-55}$$

where $X_{\mathrm{H_2O}}$ is the mole fraction of water in the aqueous phase.

Table 3-6
Physical Constants of Inhibitors
(Hammerschmidt, 1934)

Inhibitor	MW	*K*
Methanol	32.04	2335
Ethanol	46.07	2335
Isopropanol	60.10	2335
Ethylene glycol	62.07	2200
Propylene glycol	76.10	3590
Diethylene glycol	106.10	4370
Triethylene glycol	150.17	5400

They claim that this equation is accurate up to 90 wt% methanol, which gives the maximum suppression because methanol freezes at concentrations above 90 wt%. However, studies of GPSA (1998) only recommend the Nielsen and Bucklin (1983) equation for methanol concentrations ranging from 25 to 50 wt%. Equation (3-55) was developed for use with methanol; however, this equation is actually independent of the choice of inhibitor and therefore, theoretically, it can be used for any glycols.

Maddox *et al.* (1991) described a graphical procedure for estimating the required inhibitor concentration for both methanol and glycol. This method is a trial-and-error approach, which can be used when the activity coefficients of water in methanol and glycol are available. Although this method provides better accuracy, no recommended applicable range is provided (GPSA, 1998, Esteban *et al.*, 2000).

If the produced water or seawater is in contact with the hydrocarbon fluid, the salinity of the water will itself inhibit hydrate formation. Therefore, it is important to be able to estimate the effect of the brine in the produced water on the hydrate formation temperature. For this purpose, McCain (1990) presented the following equation:

$$\Delta T = \mathrm{AS} + \mathrm{BS}^2 + \mathrm{CS}^3 \tag{3-56}$$

where ΔT is temperature depression, °F; and S is water salinity, wt%.

$$\mathrm{A} = 2.20919 - 10.5746\,(\mathrm{SG}) + 12.1601\,(\mathrm{SG}^2) \tag{3-56-1}$$

$$B = -0.10605 + 0.72269\,(SG) - 0.85093\,(SG^2) \qquad (3\text{-}56\text{-}2)$$

$$C = 0.00347 - 0.01655\,(SG) + 0.01976\,(SG^2) \qquad (3\text{-}56\text{-}3)$$

Equation (3-56) is limited to salt concentrations of less than 20 wt% and for gas-specific gravities (SG) ranging from 0.55 to 0.68.

All of these simple methods predict the depression of the hydrate formation temperature; they do not predict the actual hydrate formation conditions. However, several thermodynamic models have been proposed for predicting the hydrate formation conditions in aqueous solutions containing methanol/glycols and electrolytes (Edmonds *et al.*, 1996; Nasrifar *et al.*, 1998; Zuo and Zhang, 1999). These rigorous models can also account for the effect of pressure and the type of hydrate neglected in the simple mentioned methods. However, available models have limitations that include the types of liquid, compositions of fluids, and inhibitors used.

The amount of inhibitor required to treat the water phase, plus the amount of inhibitor lost to the vapor phase and the amount that is a soluble in the hydrocarbon liquid, equals the total amount required. Methanol vaporization loss can be estimated from Figure 3-16, while glycol

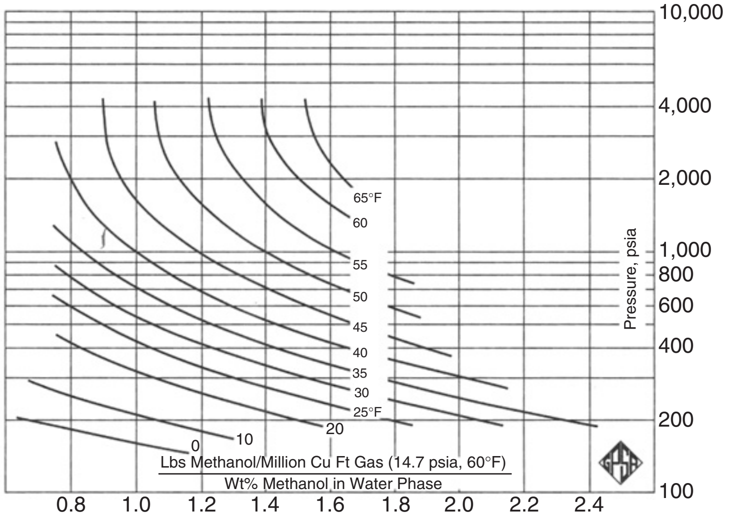


Figure 3-16. Prediction of methanol vaporization loss (GPSA, 1998).

vaporization losses are generally very small and typically can be ignored (Chen *et al.*, 1988; Sloan, 2000).

In addition, prediction of inhibitor losses to the hydrocarbon liquid phase is difficult. However, many of the commercially available software programs include proper calculations to account for the loss of methanol and glycols to the hydrocarbon liquid phase.

3.8.1.3.2.3 Design of Injection Systems Proper design of an inhibitor injection system is a complex task that involves optimum inhibitor selection, determination of the necessary injection rates, pump sizing, and pipeline diameters. Inhibitors for a subsea gas transmission system are selected before gas production is started on the facility. This makes inhibitor selection difficult, as a large number of factors, including brine composition, temperature, and pressure, that affect the performance of inhibitors are unknown (Ramachandran *et al.*, 2000). Therefore, at this stage, an appropriate multiphase flow simulation package must be used to calculate some of the unknown necessary variables, which are required for injection systems design.

To determine the appropriate injection rate, it is preferable to determine inhibitor requirements by field testing. The inhibitor-dosing requirements are used later to determine the requirements for inhibitor storage, pumping capacities, and number of inhibitor lines in order to ensure that the inhibitor can be delivered at the required rates for treating wells and flow lines during start-up and shut-in operations. In addition, points of injection should be chosen to provide maximum benefit in the pipeline system. The most effective position for injection is generally at the center of the pipe in the direction of the product flow. However, the injection rate and the location of injection points are a function of flow geometry, fluid properties, and pressure–temperature relationships that are encountered in the actual field application. For instance, hydrate inhibitors for a deepwater production system are often pumped through long umbilicals that are one-half to one-fourth inch in diameter. The injection pump is a positive displacement metering pump (Kumar, 1987) capable of generating sufficient injection line pressure (normally between 3000 and 4000 psi) to overcome the line operating pressure. Ideally the injection pressure should be 100 psi above line pressure. However, varied injection rates can be achieved by changing the pressure differential. Note that Ramachandran *et al.* (2000) provide a discussion on the proper design of deepwater injection systems and predicting systems behavior.

3.8.2 Corrosion

One of the common problems in multiphase flow transmission pipelines is metal corrosion. Corrosion is defined as the deterioration of material, usually a metal, due to its reaction with the environment or handling media. The cause of corrosion can be directly attributed to the impurities found in the produced gas, as well as the corrosive components that are by-produced. Because of the nonspecificity of the components produced from a production well, some or all of these components may be active to create a corrosive environment in the pipelines. Corrosion in multiphase systems is a complex phenomenon, including dependency on the partial pressure, temperature, pH, and concentration of corrosion products. Consequently, corrosion prediction requires substantial understanding of the simultaneous interaction of the many process variables that govern both flow and corrosion conditions.

An important aspect of maintaining pipeline performance is adequate control of corrosion both internally—caused by the flow components and their by-products—and externally—because of pipeline exposure to the soil and water (Hartt and Chu, 2004). Pipeline corrosion can be inhibited by several means.

- Choice of corrosion-resistant metals, alloys
- Injection of corrosion inhibitors
- Cathodic protection
- External and/or internal protective coatings

While corrosion control can be achieved by selection of an appropriate corrosion-resistant metal, operating considerations usually dictate that a high-efficiency corrosion control system be used. Protecting pipelines from corrosion is achieved internally by the injection of inhibitors to mitigate internal corrosion and externally by use of cathodic protection and/or a combination of coatings and cathodic protection (for buried or subsea pipelines).

3.8.2.1 Choice of Corrosion-Resistant Metals

Corrosion resistance is a basic property related to the ease with which materials react with a given environment. All metals have a tendency to return to stable conditions. This tendency causes metals to be classified according to rising nobleness, which again leads to classification of decreasing activity and increasing potential. For specific recommendations concerning materials selection, refer to the ANSI B31.3 and B31.8, API RP 14E (1991) and NACE (1975) MR-01-75. Corrosion resistance is not the only property to be considered in the material selection process, and the final selection will generally be the result of several compromises between corrosion resistance and economic factors. Historically, high-strength steels and alloys[1] have been the safest and most economical materials for the construction of offshore transmission pipelines. However, the inherent lack of corrosion resistance of these materials in subsea pipelines requires a corrosion control system with a high degree of reliability (Hartt and Chu, 2004).

3.8.2.2 Corrosion Inhibitors

For internal corrosion protection, chemical corrosion inhibitors are used. Corrosion inhibitors are cationic surfactant chemicals, which, when added in a small concentration, effectively reduce the corrosion rate of a metal. Inhibitors, which are normally distributed from a solution or dispersion, adsorb on the metallic surface and form a corrosion-resistant film. One major advantage of the corrosion inhibitors is that they are able to respond to an increase in the corrosive environment in the pipelines to reduce its effects. Corrosion inhibitors have been classified differently by various researchers. The most common inhibitor classification schemes are based on their chemical functionality. For application in petroleum production operations, organic inhibitors containing nitrogen (amines) dominate because of their high effectiveness and availability.

Corrosion inhibitors can be applied in batches, resulting in a protective film, which may last for weeks or months. Enough inhibitor should be introduced to provide an appropriate wall coating thickness. Frequency of the treatment is governed by the remaining effectiveness of the inhibitor after a specified amount of commodity has been moved through the line. Inhibitors are often continuously injected in low concentrations where

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[1] Corrosion-resistant alloys, such as 13% Cr steel and duplex stainless steel, are often used downhole and, recently, for short flow lines; however, for long-distance, large-diameter pipelines, carbon steel is the only economically feasible material (Nyborg, 2003).

a thin film is laid down and maintained over time (Tiratsoo, 1992). Sometimes a combination of continuous and batch treatment is employed with a periodic large concentrated slug followed by a continuous treatment of only a few parts per million to maintain a film.

Although previous studies have focused on the problem of corrosion based on test results in small-scale laboratory systems, a considerable amount of effort has been directed towards the development of a more environmentally friendly approach for corrosion control in practical multiphase pipelines carrying gas and condensate. The technique involves increasing the pH of the water phase using pH stabilization agents such as sodium hydroxide so as to facilitate the formation of a dense iron carbonate film on the pipeline wall (IFE, 2000). However, such carbonate films may not hold under slugging conditions (Kang *et al.*, 1996) and are not appropriate for brines containing significant concentrations of chloride and some other ions.

3.8.2.3 Cathodic Protection

Cathodic protection is the most successful method for reducing or eliminating corrosion for buried or submerged metallic structures that involves using electric voltage to prevent corrosion (Colson and Moriber, 1997). When two metals are connected to each other electrically in an electrolyte (e.g., seawater), electrons will flow from the more active metal (anode) to the other (cathode) due to the difference in the electrochemical potential. The anode supplies current, and it will gradually dissolve into ions in the electrolyte and at same time produce electrons that the cathode will receive through the metallic connection with the anode. The result is that the cathode will be negatively polarized and hence be protected against corrosion. The two methods of achieving cathodic protection are (1) the use of sacrificial or reactive anodes with a corrosion potential lower than the metal to be protected and (2) applying a direct current (Uhlig and Revie, 1985). Use of a direct current system is less costly than sacrificial anodes and provides a higher range of possible potential differences (Copp, 1970), although they may require greater maintenance during the lifetime of the operation.

3.8.2.4 Protective Coatings

While cathodic protection has historically been employed as the sole corrosion control methodology for subsea gas production systems, the nature of multiphase pipelines is such that the combined use of protective coatings with cathodic protection is necessary to achieve the effective protection[2] (Samant, 2003). In fact, protective coatings help control pipeline corrosion by providing a barrier against reactants such as oxygen and water. However, because all organic coatings are semipermeable to oxygen and water, coatings alone cannot prevent corrosion and so a combination of cathodic protection is often used (Varughese, 1993).

Pipelines are often internally coated to minimize corrosion when lying in pipeline dumps prior to construction and to provide a smooth surface that reduces friction when fluids are in transit. It should be noted that the initial period in a production well's life can be its most corrosive time due to the high partial pressure of the carbon dioxide. Therefore, it may be more economical to protect the multiphase pipelines from a young field. The coating must be compatible with the commodity and should have suitable resistance to attack by the commodity as well as any contaminants, corrosives, or inhibitor associated with it. Coatings such as epoxies, plastics, or metallic compounds can be used for selected applications. The oldest and still used approach consists of hot-applied bituminous material wrapped with an appropriate covering (Polignano, 1982). External coatings are used to reduce the value of the external current to an economic level by imposing a barrier of high electrical resistance between metal and its environment (Copp, 1970). Rhodes (1982) reports the use of many such coatings, including polyurethane, phenolic resin, phenolic epoxy, and fusion-bonded epoxy powder.

Pipelines are susceptible to both internal and external corrosion. Therefore, both internal and external monitoring of a pipeline is required for a complete assessment, thus providing the direction that will ensure proper utilization of materials and corrosion control methodologies. While traditional inspection and monitoring techniques may represent effective solutions for assessment of the condition and integrity of a pipeline, the sensitivity and accuracy of these methods may be inadequate for monitoring inhibitor performance. However, the field signature method (FSM) developed by the Center for Industrial Research (Strommen *et al.*, 1993) continuously monitors corrosion taking place at a given point on the subsea pipeline and remaining pipe wall thickness at any time. This information can be transferred immediately to the surface, allowing the

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[2] For some installations (e.g., deepwater) one might choose continuous inhibition over protective coating due to the implications of a coating failure.

operator to take immediate action. Compared with traditional corrosion-monitoring methods, the FSM reduces costs, improves safety, and may reduce the required frequency of intelligent pigging. The method requires virtually no maintenance or replacement of consumables. Its service life equals that of the pipe itself. However, it is still a relatively new technique and experience is improving.

3.8.3 Wax

Multiphase flow can be severely affected by the deposition of organic solids, usually in the form of wax crystals, and their potential to disrupt production due to deposition in the production/transmission systems. The wax crystals reduce the effective cross-sectional area of the pipe and increase the pipeline roughness, which results in an increase in pressure drop. The deposits also cause subsurface and surface equipment plugging and malfunction, especially when oil mixtures are transported across Arctic regions or through cold oceans. Wax deposition leads to more frequent and risky pigging requirements in pipelines. If the wax deposits get too thick, they often reduce the capacity of the pipeline and cause the pigs to get stuck. Wax deposition in well tubings and process equipment may lead to more frequent shutdowns and operational problems.

3.8.3.1 Wax Deposition

Precipitation of wax from petroleum fluids is considered to be a thermodynamic molecular saturation phenomenon. Paraffin wax molecules are initially dissolved in a chaotic molecular state in the fluid. At some thermodynamic state the fluid becomes saturated with the wax molecules, which then begin to precipitate. This thermodynamic state is called the onset of wax precipitation or solidification. It is analogous to the usual dew point or condensation phenomenon, except that in wax precipitation a solid is precipitating from a liquid, whereas in condensation a liquid is precipitating from a vapor. In wax precipitation, resin and asphaltene micelles behave like heavy molecules. When their kinetic energy is sufficiently reduced due to cooling, they precipitate out of solution but they are not destroyed. If kinetic energy in the form of heat is supplied to the system, these micelles will desegregate and go back into stable suspension and Brownian motion.

3.8.3.1.1 Wax Deposition Envelope Many reservoir fluids at some frequently encountered field conditions precipitate field waxes. It is very important to differentiate field waxes from paraffin waxes. Field waxes usually consist of a mixture of heavy hydrocarbons such as asphaltenes, resins, paraffins (or paraffin waxes), cyclo-paraffins, and heavy aromatics. Wax precipitation depends primarily on fluid temperature and composition and is dominated by van der Waals or London dispersion type of molecular interactions. Pressure has a smaller effect on wax precipitation. As with asphaltenes, the fact that waxes precipitate at some and not at other thermodynamic states, for a given fluid, indicates that there is a portion of the thermodynamic space that is enclosed by some boundary within which waxes precipitate. This bounded thermodynamic space has been given the name wax deposition envelope (WDE) (Leontaritis, 1995). A typical WDE is shown in Figure 3-17.

The upper WDE boundary can have either a positive slope or a negative slope. In most cases where the WDE has been obtained experimentally, the upper WDE boundary was very close to a vertical line.

Intersection of the WDE boundary with the bubble-point line is generally expected to be to the left of the onset of wax crystallization (cloud

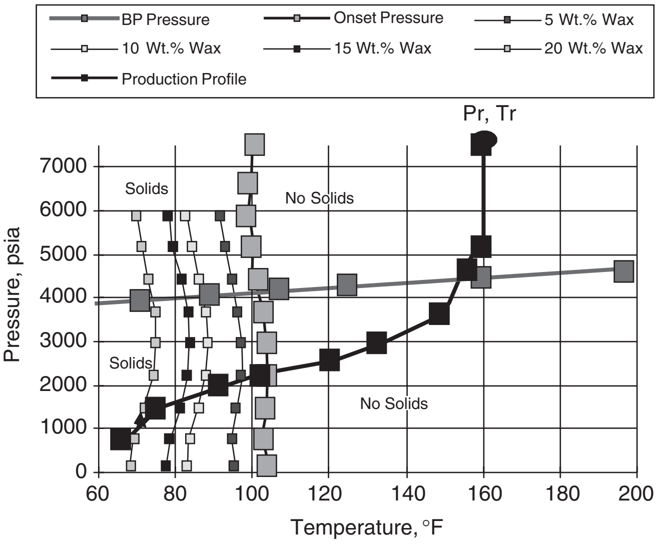

Figure 3-17. Typical wax deposition envelope (Leontaritis, 1995).

point) of the stock tank oil. This has been the case with many reservoir fluids whose WDE has been measured (Leontaritis, 1995). This is due to the fact that light ends, when pressured into oil, usually cause a suppression of the onset of wax crystallization temperature. The actual shape of the lower WDE boundary is primarily a function of the compositions of the intermediates and light ends of the reservoir fluid. The technology for measuring WDEs has been perfected only recently. Because the technology is new and the cost of measurements is relatively high, most companies do not find it economical to obtain the complete WDE of their reservoir fluid through laboratory measurements. Hence, obtaining only a few experimental data points and using them to fine-tune phase behavior models that are then used to calculate the remaining WDE more economically is preferred. It should be noted that most of these models have very little predictive capacity at this time and are used primarily as combination correlational-predictive type of tools. One of the main reasons for the lack of pure predictive capability is improper or inadequate crude oil characterization (Leontaritis, 1997a).

3.8.3.1.2 Gas/Condensate Wax Deposition Envelope Some gas/condensates, especially rich gas/condensates with yields in excess of 50 bbls/MMSCF, are known to contain high carbon number paraffins that sometimes crystallize and deposit in the production facilities. The obvious question is what is the shape of the thermodynamic envelope (i.e., P and T surface) of these gas/condensates within which waxes crystallize or, in order to maintain the previous terminology, what is the WDE of gas/condensates typically?

The shapes of the WDEs of two gas/condensates in the Gulf of Mexico are presented here. The shapes of the aforementioned WDEs indicate potential wax deposition in those cases where the gas/condensate contains very high carbon number paraffins that precipitate in solid state at reservoir temperature. In other words, the temperature of the reservoir may not be high enough to keep the precipitating waxes in liquid state. Hence, the gas/condensate, which is a supercritical fluid, enters the WDE at the "dew point" pressure. This casts new insight into the conventional explanation that the productivity loss in gas/condensate reservoirs, when the pressure near the wellbore reaches the dew point, is only due to relative permeability effects.

Figure 3-18 shows the vapor–liquid (V-L) envelope of what one might call a typical Gulf of Mexico gas/condensate. This gas/condensate

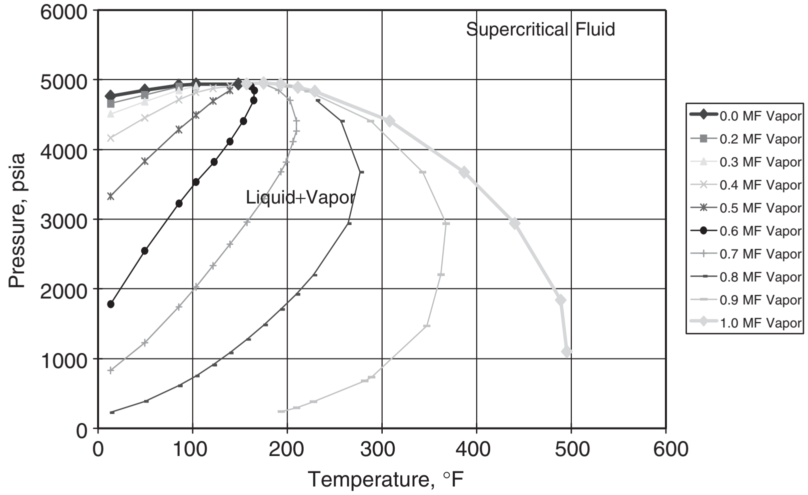

Figure 3-18. Vapor–liquid envelope, Gulf of Mexico gas/condensate "A" (Leontaritis, 1998).

(called gas/condensate "A" for our purposes here) was analyzed with PARA (paraffin-aromatic-resin-asphaltene) analysis (Leontaritis, 1997a) and found to contain normal paraffins with carbon numbers exceeding 45. The V-L envelope was simulated using the Peng and Robinson (1976) original equation of state (EOS) that had been fine-tuned to PVT data obtained in a standard gas/condensate PVT study. The first question that was addressed in a wax study involving this fluid was what happens as the fluid is cooled at some constant supercritical pressure? What actually happened is shown in Figure 3-19.

Figure 3-19 shows several onset of wax crystallization data points obtained with near-infrared (NIR) equipment (Leontaritis, 1997b) by cooling the gascondensate "A" at different constant pressures. It was evident from NIR data that there was a thermodynamic envelope, similar to the one defined and obtained experimentally for oils, to the left of which (i.e., at lower temperatures) wax crystallization occurred. The complete wax deposition envelope shown in Figure 3-19 was calculated with a previously tuned wax phase behavior model (Narayanan *et al.*, 1993). Despite the clarity of the WDE obtained for gas/condensate "A" as shown in Figure 3-19, more data were needed to confirm the presence of WDE in other condensates and establish its existence as a standard thermodynamic diagram.

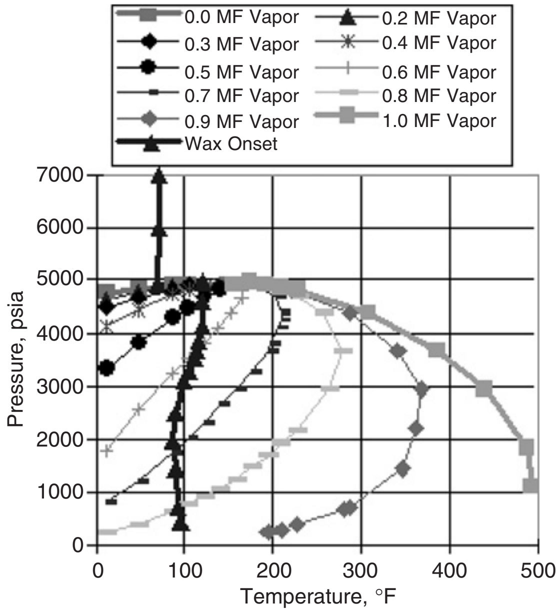

Figure 3-19. Wax deposition envelope, Gulf of Mexico gas/condensate "A" (Leontaritis, 1998).

Figure 3-20 shows the V-L envelope of another typical Gulf of Mexico condensate. This condensate (called gas/condensate "B" for our purposes here) also contains paraffins with carbon numbers exceeding 45, although data show that gas/condensate "B" is lighter than gas/condensate "A". The V-L envelope was again simulated using the Peng and Robinson (1976) original EOS after it had been tuned to PVT data obtained in a standard gas/condensate PVT study. Figure 3-21 shows NIR onset data superimposed on the V-L envelope.

It is evident again from NIR data that there is a thermodynamic envelope to the left of which (i.e., at lower temperatures) wax crystallization occurs. Once again, the complete wax deposition envelope shown in Figure 3-21 was calculated with a previously tuned wax phase behavior model (Narayanan *et al.*, 1993).

Data presented here confirm the presence of a WDE in gas condensates that contain high carbon number paraffin waxes (≥ 45). This WDE is similar to oil WDEs and as a result it should be considered a standard thermodynamic diagram. The shape of the WDE inside the V-L envelope

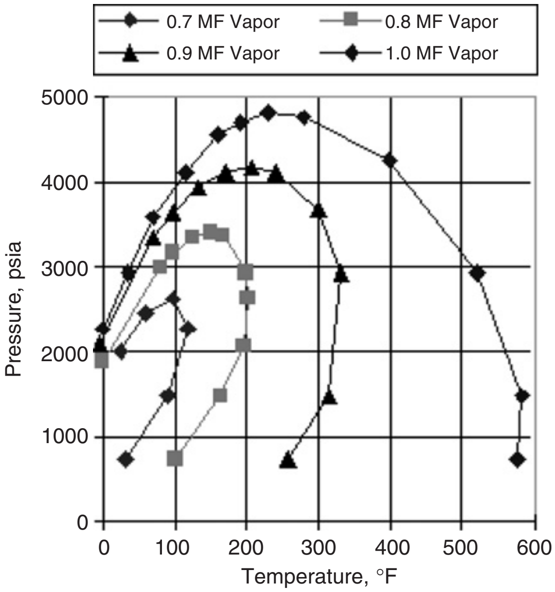

Figure 3-20. Vapor–liquid envelope, Gulf of Mexico gas/condensate "B" (Leontaritis, 1998).

seems to be consistent with existing information regarding the effect of light hydrocarbons on the onset of wax crystallization or wax appearance temperature. That is, as the pressure rises, the WDE tilts to the left (negative slope) due to the ability of light hydrocarbons to depress wax crystallization. However, at the pressure where retrograde condensation begins the WDE turns forward, thus acquiring a positive slope. This is because the light ends begin to vaporize and waxes remaining in the liquid phase begin to concentrate. This is simply caused by the change in normal paraffin concentration, which in turn is caused by retrograde concentration. In most condensates the V-L envelope is fairly horizontal at the saturation line (dew point or bubble point). Hence, when this general pressure is reached the WDE seems to coincide with the V-L saturation line until the temperature becomes low enough for the waxes to begin crystallizing from the supercritical condensate. This is in agreement with prior observations that indicate a substantial increase in the solvent power of some fluids when they become supercritical (i.e., propane, CO_2). That is, supercritical hydrocarbon fluids are expected to require cooling to much

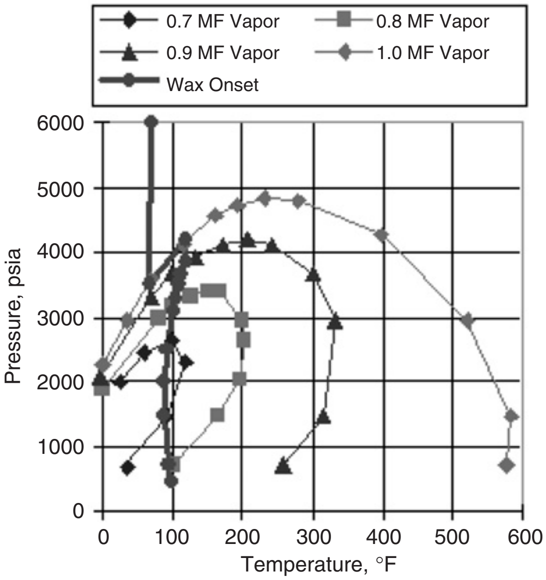

Figure 3-21. Wax deposition envelope, Gulf of Mexico gas/condensate "B" (Leontaritis, 1998).

lower temperatures before paraffin waxes begin to crystallize because of their increased solvent power.

3.8.3.2 Wax Formation in Multiphase Gas/Condensate Pipelines

When the production pressure and temperature profile of a gas/condensate crosses the WDE and the hydrate envelope (HE), waxes and hydrates may form. The formed hydrates can grow in mass and yield strength to the point that they restrict and finally stop the flow. If the waxes form while the liquid is in contact with the wall, some of them should attach to the wall, thus resulting in wax deposition. Figure 3-22 shows a P-T diagram of a Gulf of Mexico gas/condensate demonstrating this situation.

The diagram indicates that there was not any live NIR done with this fluid. The only point on the WDE that was measured was the cloud point of the condensate, shown in Figure 3-22 at 95°F at the bottom of the vertical line. The clean tubing and flow line of this system would deliver about 14.5 MMSCFD on 6/17/2002. From the light blue P and T profile line and dark blue hydrate line, it was very clear at the outset that the

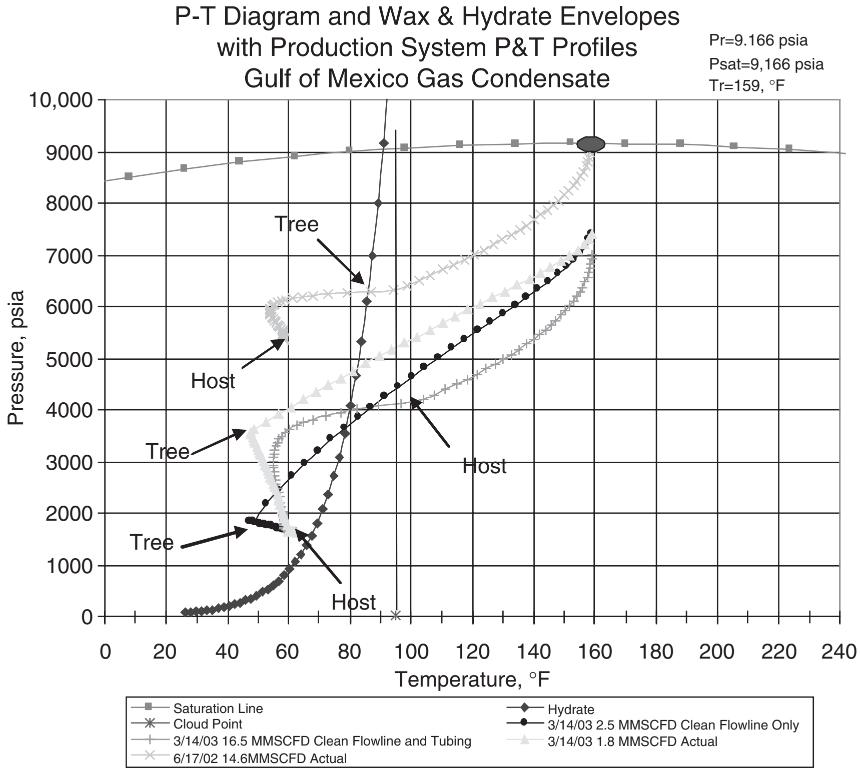

Figure 3-22. Gulf of Mexico gas/condensate wax and hydrate deposition envelopes (AsphWax, Inc., 2003).

main flow assurance issues in this system would be hydrate and wax formation and deposition. Hence, because of this illustrative important information, the effort to obtain the aforementioned diagram is obviously very worthwhile.

Sometimes oil and gas operators decide to design the facilities such that their operation is to the left of the WDE and HE (such as the example in Figure 3-22). In these situations, wax and hydrate formation takes place if the fluids are left untreated with chemicals. In the aforementioned example, methanol was injected to inhibit successfully the formation of hydrates formed due to the production of reservoir equilibrium water. However, because the operator was not aware of the wax phase behavior of this condensate at start-up, no wax chemical was injected. Wax deposition was severe enough to cause the production rate to drop to 1.8 MMSCFD on 3/14/2003. The estimated via simulation maximum production rate on 3/14/2003 was 16.5 MMSCFD. Data in the plot show that the maximum friction loss was in the upper part of the tubing.

3.8.3.2.1 Identification of Wax Deposition Problems A rather simple chart that allows daily monitoring of wax deposition problems in multiphase gas/condensate pipelines is shown in Figure 3-23. This chart is for the gas/condensate shown in Figure 3-22. The chart was made by plotting the performance index (PI) versus time. The PI is calculated from the following equation:

$$\mathrm{PI} = \frac{\Delta P^2}{Q^{\frac{1}{0.54}}} \qquad (3\text{-}57)$$

where ΔP is the pressure drop in the pipeline, psi, and Q is the flow rate, MMSCFD.

The PI should remain as a horizontal straight line during production if there is no restriction formed in the line. It is evident in Figure 3-23

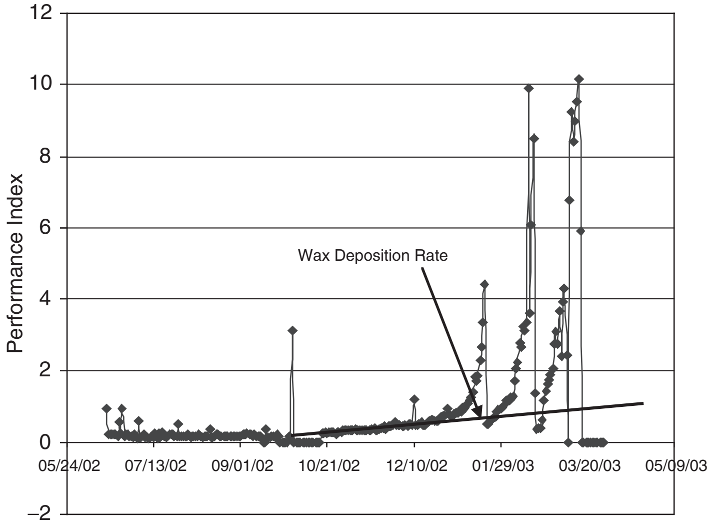

Figure 3-23. Gas/condensate flowline performance index (AsphWax, Inc., 2003).

that around 10/21/2002 a restriction was being formed in the line. At the beginning of January there was a very sharp loss of hydraulic capacity. This cannot be due to wax deposition because the rate of wax deposition does not change so sharply in a produced fluid. Indeed, the sharp rise in the PI was caused by hydrate formation that occurred on top of wax crystal formation and deposition. Several sharp rises in PI followed caused by hydrates until it was determined that the well started producing free water and required a much higher methanol injection rate. While all of this is happening with the advent of water and hydrate formation the wax deposition rate was increasing steadily, as shown on the chart.

It should be noted that the engineer needs to be very careful in attributing a rise in the PI only to wax and hydrates. One needs to make sure that other culprits, such as fines, salts, returning drilling, and completion fluids, are not present.

3.8.3.2.2 Wax Deposition Inhibition/Prevention It is not always clear to the engineer what would be the wax deposition mechanism and what type of chemical treatment would be the most effective in gas/condensate situations. In oil flow lines, wax deposition occurs by diffusion of wax molecules and crystals toward and attachment at the wall. Prevention or inhibition of wax deposition is mainly accomplished by injection of a special class of molecules that interact with paraffin molecules at temperatures above the cloud point and influence their crystallization process in a way that diminishes the attraction of the formed crystals toward the wall. The inhibited formed wax crystals are removed from the system by the shear forces caused by the flowing oil.

The aforementioned mechanism occurs when the wax crystal forms at the wall. This is the case with most liquid-filled bare (uninsulated) lines. The two important requirements are fast heat transfer and fast diffusion of paraffin molecules toward the wall. This is not the case in many situations of flow lines carrying gas/condensates. There is fast cooling but the majority of the gas cools down while in the main flow, thus forming wax crystals with the main flow that would have a tendency to deposit or sediment by gravity in the liquid holdup. Also, the liquid that forms near the inlet of the bare line that has not reached its cloud point cools with a velocity that is almost standstill compared to the main flow. Slugging in this case would be beneficial in removing wax slush from the line. Additionally, these flow lines are essentially soaked in methanol to prevent hydrate formation. It just so happens that most (if not all) chemicals with wax crystal modification properties, as discussed later, are incompatible with alcohols and glycols. Hence, the inhibited wax crystals accumulate in the flowline in the liquid holdup, thus adding to the wax accumulation. This accumulation forms what has been called “wax slush.” This wax slush is viscous but it moves and can flow given enough shear force.

The aforementioned theory has been documented several times from field data. In the year 2000 enough production and laboratory data and simulations were available to make the case for one field, but there was no direct evidence of wax slush from the field. During a shutdown caused by a hurricane, the flowline was depressurized at the tree in an effort to dissolve any existing hydrates. At start-up, the well was started very fast to try to initiate movement and flow of any liquids (via an induced slug) and the previously theorized existing “wax slush.” The fast flow did mobilize a phase that was loaded with wax. The wax slush flowed into the separators and plugged lines and equipment, making them temporarily incapable of operation.

It is appropriate at this time to give a brief description of the main two chemical classes used to treat wax deposition. There are three types of wax crystals:

- Plate crystals
- Needle crystals
- Mal or amorphous crystals

Paraffinic oils form plate or needle crystals. Asphaltenic oils form primarily mal or amorphous crystals. Asphaltenes act as nucleation sites for wax crystal growth into mal crystals.

Plate crystals look as their name implies, like plates under the microscope. Needle crystals look like needles, and mal crystals are amorphous and generally look like small round spheres. The interaction between crystals and pipe wall increases from mal to needle to plates. Thus, maintaining newly formed wax crystals small and round, i.e., like mal crystals, is desirable.

The behavior and properties, e.g., cloud point and pour point, of paraffin crystals precipitating from a hydrocarbon can be affected in three ways.

1. Crystal size modification: modification of the crystal from larger sizes to smaller sizes.

2. Nucleation inhibition: inhibition of the growth rate of the crystal and its ultimate size.
3. Crystal type or structure modification: modification of the crystal from one type to the other. For instance, modify a crystal from needle to mal type.

A wax crystal modifier works primarily to modify the crystal size. The plate crystals of an *n*-paraffin look much smaller under the microscope when precipitating from hydrocarbons inhibited with a wax crystal modifier. Smaller crystals have lower molecular weights and thus higher solubility in oil. Furthermore, smaller particles have smaller energy of interaction among themselves and the pipe wall. A crystal modifier interrupts the normal crystal growth of the *n*-paraffins by inserting itself in the crystal, thus stopping its growth.

It is noted that another name for a wax crystal modifier is pour point depressant (PPD). As the name implies, wax crystal modifiers are very effective at suppressing the pour points of crude oils because they suppress the wax crystal growth, thus minimizing the strength of their interactions.

A wax dispersant may act to inhibit wax nucleation and change the type of wax crystals from plate or needle to mal or amorphous. Inhibited amorphous crystals are smaller and carried much easier by the hydrodynamic drag of the flowing fluid. The presence of asphaltenes and resins facilitates the effect of the dispersant. The dispersant interacts with the asphaltenes and resins and ties them up, thus removing nucleation sites required by wax crystals to grow. A dispersant, when added to the oil at a temperature above the cloud point, has occasionally an additional benefit of suppressing the cloud point by interacting with and tying up the asphaltenes and resins that tend to come out first. A dispersant tends to usually disperse the wax particles at the water–oil interface.

Both wax crystal modifiers and wax dispersants are useful chemicals that have the ability to diminish wax formation and deposition, although through different mechanisms. Wax dispersants are usually much smaller in molecular weight and size than wax crystal modifiers. Hence, their viscosity and flow properties in general are more favorable in cold applications. Also, some wax dispersants are soluble with alcohols and glycols, thus making them compatible for simultaneous injection. The selection of the wax chemical should be made after careful consideration of the produced hydrocarbon and facilities.

3.8.3.2.3 Wax Deposit Remediation On occasion, if a substantial amount of wax accumulates in the line, as evidenced by the PI chart such as the one shown in Figure 3-23, a temporary shutdown to do a chemical soak, with a potential modification to the chemical to give it more penetrating power at cold temperatures, and a fast start-up might be necessary to cause the wax slush cough.

Pigging is an option, but only after very careful consideration of the system's performance to understand the dynamics of the moving pig and continuous removal of the wax cuttings ahead of it. This is a very difficult job. Controlling the bypass of flow around the pig for such a purpose is difficult. Hence, many pigging operations end up in failure with stuck pigs. The pigging analysis and decision must be left to true experts. Starting a pigging program at the beginning of the life of the system has a better chance of success than at any other time. Even then, excellent monitoring of the system's PI is a must. It is recommended that a short shutdown, chemical soak, and fast start-up be considered first, because it is the safest option.

3.8.3.2.4 Controlled Production of Wax Deposits From a technical standpoint, spending enough capital initially to design the facilities to operate outside the wax and hydrate forming conditions or to the right of the WDE and HE is obviously the best solution. However, very often in practice, controlling wax deposits during production is preferred because of the lower facilities cost. Economically marginal fields can only be produced under this scenario. Hence, in these cases, the following three options prevail.

1. Pigging only
2. Chemical injection only
3. Combination of pigging and chemical injection

While frequent pigging of the line clears up any wax deposits, it may be necessary to chemically inhibit wax deposition during times when pigging is unavailable. Pigging often is inadequate or uneconomical unless used in conjunction with a chemical treatment program. This program is often performed into two stages: (1) removal of wax deposits in the production/transmission lines and (2) continuous chemical injection or periodic treatment (such as batch treatments, etc.) in order to ensure pipeline integrity.

Chemical injection is the safest of the three if good technical support and testing are available. The approach is discussed as follows.

1. Inject a strong chemical dispersant/inhibitor down the hole to keep the formed wax particles small and suspended in the flow line. The majority of them would be carried away with the gas/liquid flow.
2. The chemical dispersant/inhibitor must be compatible and soluble with methanol to prevent precipitation of the chemical itself in the line.
3. Monitor the hydraulics in the flow line and occasionally, if necessary, cause a "slug" or "cough" to cough up any accumulated wax.
4. On occasion, if a substantial amount of wax accumulates in the line, a temporary shutdown to do a chemical soak, with a modification to the chemical to give it more penetrating power at cold temperatures, and a fast start-up might be necessary to cause the wax slush cough.

When implemented properly, the aforementioned approach should provide the lowest cost wax deposition control in gas/condensate transmission lines.

3.8.4 Severe Slugging

One of the most complex multiphase flow patterns with unsteady characteristics is intermittent or slug flow. Slug flow exists in the whole range of pipe inclinations and over a wide range of fluid flow rates. However, liquid slugs may, depending on the pipeline geometry and flow conditions, become very large (severe slugging) and threaten the safe and reliable operation of a production system. Severe slugging can occur in a multiphase flow system, operating at low liquid and gas flow rate, where a downward inclined pipe segment is followed by an upward inclined segment/riser. For this system, liquid will accumulate in the riser and the pipeline, blocking the flow passage for gas flow. This results in a compression and pressure buildup in the gas phase that will eventually push the liquid slug up the riser and a large liquid volume will be produced into the separator that might cause possible overflow and shut down of the separator. The severe slugging phenomenon is very undesirable due to pressure and flow rate fluctuations, resulting in unwanted flaring and reducing the operating capacity of the separation and compression units (Schmidt *et al.*, 1980). Figures 3-24 and 3-25 show example time traces

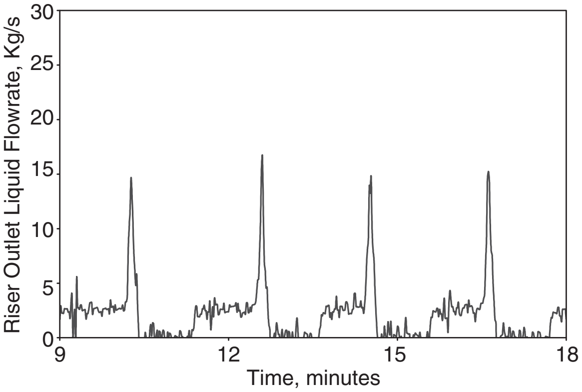

Figure 3-24. Example time trace of the riser outlet liquid flow rate during severe slugging (Mokhatab, 2006b).

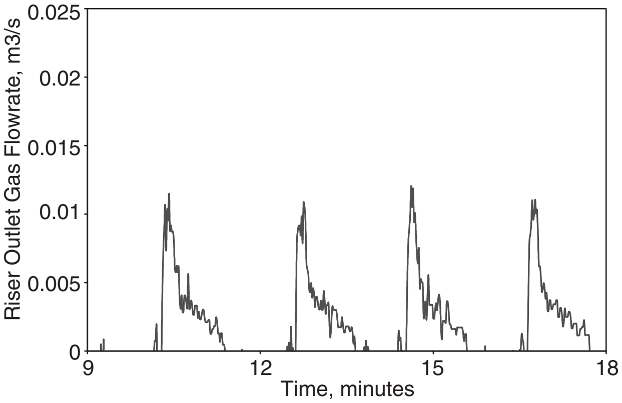

Figure 3-25. Example time trace of the riser outlet gas flow rate during severe slugging (Mokhatab, 2006b).

for the riser outlet liquid and gas flow rates during severe slugging, respectively. These figures show the large surges in liquid and gas flow rates accompanying the severe slugging phenomenon. Clearly such large transient variations could present difficulties for topside facilities unless they are designed to accommodate them.

Usually, the operators try not to operate in the severe slugging region. However, the inlet conditions of a production pipeline are linked to the number and the capacity of the producing wells, the availability of wells, and also to some undesirable operation such as shut down or restart. The natural trend when dimensioning a production line is to do whatever possible to avoid critical flooding of the separator, and therefore to overdimension the separator unit. However, in offshore production, overdimensioning the installation is very costly and not always possible so the design engineers require more accurate dynamic simulations to correctly design and dimension their production schemes (which are more sensitive to transients occurring when slug flow conditions buildup and require high-performance control systems to maintain the plant within the preset operating ranges) and to be able to propose new concepts suitable to every situation they can be faced with (Sagatun, 2004). The prediction of pipeline simulation with respect to slug dynamics will be strongly affected by the specification of boundary conditions. Hence, a tight integration of the hydrodynamic model of pipeline-riser system with the dynamic model of the receiving plant is very important for achieving correct simulation results, especially for studying severe slugging dynamics (Mokhatab, 2006c).

Given these potential problems of severe slugging in such systems, an understanding of how and when severe slugging will form, as well as providing significant information on the best method to prevent and control this phenomenon, is a highly necessary undertaking.

Although the research community has undertaken a thorough exposition of severe slugging phenomenon in flexible risers, there is a lack of sufficiently wide-ranging and openly available transient code testing and full-scale data on flexible risers for better understanding of the physics and characteristics of severe slugging and carrying out more vigorous verification (Mokhatab, 2006d). Unless such a vigorous link with reality is maintained, designers and operators will not have the confidence to expand the use of flexible risers in more critical applications.

3.8.4.1 Severe Slugging Mechanism

The process of severe slugging in a pipeline-riser system consists of four steps: (1) slug formation, (2) slug production, (3) bubble penetration, and (4) gas blow down. This phenomenon had been identified previously by Schmidt *et al*. (1980) as a cyclic flow rate variation, resulting in periods of both no flow and very high flow rates substantially greater than the time average. Figure 3-26 illustrates the stages of a severe slugging cycle.

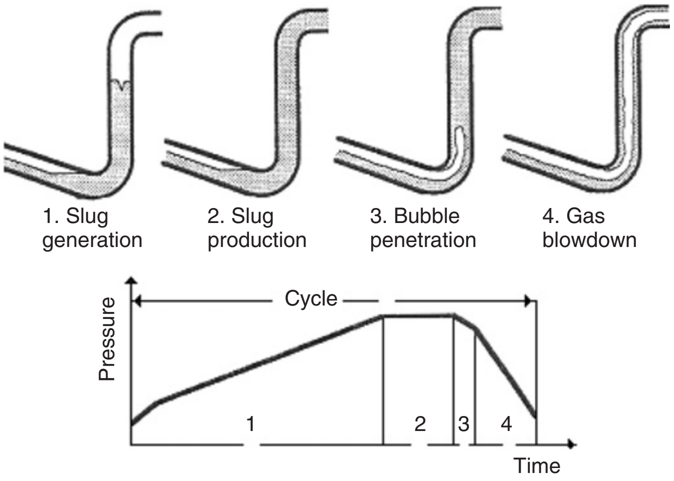

Figure 3-26. Description of severe slugging in pipeline/riser systems (Fabre *et al.*, 1990).

The first step, slug formation, corresponds to an increase of the pressure in bottom of the riser. The liquid level does not reach the top of the riser. During this period, the liquid is no longer supported by the gas and begins to fall, resulting in riser entrance blockage and pipeline pressure buildup until the liquid level in the riser reaches the top. During the second step, slug production, the liquid level reaches the riser outlet, and the liquid slug begins to be produced until the gas reaches the riser base. In the third step, bubble penetration, gas is again supplied to the riser so the hydrostatic pressure decreases. As a result, the gas flow rate increases. The fourth step corresponds to gas blow down. When the gas produced at the riser bottom reaches the top, the pressure is minimal and the liquid is no longer gas lifted. The liquid level falls and a new cycle begins (Fabre *et al.*, 1990). This cyclic process becomes steady state when the rate of penetration of the gas into the riser is always positive. However, it is also possible that penetration of the gas into the riser becomes zero. In this case, liquid blocks the bottom of the riser. This is followed by a movement of the liquid interface into the pipeline and blocking of the gas passage into the riser until the liquid interface reaches the bottom of the riser. At this point, penetration of gas into the riser starts and a new cycle begins again.

When liquid penetrates into the pipeline, the gas in the riser propagates to the top until all of the gas in the riser disappears. When the liquid input is very low, the propagation of gas toward the top of the riser causes accumulation of all the gas at the top as the liquid falls back. This process is termed cyclic process with fallback, while the former case is termed cyclic process without fallback. In summary, three different possibilities can occur as a result of penetration of gas into a liquid column in a quasi-steady severe slugging process (Jansen *et al.*, 1996).

1. Penetration of the gas that leads to oscillation, ending in a stable steady-state flow.
2. Penetration of the gas that leads to a cyclic operation without fallback of liquid.
3. Penetration of the gas that leads to a cyclic operation with fallback of liquid.

Severe slugging in a pipeline-riser system can be considered a special case of flow in low-velocity hilly terrain pipelines, which are often encountered in an offshore field. This is a simple case of only one downward inclined section (pipeline), one riser, and constant separator pressure. For this reason, severe slugging has been termed “terrain-induced slugging” (Fuchs, 1987). This phenomenon has also various names in the industry, including “riser-base slugging” and “riser-induced slugging.”

3.8.4.2 Stability Analysis

The flow characteristics of multiphase flow in a pipeline-riser system are divided into two main regions: stable (steady flow) and unstable (pressure cycling), in which the stability line on the flow pattern map separates the two regions. The steady region includes acceptable slugging, annular mist, and bubble flows, whereas the pressure cycling region includes severe slugging and transitional flows. Figure 3-27 shows a typical flow map for a pipeline-riser system developed by Griffith and Wallis (1961) featuring regions of stable and unstable behavior. In Figure 3-27, N_{Fr} and λ_G are the Froude number of two-phase gas/liquid flow and no-slip gas holdup, defined previously. As it can be seen from Figure 3-27, at low Froude numbers, bubble flow prevails and fluids will flow through riser pipes without slug formation. However, as the Froude number increases, the slug flow range is entered.

The stability analysis predicts the boundary between stable and unstable regions, where the resultant stability map helps the engineer to design

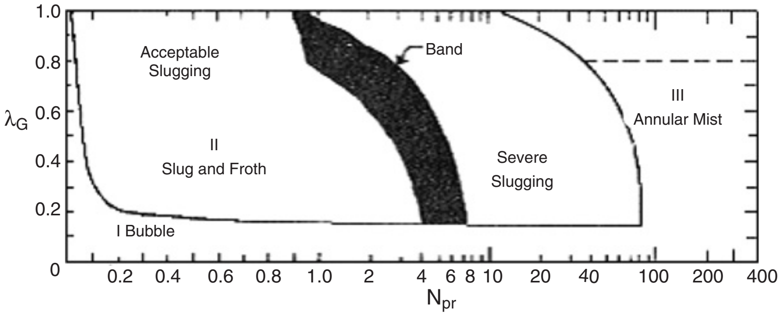

Figure 3-27. Griffith and Wallis flow pattern map with Yocum (1973) transition band (Brill and Beggs, 1991).

systems that operate well into the stable zone, thus offering an adequate margin of safety. The stability analysis seeks to model a particular process required for severe slugging and hence predicts the likelihood of severe slugging, as such these stability models are termed criteria for severe slugging. The severe slugging process was first modeled by Schmidt *et al*. (1980); however, their model formed the basis of much of the early work for the stability of severe slugging. A review of the existing stability criteria for predicting severe slugging in a pipeline-riser system can be found in Mokhatab (2005).

3.8.4.3 Prevention and Control of Severe Slugging

The flow and pressure oscillations due to severe slugging phenomenon have several undesirable effects on downstream topside facilities unless they are designed to accommodate them. However, designing topside facilities to accept these transients may dictate large and expensive slug catchers with compression systems equipped with fast-responding control systems. This may not be cost-effective and it may be more prudent to design the system to operate in a stable manner (Sarica and Tengesdal, 2000). While lowering production rates (slowing fluid velocity) can minimize severe slugging, operators are investigating alternatives that would allow for maximum production rates without the interruptions caused by slugs (Furlow, 2000). Mokhatab *et al*. (2006a) reference a combination of industrial experience and information from the literature to compile a list of methods of remediating the problems associated with severe slugging in pipeline-riser systems. In general, severe slugging prevention and elimination strategies seek the three following approaches.

3.8.4.3.1 Riser Base Gas Injection This method provides artificial lift for the liquids, moving them steadily through the riser. This technique can alleviate the problem of severe slugging by changing the flow regime from slug flow to annular or dispersed flow, but does not help with transient slugging in which the liquid column is already formed before it reaches the riser base. It is one of the most frequently used methods for current applications. However, for deepwater systems, increased frictional pressure loss and Joule–Thomson cooling are potential problems resulting from high-injection gas flow rates.

The riser base gas injection method was first used to control hydrodynamic slugging in vertical risers. However, Schmidt *et al*. (1980) dismissed it as not being economically feasible due to the cost of a compressor or pressurizing the gas for injection and the piping required to transport the gas to the base of the riser. Pots *et al*. (1985) investigated the application of the method to control severe slugging. They concluded that the severity of the cycle was considerably lower for riser injection of about 50% inlet gas flow. Hill (1990) described the riser base gas injection tests performed in the S.E. Forties field to eliminate severe slugging. The gas injection was shown to reduce the extent of the severe slugging.

The riser base gas lift method may cause additional problems due to Joule–Thomson cooling of the injected gas, where the lift gas will cause cooling and make the flow conditions more susceptible for wax precipitation and hydrate formation. Hence, Johal *et al*. (1997) proposed an alternative technique, “multiphase riser base lift” that requires nearby high-capacity multiphase lines diverted to the pipeline-riser system that alleviate the severe slugging problem without exposing the system to other potential problems.

Sarica and Tengesdal (2000) proposed a new technique for sourcing riser base gas lift. The principle of the proposed technique is to connect the riser to the downward inclined segment of the pipeline with a small-diameter conduit, where the conduit will transfer the gas from the downward inclined segment to the riser at points above the riser base (multiposition gas injection). The transfer process reduces both the hydrostatic head in the riser and the pressure in the pipeline, consequently lessening or eliminating the severe slugging occurrence. This method can be considered as self-gas lifting. Sarica and Tengesdal (2000) claimed that the proposed technique is expected to increase production, as it does not impose additional backpressure to the production system. The cost of implementing and operation of the proposed systems in the field application is also expected to be low compared to other elimination methods.

3.8.4.3.2 Topside Choking This method induces bubble flow or normal slug flow in the riser by increasing the effective backpressure at the riser outlet. While a topside choke can keep liquids from overwhelming the system, it cannot provide required control of the gas surges that might be difficult for the downstream system to manage. This is a low-cost, slug mitigation option, but its application might be associated with considerable production deferment.

Topside chocking was one of the first methods proposed for control of the severe slugging phenomenon (Yocum, 1973). Yocum observed that increased backpressure could eliminate severe slugging but would reduce the flow capacity severely. Contrary to Yocum's claim, Schmidt *et al*. (1980) noted that the severe slugging in a pipeline-riser system could be eliminated or minimized by choking at the riser top, causing little or no changes in flow rates and pipeline pressure. Taitel (1986) provided a theoretical explanation for the success of choking to stabilize the flow as described by Schmidt *et al*. (1980).

Jansen (1996) investigated different elimination methods, such as gas-lifting, chocking, and gas-lifting and chocking combination. He proposed the stability and the quasi-equilibrium models for the analysis of the aforementioned elimination methods. He experimentally made three observations: (1) large amounts of injected gas were needed to stabilize the flow with the gas-lifting technique, (2) careful chocking was needed to stabilize the flow with minimal backpressure increase, and (3) the gas-lifting and chocking combination were the best elimination method, reducing the amount of injected gas and the degree of choking to stabilize the flow.

3.8.4.3.3 Control Methods Control methods (feedforward control, slug choking, active feedback control) for slug handling are characterized by the use of process and/or pipeline information to adjust available degrees of freedom (pipeline chokes, pressure, and levels) to reduce or eliminate the effect of slugs in the downstream separation and compression unit. Control-based strategies are designed based on simulations using rigorous multiphase simulators, process knowledge, and iterative procedures. To design efficient control systems, it is therefore advantageous to have an accurate model of the process (Storkaas *et al*., 2001; Bjune *et al*., 2002).

The feedforwarded control aims to detect the buildup of slugs and, accordingly, prepares the separators to receive them, e.g., via feedforwarded control to the separator level and pressure control loops. The aim of slug choking is to avoid overloading the process facilities with liquid or gas. This method makes use of a topside pipeline choke by reducing its opening in the presence of a slug, thereby protecting the downstream equipment (Courbot, 1996). Like slug choking, active feedback control makes use of a topside choke. However, with dynamic feedback control, the approach is to solve the slug problem by stabilizing the multiphase flow. Using feedback control to prevent severe slugging has been proposed by Hedne and Linga (1990) and by other researchers (Henriot *et al*., 1999; Havre *et al*., 2000; Molyneux *et al*., 2000; Havre and Dalsmo, 2001; Bjune *et al*., 2002). The use of feedback control to stabilize an unstable operating point has several advantages. Most importantly, one is able to operate with even, nonoscillatory flow at a pressure drop that would otherwise give severe slugging. Figure 3-28 shows a typical application of an active feedback control approach on a production flowline/pipeline system and illustrates how the system uses pressure and temperature measurements (PT and TT) at the pipeline inlet and outlet to adjust the choke valve. If the pipeline flow measurements (FT) are also available, these can be used to adjust the nominal operating point and tuning parameters of the controller.

Note that the response times of large multiphase chokes are usually too long for such a system to be practical. The slug suppression system (S3) developed by Shell has avoided this problem by separating the fluids into a gas and liquid stream, controlling the liquid level in the separator by throttling the liquid stream and controlling the total volumetric flow rate by throttling the gas stream. Hence, the gas control valve back pressures the separator to suppress surges and as it is a gas choke, it is smaller and therefore more responsive than a multiphase chock.

The S3 is a small separator with dynamically controlled valves at the gas and liquid outlets, positioned between the pipeline outlet and the production separator. The valves of the outlet are regulated by the control system using signals calculated from locally measured parameters, including pressure and liquid level in the S3 vessel and gas and liquid

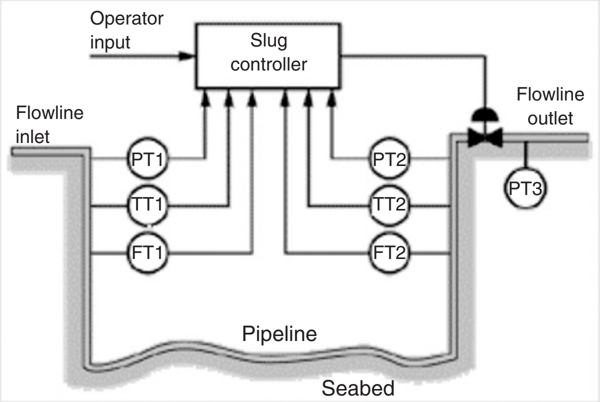

Figure 3-28. Typical configuration of feedback control technique in flow line/riser systems (Bjune *et al.*, 2002).

flow rates. The objective is to maintain constant total volumetric outflow. The system is designed to suppress severe slugging and decelerate transient slugs so that associated fluids can be produced at controlled rates. In fact, implementation of the S3 results in a stabilized gas and liquid production approximating the ideal production system. Installing S3 is a cost-effective modification and has lower capital costs than other slug catchers on production platforms. The slug suppression technology also has two advantages over other slug-mitigation solutions, where unlike a topside choke, the S3 does not cause production deferment and controls gas production, and the S3 controller uses locally measured variables as input variables and is independent of downstream facilities (Kovalev *et al.*, 2003).

The design of stable pipeline-riser systems is particularly important in deepwater fields, as the propensity toward severe slugging is likely to be greater and the associated surges more pronounced at greater water depths. Therefore, system design and methodology used to control or eliminate severe slugging phenomenon become very crucial when considering the safety of the operation and the limited available space on the platform. Currently, there are three basic elimination methods that have been already proposed. However, the applicability of current elimination methods to deepwater systems is very much in question. Anticipating this problem, different techniques should be developed to be suitable for different types of problems and production systems (Mokhatab *et al.*, 2006a).

3.8.5 Real-Time Flow Assurance Monitoring

Significant steps are now being taken to minimize the occurrence of flow assurance problems. In addition to design constraints, flow assurance analysis should lead to operational constraints to include procedures for planned and unplanned shutdowns. In this regard, a real-time flow assurance monitoring system provides optimum asset management by continuously monitoring the operating conditions of the pipeline for anomalous readings, which may indicate the formation of restrictions leading to potential blockage of the asset (i.e., observation of erratic pressure fluctuations is usually indicative of hydrate formation). In fact, reliable real-time data provided quickly and continuously throughout the life cycle of the pipeline can be tied into a software simulation program and analyzed and modeled in a timely fashion. When a process interrupt occurs, the software can project where problems will occur (some can recommend the best corrective action sequence). This faster detection and diagnosis of problems decrease the cost of multiphase production and substantially reduce the risk of environmental disasters due to the failure of unmonitored pipelines.

This monitoring system can provide wide bandwidth capability via fiber optic distributed sensors over which real-time data can pass at optimal transmission rates for analysis, ensuring that data are moved reliably and appropriately referenced at each interval along the pipeline. However, in cases of deep water installations, the need for accurate and advanced distributed sensor systems is heightened further as such systems offer even greater opportunities for the occurrence of flow assurance problems.

3.9 MULTIPHASE PIPELINE OPERATIONS

After a pipeline is installed, efficient operation and automatic controls must be provided to maintain safe operation in the face of unexpected upsets. Leak detection and pigging are typically important procedures.

3.9.1 Leak Detection

Leaks in the pipelines originate from a variety of causes, which may include material-related damage and physical damage caused by construction in the right of way. Because accidental product discharges cannot be eliminated entirely, one of the most effective methods of reducing the impact of spills is to detect the leak quickly and to act quickly to stop the discharge. Methods used to detect product leaks along the pipeline can be divided into two categories: external and internal leak detection systems (API, 1995a). Externally based methods include traditional procedures such as visual inspection, as well as technologies such as hydrocarbon sensing using fiber optic or dielectric cables (Sandberg *et al*., 1989). Internally based methods, also known as computational pipeline monitoring (CPM), use instruments to monitor internal pipeline parameters (i.e. flow, pressure, temperature, and fluid properties), which are input continuously into a computer simulation software linked to a supervisory control and data acquisition (SCADA) system that analyzes the information mathematically or statistically. Leaks result in unexpected variations or in a well-defined deviation pattern between simulated and measured values. These patterns can be detected and assessed to determine if a leak is present (API, 1995b). The method of leak detection selected for a pipeline is dependent on a variety of factors, including physical pipeline characteristics (length, diameter, thickness, etc.), product characteristics (density, viscosity, etc.), instrumentation and communications capabilities, and economics (Muhlbauer, 1996). Compared to other leak detection methods, SCADA-based leak detection methods have the widest range of applicability to pipeline leak detection and are by far the most highly developed of the leak detection methods presented, as it can detect large leaks rapidly and, over a period of time, detect smaller leaks as well (Jolly *et al*., 1992). However, for multiphase flow pipelines, SCADA-based leak detection methods become much more difficult to apply, and the sensitivity to detecting leaks is reduced because metering multiphase lines is very difficult and inaccurate. Most condition the flow so has to create a measurable flow stream by making a uniform mist or separate the gas and liquid content. The big problem in multiphase leak detections is the varying percentage of gas and liquid content and, most notably, the formation and expulsion of liquid slugs in the pipeline. Hence, for multiphase SCADA-based leak detection systems, accurate and reliable flow measurement instrumentation and transient flow computer models will need to be developed. Moreover, the capabilities and advantages/disadvantages of new technologies should be analyzed with special consideration given to the possible application in subsea pipelines, for which the remoteness of these pipelines, coupled with a number of complex interactions between the released fluids and the subsea environment, makes detection much more difficult.

3.9.2 Pigging

Pigging is a term used to describe a mechanical method for removing contaminants and deposits within the pipe or to clean accumulated liquids in the lower portions of hilly terrain pipelines using a mechanized plunger or pigs (Tiratsoo, 1992). Because of the ability of the pig to remove both corrosion products and sludge from the line, it has been found to be a positive factor in the corrosion control of offshore pipelines. Some specially instrumented pigs, known as "intelligent pigs," may also be used intermittently for the purpose of pipeline integrity monitoring (Cranswick, 2001) which includes detecting wall defects (e.g., corrosion, weld defects, and cracks). Pipeline pigs fit the inside diameter of the pipe and scrape the pipe walls as they are pushed along the pipeline by the flowing fluid. For offshore platforms the pig is launched from offshore and received in the onshore pig catcher. The receiver is in a direct line with the sealine and can be isolated from it to allow the pigs to be removed. Pigs are available in various shapes (Figure 3-29) and are made of different materials, depending on the pigging task to be accomplished. Some have spring-loaded steel knives, wire brushes, or abrasive grit surfaces for the removal of adhering contaminants. Others are semirigid, nonmetallic spheres.

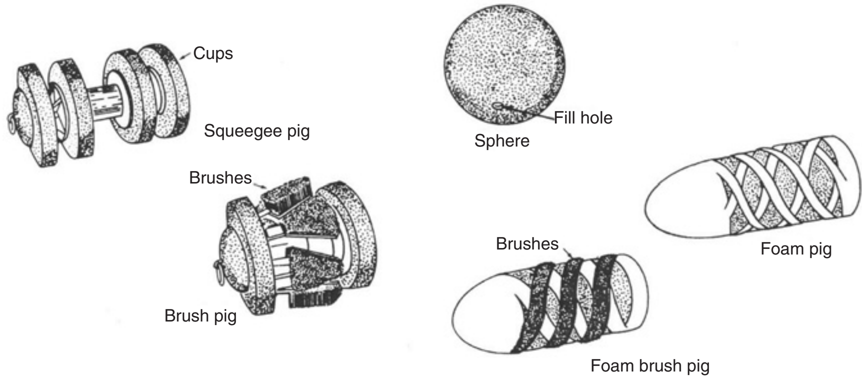

Figure 3-29. Typical types of pig (Campbell, 1992).

During pigging operations there are often technical problems due to the lack of reliable tools for the prediction of the many variables related to the motion of the pig through the pipeline. Hence, the pigging operation requires careful control and coordination. For example, overfrequent pigging causes production downtime or higher operation expenses, whereas infrequent pigging results in less production reduction, but increases the risk of pipeline blockages, including sticking a pig. The pipeline operator must therefore give serious consideration to whether the pipeline really needs to be pigged and whether it is economical to do so. Because there is no commercially available tool to determine the optimum pigging frequency, operators must choose their pigging frequency using rules of thumb based on their field experience, which often involves a high degree of uncertainty.

In pigging operations where the pipe content is unloaded, the liquid holdup builds up as a slug ahead of the moving pig. The arrival of a slug at production or processing equipment is problematic. It causes both mechanical problems (due to high velocities and momentum) and process problems (increasing liquid levels, causing surges and trips). There are in general several means to reduce slugging in pipelines (Haandrikman *et al.*, 1999; Havre *et al.*, 2000). In some cases, operators can minimize liquid accumulation by managing fields and pipelines in such a way as to create a suitable fluid flow regime (i.e., mist flow regime) under which the gas velocity is high enough to keep liquids dispersed continuously. While it is desirable to design the lines to avoid slugging, in practice this can be difficult while maintaining the ability to turn down the pipeline flow rate. In these cases, consideration should be given to providing suitable process equipment to handle possible slugging (GPSA, 1998; Xiao and Shoup, 1998).

The pigging operation in multiphase pipelines is a transient operation. Transient flow is observed not only during the pig running time, but also for a long period after the pig exits. This situation occurs even if the inlet gas and liquid flow rates and outlet pressures are kept constant (Minami, 1991). Analysis of such transient flow behavior in a pipeline is necessary not only for designing the downstream processing facilities, but also for establishing safe operating procedures. Hence, there is a definite need to develop reliable and comprehensive pigging models for better understanding of transient behavior of fluids during these operations. A simple model to simulate transient flow behavior in a two-phase flow pipeline under pigging operation has been presented by Minami (1991). In this model, Minami (1991) assumed that the gas phase can be considered to be flowing in a quasi-steady condition and then coupled it with the Taitel *et al*. (1989) simplified transient two-fluid model. The model, however, needs significant modifications in order to be used for simulating transient flow in a pipeline-riser system, where the quasi-steady-state approach is not suitable for such systems due to a high accumulation of gas upstream of the pig (Yeung and Lima, 2002). For this purpose, a new transient two-fluid model has been developed by Yeung and Lima (2002), which is appropriate for estimating two-phase flow-pigging hydraulics, especially in pipeline-riser systems.

Example 3-1

Using the Beggs and Brill (1973) two-phase flow correlation and the presented calculation algorithm in Figure 3-6, predict the flow temperature and pressure profiles of the Masjed Soliman to Mahshahr gas/condensate transmission line (104.4 miles long and 19-inch inside diameter) located in southwestern Iran. The pipeline elevation profile is given in Table 3-7. The gas composition and the pipeline data needed for all the runs are given in Tables 3-8 and 3-9, respectively. Also, the parameters of the C_6^+ fraction are molecular weight, 107.8; normal boiling point, 233.8°F;

Table 3-7
Masjed Soleiman–Mahshahr Pipeline Profile

Segment number	Length (miles)	Inlet elevation (ft)
1	7.09	1740.00
2	4.84	672.57
3	6.4	1197.51
4	3.1	688.98
5	0.62	1410.76
6	7.77	862.86
7	9.94	295.28
8	14.93	426.51
9	7.34	196.85
10	9.46	98.43
11	9.94	55.77
12	9.94	49.21
13	9.94	19.69
14	3.11	36.09

Table 3-8
Composition of Transported Gas

Component	Mole percent
H_2S	25.60
N_2	0.20
CO_2	9.90
C_1	62.90
C_2	0.70
C_3	0.20
iC_4	0.06
nC_4	0.09
iC_5	0.04
nC_5	0.05
C_6^+	0.26

Table 3-9
Other Pipeline Data

Inlet pressure, psia	1165
Inlet temperature, °F	95
Surrounding temperature,°F	77
Flow rate, MMSCFD	180
Overall heat-transfer coefficient, $Btu/hr.ft^2.°F$	0.25

critical temperature, 536.7°F; critical pressure, 374.4 psia; and acentric factor, 0.3622.

Assumptions

The following assumptions will be made in the calculations required for pipeline simulation: (1) this example uses the Peng and Robinson (1976) equation of state to implement the thermodynamic model, as it has proven reliable for gas/condensate systems (in conjunction with the thermodynamic model, empirical correlations for viscosity and surface tension are used) and (2) the overall composition of the fluid remains constant along the pipe, which can be seen as a steady-state requirement.

Solution

The calculation algorithm presented here couples a hydrodynamic model, based on the Beggs and Brill (1973) two-phase flow correlation, to obtain the pressure profile along the pipeline and a phase behavior model, based on the Peng and Robinson (1976) equation of state, for determining the phase condition at any point in the pipe and the fluid properties. Following the procedure outlined by Brill and Beggs (1991), the resulting algorithm allows one to predict the flow behavior of a natural gas/condensate system. Figures 3-30 and 3-31 show respectively, results of the temperature and pressure profiles of the Masjed Soleiman to Mahshahr gas/condensate transmission line.

Examined case results demonstrate the ability of the Beggs and Brill (1991) introduced algorithm (Figure 3-6) to predict the pressure and temperature gradients at any point in the pipe.

Note

The Masjed Soleiman to Mahshahr pipeline was intended to operate as a single-phase gas line and was designed on that basis. However, operation of the line clearly showed liquid to be present. This occurs when the pressure and the temperature changes traverse the two-phase region of the

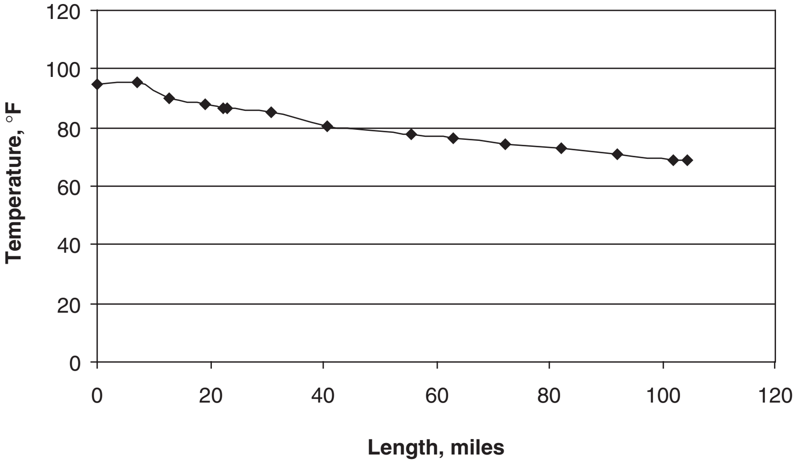

Figure 3-30. Pipeline temperature profile.

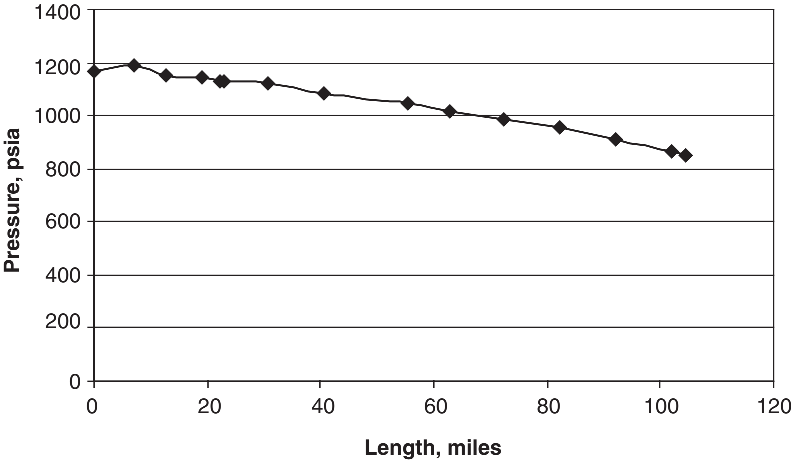

Figure 3-31. Pipeline pressure profile.

phase envelope such that the system changes from single phase to two phase. As the system enters the two-phase region, mass transfer between gas and liquid phases causes condensation and gives rise to two-phase flow in the pipeline. This example shows how the gas pipeline designer has to use phase behavior analysis as an operational tool in order to predict the presence of liquid in the system. Phase behavior analysis is conducted to determine the bubble and dew point curve and to perform flash calculations at any pressure and temperature. Then the operational pressure–temperature profile is overplotted on the phase envelope diagram in order to evaluate the condensation possibility into the gas pipeline.

When thermodynamic conditions dictate single-phase gas flow, the single-phase flow equations should be applied instead of two-phase flow models. However, in this case, the Beggs and Brill (1973) two-phase flow correlation (neglecting the liquid holdup parameter in all of calculations) was used to estimate pressure drop for pipe segments containing no liquid.

Example 3-2

Estimate the hydrate formation temperature for the following mixture at 661.37 psia using the Katz gravity chart, Towler and Mokhatab correlation, and Baillie and Wichert chart (the experimentally observed value is 57.56°F).

CH_4 82.0 mol%
CO_2 12.6 mol%
H_2S 5.4 mol%

Solution

Calculate the molar mass and gravity of the gas mixture: $0.82 \times 16.043 + 0.126 \times 44.011 + 0.054 \times 34.082 = 20.541$. Therefore, $SG = 20.541/28.966 = 0.709$.

1. Using the Katz gravity chart: $T_h = 59°F$.
2. Using the Towler and Mokhatab (2005) correlation [Equation (3-53)] representing the Katz gravity chart:

$$\begin{aligned} T_h &= 13.47\ln(661.37) + 34.27\ln(0.709) - 1.675\ln(661.37) \\ &\quad \times \ln(0.709) - 20.35 \\ T_h &= 59.1°\text{F} \end{aligned}$$

To use the Baillie and Wichert method to estimate the hydrate formation temperature, enter the main chart at the given pressure and then move to the right to the appropriate H_2S concentration. From that point, go straight down to the appropriate gas-specific gravity. Then follow the sloping lines down to the temperature axis. This value is the based temperature. In this case, the base temperature is about 66°F. Now go to the temperature correction in the upper left. Enter the chart with the H_2S concentration and move across to the appropriate propane concentration. From that point, go straight down to the appropriate pressure curve and then read the temperature correction from the left or the right axis, depending on the position in the pressure section. In this case, the temperature correction is negative (−7°F) because it comes from the left half of the correction chart. Finally, the hydrate formation temperature is obtained by adding the temperature correction to the base temperature, 59°F.

Example 3-3

Natural gas flowing at 4 MMscfd is transferred through a pipeline from a well site to a processing plant. The production also includes 7.062 scfd of water, which is transferred in the same pipeline. The inlet pressure and temperature are 1165 psia and 95°F, respectively. The hydrate formation temperature of the gas is determined to be 84.485°F at stream pressure. In transportation through the pipeline, the gas is expected to cool to 48.92°F. Calculate the amount of methanol that should be injected into the line in order to prevent hydrate formation.

Solution

1. Calculate the required temperature depression:

$$\Delta T = 84.485 - 48.920 = 35.565^\circ\mathrm{F}$$

2. Use Equation (3-54) to estimate the concentration of methanol required to yield the desired temperature depression:

$$W = [100(32.04)(35.565)]/[2335 + (32.04)(35.565)] = 32.796\%$$

   Because this is outside the range of applicability of the Hammerschmidt equation, this result should be taken with some caution. However, the value obtained from the Nielsen and Bucklin equation [Equation (3-55)] is $X_{\mathrm{methanol}} = 0.2399$, and then to calculate the weight percent from this mole fraction, the following equation is used:

$$\begin{aligned}\text{Weight fraction methanol} &= [X_{\mathrm{methanol}} \times \mathrm{MW}_{\mathrm{methanol}}]/[18.05 \\ &\quad + X_{\mathrm{methanol}}(\mathrm{MW}_{\mathrm{methanol}} - 18.015] \\ &= [0.2399(32.04)]/[18.05 + 0.2399 \\ &\quad (32.04 - 18.015)] = 35.952\ \mathrm{wt}\%\end{aligned}$$

3. Calculate the amount of methanol required to get a 35.952 wt% solution based on water and methanol:

$$\begin{aligned}0.35952 = {} & [\text{methanol mass flow rate}]/[\text{water mass flow rate} \\ & + \text{methanol mass flow rate}]\end{aligned}$$

   Therefore, methanol mass flow rate = 247.439 $\mathrm{lb_m}$/day. The density of methanol is 49.717 $\mathrm{lb_m/ft^3}$, so the methanol is injected at a rate of 4.976 $\mathrm{ft^3}$/day.

Note that the amount of water in the natural gas has been neglected in this calculation, where the gas is being produced with free water, so it can be assumed that is saturated with water. However, to estimate the amount of methanol required to saturate the gas, Figure 3-15 can be used. From that figure and then estimating the amount of methanol in the vapor phase, the total methanol requirement will be determined.

REFERENCES

Acikgoz, M., Franca, F., and Lahey Jr, R.T., An experimental study of three-phase flow regimes. *Int. J. Multiphase Flow* **18**(3), 327–336 (1992).

Adewmi, M.A., and Bukacek, R.F., Two-phase pressure drop in horizontal pipelines. *J. Pipelines* **5**, 1–14 (1985).

Amdal, J., *et al*., "Handbook of Multiphase Metering." Produced for The Norwegian Society for Oil and Gas Measurement, Norway (2001).

Ansari, A.M., Sylvester, A.D., Sarica, C., Shoham, O., and Brill, J.P., A comprehensive mechanistic model for upward two-phase flow in wellbores. *SPE Prod. Facilit. J*. 143–152 (May 1994).

API, "Computational Pipeline Monitoring." API Publication 1130, 17, American Petroleum Institute, TX (1995a).

API, "Evaluation Methodology for Software Based Leak Detection Systems." API Publication 1155, 93, American Petroleum Institute, TX (1995b).

API RP 14E, "Recommended Practice for Design and Installation of Offshore Production Platform Piping Systems," 5th Ed., p. 23. American Petroleum Institute, Washington, DC (1991).

Asante, B., "Two-Phase Flow: Accounting for the Presence of Liquids in Gas Pipeline Simulation." Paper presented at 34th PSIG Annual Meeting, Portland, Oregon (Oct. 23–25, 2002).

AsphWax's Flow Assurance course, "Fluid Characterization for Flow Assurance." AsphWax Inc., Stafford, TX (2003).

Ayala, F.L., and Adewumi, M.A., Low-liquid loading multiphase flow in natural gas pipelines. *ASME J. of Energy Res. Technol*. **125**, 284–293 (2003).

Baillie, C., and Wichert, E., Chart gives hydrate formation temperature for natural gas. *Oil Gas J.* **85**(4), 37–39 (1987).

Baker, O., Design of pipelines for simultaneous flow of oil and gas. *Oil Gas J*. **53**, 185 (1954).

Banerjee, S., “Basic Equations.” Lecture presented at the Short Course on Modeling of Two-Phase Flow Systems, ETH Zurich, Switzerland (March 17–21, 1986).

Barnea, D., A unified model for prediction flow pattern transitions for the whole range of pipe inclinations. *Int. J. Multiphase Flow* **13**(1), 1–12 (1987).

Barnea, D., and Taitel, Y., Stratified three-phase flow in pipes: Stability and transition. *Chem. Eng. Comm.* 141–142, 443–460 (1996).

Battara, V., Gentilini, M., and Giacchetta, G., Condensate-line correlations for calculating holdup, friction compared to field data. *Oil Gas J.* **30**, 148–52 (1985).

Bay, Y., “Pipelines and Risers,” Vol. 5. Elsevier Ocean Engineering Book Series (2001).

Beggs, H.D., and Brill, J.P., “A Study of Two-Phase Flow in Inclined Pipes.” JPT, 607–17; Trans., AIME, 255 (May 1973).

Bendiksen, K., Brandt, I., Jacobsen, K.A., and Pauchon, C., “Dynamic Simulation of Multiphase Transportation Systems.” Multiphase Technology and Consequence for Field Development Forum, Stavanger, Norway (1987).

Bendiksen, K., Malnes, D., Moe, R., and Nuland, S., The dynamic two-fluid model OLGA: Theory and application. *SPE Prod. Eng.* **6**, 171–180 (1991).

Bertola, V., and Cafaro, E., “Statistical Characterization of Subregimes in Horizontal Intermittent Gas/Liquid Flow.” Proc. the 4th International Conference on Multiphase Flow, New Orleans, LA (2001).

Bjune, B., Moe, H., and Dalsmo, M., Upstream control and optimization increases return on investment. *World Oil* **223**, 9 (2002).

Black, P.S., Daniels, L.C., Hoyle, N.C., and Jepson, W.P., Studying transient multiphase flow using the pipeline analysis code (PLAC). *ASME J. Energy Res. Technol.* **112**, 25–29 (1990).

Bloys, B., Lacey, C., and Lynch, P., “Laboratory Testing and Field Trial of a New Kinetic Hydrate Inhibitor.” Proc. 27th Annual Offshore Technology Conference, OTC 7772, PP. 691–700, Houston, TX (1995).

Bonizzi, M., and Issa, R.I., On the simulation of three-phase slug flow in nearly horizontal pipes using the multi-fluid model. *Int. J. Multiphase Flow* **29**, 1719–1747 (2003).

Boriyantoro, N.H., and Adewumi, M.A., “An Integrated Single-Phase/Two-Phase Flow Hydrodynamic Model for Predicting the Fluid Flow

Behavior of Gas Condensate Pipelines." Paper presented at 26th PSIG Annual Meeting, San Diego, CA (1994).

Brauner, N., and Maron, M.D., Flow pattern transitions in two-phase liquid-liquid flow in horizontal tubes, *Int. J. Multiphase Flow* **18**, 123–140 (1992).

Brauner, N., The prediction of dispersed flows boundaries in liquid-liquid and gas-liquid systems, *Int. J. Multiphase Flow* **27**, 885–910 (2001).

Brill, J.P., and Beggs, H.D., "Two-Phase Flow in Pipes," 6th Ed. Tulsa University Press, Tulsa, OK (1991).

Bufton, S.A., Ultra Deepwater will require less conservative flow assurance approaches. *Oil Gas J*. **101**(18), 66–77 (2003).

Campbell, J.M., "Gas Conditioning and Processing," 3rd Ed. Campbell Petroleum Series, Norman, OK (1992).

Carroll, J.J., "Natural Gas Hydrates: A Guide for Engineers." Gulf Professional Publishing, Amsterdam, The Netherlands (2003).

Carroll, J.J., "An Examination of the Prediction of Hydrate Formation Conditions in Sour Natural Gas." Paper presented at the GPA Europe Spring Meeting, Dublin, Ireland (May 19–21, 2004).

Carson, D.B., and Katz, D.L., Natural gas hydrates. *Petroleum Trans. AIME* **146**, 150–158 (1942).

Chen, C.J., Woo, H.J., and Robinson, D.B., "The Solubility of Methanol or Glycol in Water-Hydrocarbon Systems." GPA Research Report RR-117, Gas Processors Association, OK (March 1988).

Chen, X., and Guo, L., Flow patterns and pressure drop in oil-air-water three-phase flow through helically coiled tubes. *Int. J. Multiphase Flow* **25**, 1053–1072 (1999).

Cheremisinoff, N.P., "Encyclopedia of Fluid Mechanics," Vol. 3. Gulf Professional Publishing, Houston, TX (1986).

Chisholm, D., and Sutherland, L.A., "Prediction of Pressure Gradient in Pipeline Systems during Two-Phase Flow." Paper presented at Symposium on Fluid Mechanics and Measurements in Two-Phase Flow Systems, Leeds (Sept. 1969).

Cindric, D.T., Gandhi, S.L., and Williams, R.A., "Designing Piping Systems for Two-Phase Flow." Chem. Eng. Progress, 51–59 (March 1987).

Collier, J.G., and Thome, J.R., "Convective Boiling and Condensation," 3rd Ed. Clarendon Press, Oxford, UK (1996).

Colson, R., and Moriber, N.J., Corrosion control. *Civil Eng*. **67**(3), 58–59 (1997).

Copp, D.L., “Gas Transmission and Distribution.” Walter King Ltd., London (1970).

Courbot, A., “Prevention of Severe Slugging in the Dunbar 16 Multiphase Pipelines.” Paper presented at the Offshore Technology Conference, Houston, TX (May 1996).

Covington, K.C., Collie, J.T., and Behrens, S.D., “Selection of Hydrate Suppression Methods for Gas Streams.” Paper presented at the 78th GPA Annual Convention, Nashville, TN (1999).

Cranswick, D., “Brief Overview of Gulf of Mexico OCS Oil and Gas Pipelines: Installation, Potential Impacts, and Mitigation Measures.” U.S. Department of the Interior, Minerals Management Service, Gulf of Mexico OCS Region, New Orleans, LA (2001).

Dahl, E., *et al*., “Handbook of Water Fraction Metering,” Rev. 1. Norwegian Society for Oil and Gas Measurements, Norway (June 2001).

Decarre, S., and Fabre, J., Etude Sur La Prediction De l’Inversion De Phase. *Revue De l’Institut Francais Du Petrole* **52**, 415–424 (1997).

De Henau, V., and Raithby, G.D., A transient two-fluid model for the simulation of slug flow in pipelines. *Int. J. Multiphase Flow* **21**, 335–349 (1995).

Dukler, A.E., “Gas-Liquid Flow in Pipelines Research Results.” American Gas Assn. Project NX-28 (1969).

Dukler, A.E., and Hubbard, M.G., “The Characterization of Flow Regimes for Horizontal Two-Phase Flow.” Proc. Heat Transfer and Fluid Mechanics Institute, 1, 100–121, Stanford University Press, Stanford, CA (1966).

Eaton, B.A., *et al*., The prediction of flow pattern, liquid holdup and pressure losses occurring during continuous two-phase flow in horizontal pipelines. *J. Petro. Technol*. **240**, 815–28 (1967).

Edmonds, B., *et al*., “A Practical Model for the Effect of Salinity on Gas Hydrate Formation.” European Production Operations Conference and Exhibition, Norway (April 1996).

Edmonds, B., Moorwood, R.A.S., and Szczepanski, R., “Hydrate Update.” GPA Spring Meeting, Darlington (May 1998).

Esteban, A., Hernandez, V., and Lunsford, K., “Exploit the Benefits of Methanol.” Paper presented at the 79th GPA Annual Convention, Atlanta, GA (2000).

Fabre, J., *et al*., Severe slugging in pipeline/riser systems. *SPE Prod. Eng*. **5**(3), 299–305 (1990).

Faille, I., and Heintze, E., "Rough Finite Volume Schemes for Modeling Two-Phase Flow in a Pipeline." In Proceeding of the CEA.EDF.INRIA Course, INRIA Rocquencourt, France (1996).

Fidel-Dufour, A., and Herri, J.S., "Formation and Dissociation of Hydrate Plugs in a Water in Oil Emulsion." Paper presented at the 4th International Conference on Gas Hydrates, Yokohama, Japan (2002).

Frostman, L.M., "Anti-aggolomerant Hydrate Inhibitors for Prevention of Hydrate Plugs in Deepwater Systems." Paper presented at the SPE Annual Technical Conference and Exhibition, Dallas, TX (Oct. 1–4, 2000).

Frostman, L.M., *et al.*, "Low Dosage Hydrate Inhibitors (LDHIs): Reducing Total Cost of Operations in Existing Systems and Designing for the Future." Paper presented at the SPE International Symposium on Oilfield Chemicals, Houston, TX (Feb. 5–7, 2003).

Frostman, L.M., "Low Dosage Hydrate Inhibitor (LDHI) Experience in Deepwater." Paper presented at the Deep Offshore Technology Conference, Marseille, France (Nov. 19–21, 2003).

Fuchs, P., "The Pressure Limit for Terrain Slugging." Proceeding of the 3rd BHRA International Conference on Multiphase Flow, Hague, The Netherlands (May 1987).

Furlow, W., "Suppression System Stabilizes Long Pipeline-Riser Liquid Flows." Offshore, Deepwater D&P, 48 + 166 (Oct., 2000).

Giot, M., "Three-Phase Flow," Chap. 7.2. McGraw-Hill, New York (1982).

Goulter, D., and Bardon, M., Revised equation improves flowing gas temperature prediction. *Oil Gas J.* **26**, 107–108 (1979).

Govier, G.W., and Aziz, K., "The Flow of Complex Mixtures in Pipes." Van Nostrand Reinhold Co., Krieger, New York (1972).

GPSA Engineering Data Book, 11th Ed. Gas Processors Suppliers Association, Tulsa, OK (1998).

Gregory, G.A., and Aziz, K., Design of pipelines for multiphase gas-condensate flow. *J. Can. Petr. Technol.* 28–33 (1975).

Griffith, P., and Wallis, G.B., Two-phase slug flow. *J. Heat Transfer Trans. ASME.* **82**, 307–320 (1961).

Haandrikman, G., Seelen, R., Henkes, R., and Vreenegoor, R., "Slug Control in Flowline/Riser Systems." Paper presented at the 2nd International Conference on Latest Advance in Offshore Processing, Aberdeen, UK (Nov. 9–10, 1999).

Hall, A.R.W., "Flow Patterns in Three-Phase Flows of Oil, Water, and Gas." Paper presented at the 8th BHRG International Conference on Multiphase Production, Cannes, France (1997).

Hammerschmidt, E.G., Formation of gas hydrates in natural gas transmission lines. *Ind. Eng. Chem*. **26**, 851–855 (1934).

Hart, J., and Hamersma P.J. Correlations predicting frictional pressure drop and liquid holdup during horizontal gas-liquid pipe flow with small liquid holdup. *Int. J. Multiphase Flow* **15**(6), 974–964 (1989).

Hartt, W.H., and Chu, W., New methods for CP design offered. *Oil Gas J*. **102**(36), 64–70 (2004).

Hasan, A.R., Void fraction in bubbly and slug flow in downward two-phase flow in vertical and inclined wellbores, *SPE Prod. Facil.* **10**(3), 172–176 (1995).

Hasan, A.R., and Kabir, C.S., "Predicting Multiphase Flow Behavior in a Deviated Well." Paper presented at the 61st SPE Annual Technical Conference and Exhibition, New Orleans, LA (1986).

Hasan, A.R., and Kabir, C.S., Gas void fraction in two-phase up-flow in vertical and inclined annuli. *Int. J. Multiphase Flow* **18**(2), 279–293 (1992).

Hasan, A.R., and Kabir, C.S., Simplified model for oil/water flow in vertical and deviated wellbores. *SPE Prod. Facil.* **14**(1), 56–62 (1999).

Hasan, A.R., and Kabir, C.S., "Fluid and Heat Transfers in Wellbores." Society of Petroleum Engineers (SPE) Publications, Richardson, TX (2002).

Havre, K., and Dalsmo, M., "Active Feedback Control as the Solution to Severe Slugging." Paper presented at the SPE Annual Technical Conference and Exhibition, New Orleans, LA (Oct. 3, 2001).

Havre, K., Stornes, K., and Stray, H., "Taming Slug Flow in Pipelines." ABB Review No. 4, 55–63, ABB Corporate Research AS, Norway (2000).

Hedne, P., and Linga, H., "Suppression of Terrain Slugging with Automatic and Manual Riser Chocking." Advances in Gas-Liquid Flows, 453–469 (1990).

Hein, M., "HP41 Pipeline Hydraulics and Heat Transfer Programs." PennWell Publishing Company, Tulsa, OK (1984).

Henriot, V., Courbot, A., Heintze, E., and Moyeux, L., "Simulation of Process to Control Severe Slugging: Application to the Dunbar

Pipeline." Paper presented at the SPE Annual Conference and Exhibition, Houston, TX (1999).

Hewitt, G.F., and Roberts, D.N., "Studies of Two-Phase Flow Patterns by Simultaneous X-ray and Flash Photography." AERE-M 2159, HMSO (1969).

Hill, T.H., Gas injection at riser base solves slugging, flow problems. *Oil Gas J.* **88**(9), 88–92 (1990).

Hill, T.J., "Gas-Liquid Challenges in Oil and Gas Production." Proceeding of ASME Fluids Engineering Division Summer Meeting, Vancouver, BC, Canada (June 22–26, 1997).

Holt, A.J., Azzopardi, B.J., and Biddulph, M.W., "The Effect of Density Ratio on Two-Phase Frictional Pressure Drop." Paper presented at the 1st International Symposium on Two-Phase Flow Modeling and Experimentation, Rome, Italy (Oct. 9–11, 1995).

Holt, A.J., Azzopardi, B.J., and Biddulph, M.W., Calculation of two-phase pressure drop for vertical up flow in narrow passages by means of a flow pattern specific models. *Trans. IChemE* **77**(Part A), 7–15 (1999).

Hunt, A., Fluid properties determine flow line blockage potential. *Oil Gas J.* **94**(29), 62–66 (1996).

IFE, "Mitigating Internal Corrosion in Carbon Steel Pipelines." Institute of Energy Technology News, Norway (April 2000).

Jansen, F.E., Shoham, O., and Taitel, Y., The elimination of severe slugging, experiments and modeling. *Int. J. Multiphase Flow* **22**(6), 1055–1072 (1996).

Johal, K.S., *et al.*, "An Alternative Economic Method to Riser Base Gas Lift for Deepwater Subsea Oil/Gas Field Developments." Proceeding of the Offshore Europe Conference, 487–492, Aberdeen, Scotland (9–12 Sept., 1997).

Jolly, W.D., Morrow, T.B., O'Brien, J.F., Spence, H.F., and Svedeman, S.J., "New Methods for Rapid Leak Detection in Offshore Pipelines." Final Report, SWRI Project No. 04-4558, SWRI (April 1992).

Jones, O.C., and Zuber, N., The interrelation between void fraction fluctuations and flow patterns in two-phase flow *Int. J. Multiphase Flow.* **2**, 273–306 (1975).

Kang, C., Wilkens, R.J., and Jepson, W.P., "The Effect of Slug Frequency on Corrosion in High-Pressure, Inclined Pipelines." Paper presented at the NACE International Annual Conference and Exhibition, Paper No. 96020, Denver, CO (March 1996).

Katz, D.L., Prediction of conditions for hydrate formation in natural gases. *Trans. AIME* **160**, 140–149 (1945).

Kelland, M.A., Svartaas, T.M., and Dybvik, L., "New Generation of Gas Hydrate Inhibitors." 70th SPE Annual Technical Conference and Exhibition, 529–537, Dallas, TX (Oct. 22–25, 1995).

Kelland, M.A., Svartaas, T.M., Ovsthus, J., and Namba, T., A new class of kinetic inhibitors. *Ann. N.Y. Acad. Sci.* **912**, 281–293 (2000).

King, M.J.S, "Experimental and Modelling Studies of Transient Slug Flow." Ph.D. Thesis, Imperial College of Science, Technology, and Medicine, London, UK (March 1998).

Klemp, S., "Extending the Domain of Application of Multiphase Technology." Paper presented at the 9th BHRG International Conference on Multiphase Technology, Cannes, France (June 16–18, 1999).

Kohl, A.L., and Risenfeld, F.C., "Gas Purification." Gulf Professional Publishing, Houston, TX (1985).

Kovalev, K., Cruickshank, A., and Purvis, J., "Slug Suppression System in Operation." Paper presented at the 2003 Offshore Europe Conference, Aberdeen, UK (Sept. 2–5, 2003).

Kumar, S., "Gas Production Engineering." Gulf Professional Publishing, Houston, TX (1987).

Lagiere, M., Miniscloux, and Roux, A., Computer two-phase flow model predicts pipeline pressure and temperature profiles. *Oil Gas J.* 82–91 (April 9, 1984).

Langner, *et al.*, "Direct Impedance Heating of Deepwater Flowlines." Paper presented at the Offshore Technology Conference (OTC), Houston, TX (1999).

Lederhos, J.P., Longs, J.P., Sum, A., Christiansen, R.l., and Sloan, E.D., Effective kinetic inhibitors for natural gas hydrates. *Chem. Eng. Sci.* **51**(8), 1221–1229 (1996).

Lee, H.A., Sun, J.Y., and Jepson, W.P., "Study of Flow Regime Transitions of Oil-Water-Gas Mixture in Horizontal Pipelines." Paper presented at the 3rd International Offshore and Polar Engng Conference, Singapore (June 6–11 1993).

Leontaritis, K.J., "The Asphaltene and Wax Deposition Envelopes." The Symposium on Thermodynamics of Heavy Oils and Asphaltenes, Area 16C of Fuels and Petrochemical Division, AIChE Spring

National Meeting and Petroleum Exposition, Houston, TX (March 19–23, 1995).

Leontaritis, K.J., "PARA-Based (Paraffin-Aromatic-Resin-Asphaltene) Reservoir Oil Characterization." Paper presented at the SPE International Symposium on Oilfield Chemistry, Houston, TX (Feb.18–21, 1997a).

Leontaritis, K.J., Asphaltene destabilization by drilling/completion fluids. *World Oil* **218**(11), 101–104 (1997b).

Leontaritis, K.J., "Wax Deposition Envelope of Gas Condensates." Paper presented at the Offshore Technology Conference (OTC), Houston, TX (May 4–7 1998).

Lin, P.Y., and Hanratty, T.J., Detection of slug flow from pressure measurements. *Int. J. Multiphase Flow* **13**, 13–21 (1987).

Lockhart, R.W., and Martinelli, R.C., Proposed correlation of data for isothermal two-phase, two-component flow in pipes. *Chem. Eng. Prog*. **45**, 39–48 (1949).

Maddox, R.N., *et al*., "Predicting Hydrate Temperature at High Inhibitor Concentration." Proceedings of the Laurance Reid Gas Conditioning Conference, 273–294, Norman, OK (1991).

Mann, S.L., *et al*., "Vapour-Solid Equilibrium Ratios for Structure I and II Natural Gas Hydrates." Proceedings of the 68th GPA Annual Convention, 60–74, San Antonio, TX (March 13–14, 1989).

Martinelli, R.C., and Nelson, D.B., Prediction of pressure drop during forced circulation boiling of water. *Trans. ASME* **70**, 695 (1948).

Masella, J.M., Tran, Q.H., Ferre, D., and Pauchon, C., Transient simulation of two-phase flows in pipes. *Oil Gas Sci. Technol*. **53**(6), 801–811 (1998).

McCain, W.D., "The Properties of Petroleum Fluids," 2nd Ed. Pennwell Publishing Company, Tulsa, OK (1990).

McLaury, B.S., and Shirazi, S.A., An alternative method to API RP 14E for predicting solids erosion in multiphase flow. *ASME J. Energy Res. Technol.* **122**, 115–122 (2000).

McLeod, H.O., and Campbell, J.M., Natural gas hydrates at pressure to 10,000 psia. *J. Petro. Technol.* **13**, 590–594 (1961).

Mehta, A.P., and Sloan, E.D., "Structure H Hydrates: The State-of-the-Art." Paper presented at the 75th GPA Annual Convention, Denver, CO (1996).

Mehta, A., Hudson, J., and Peters, D., "Risk of Pipeline Over-Pressurization during Hydrate Remediation by Electrical Heating."

Paper presented at the Chevron Deepwater Pipeline and Riser Conference, Houston, TX (March 28–29, 2001).

Mehta, A.P., Hebert, P.B., and Weatherman, J.P., "Fulfilling the Promise of Low Dosage Hydrate Inhibitors: Journey from Academic Curiosity to Successful Field Implementations." Paper presented at the 2002 Offshore Technology Conference, Houston, TX (May 6–9, 2002).

Minami, K., "Transient Flow and Pigging Dynamics in Two-Phase Pipelines." Ph.D. Thesis, University of Tulsa, Tulsa, OK (1991).

Mokhatab, S., Correlation predicts pressure drop in gas-condensate pipelines. *Oil Gas J.* **100**(4), 66–68 (2002a).

Mokhatab, S., New correlation predicts liquid holdup in gas-condensate pipelines. *Oil Gas J.* **100**(27), 68–69 (2002b).

Mokhatab, S., Model aids design of three-phase, gas-condensate transmission lines. *Oil Gas J.* **100**(10), 60–64 (2002c).

Mokhatab, S., Three-phase flash calculation for hydrocarbon systems containing water. *J. Theor. Found. Chem. Eng.* **37**(3), 291–294 (2003).

Mokhatab, S., Upgrade velocity criteria for sizing multiphase pipelines. *J. Pipeline Integrity* **3**(1), 55–56 (2004).

Mokhatab, S., "Interaction between Multiphase Pipelines and Downstream Processing Plants." Report No. 3, TMF3 Sub-Project VII: Flexible Risers, Transient Multiphase Flow (TMF) Program, Cranfield University, Bedfordshire, England (May 2005).

Mokhatab, S., "Explicit method predicts temperature and pressure profiles of gas-condensate pipelines." Accepted for publication in Energy Sources: Part A (2006a).

Mokhatab, S., Severe slugging in a catenary-shaped riser: Experimental and simulation studies." Accepted for publication in *J. Petr. Sci. Technol*. (2006b).

Mokhatab, S., Dynamic simulation of offshore production plants. Accepted for publication in *J. Petr. Sci. Technol.* (2006c).

Mokhatab, S., Severe slugging in flexible risers: Review of experimental investigations and OLGA predictions. Accepted for publication in *J. Petro. Sci. Technol.* (2006d).

Mokhatab, S., and Bonizzi, M., "Model predicts two-phase flow pressure drop in gas-condensate transmission lines." Accepted for publication in Energy Sources: Part A (2006).

Mokhatab, S., Towler, B.F., and Purewal, S., A review of current technologies for severe slugging remediation." Accepted for publication in *J. Petro. Sci. Technol*. (2006a).

Mokhatab, S., Wilkens, R.J., and Leontaritis, K.J., "A Review of strategies for solving gas hydrate problems in subsea pipelines." Accepted for publication in Energy Sources: Part A (2006b).

Molyneux, P., Tait, A., and Kinving, J., "Characterization and Active Control of Slugging in a Vertical Riser." Proceeding of the 2nd North American Conference on Multiphase Technology, 161–170, Banff, Canada (June 21–23, 2000).

Moody, L.F., Friction factors for pipe flow. *Trans. ASME* **66**, 671–684 (1944).

Mucharam, L., Adewmi, M.A., and Watson, R., Study of gas condensation in transmission pipelines with a hydrodynamic model. *SPE J.* 236–242 (1990).

Muhlbauer, K.W., "Pipeline Risk Management Manual," 2nd Ed. Gulf Professional Publishing, Houston, TX (1996).

Narayanan, L., Leontaritis, K.J., and Darby, R., "A Thermodynamic Model for Predicting Wax Precipitation from Crude Oils." The Symposium of Solids Deposition, Area 16C of Fuels and Petrochemical Division, AIChE Spring National Meeting and Petroleum Exposition, Houston, TX (March 28–April 1, 1993).

Nasrifar, K., Moshfeghian, M., and Maddox, R.N., Prediction of equilibrium conditions for gas hydrate formation in the mixture of both electrolytes and alcohol. *Fluid Phase Equilibria* **146**, 1–2, 1–13 (1998).

National Association of Corrosion Engineers, "Sulfide Stress Cracking Resistant Metallic Materials for Oil Field Equipment." NACE Std MR-01-75 (1975).

Ng, H.J., and Robinson, D.B., The measurement and prediction of hydrate formation in liquid hydrocarbon-water systems. *Ind. Eng. Chem. Fund*. **15**, 293–298 (1976).

Nielsen, R.B., and Bucklin, R.W., Why not use methanol for hydrate control? *Hydrocarb. Proc.* **62**(4), 71–78 (1983).

Nyborg, R., Corrosion control in oil and gas pipelines. *Business Briefing Exploration Prod. Oil Gas Rev.* **2**, 79–82 (2003).

Oram, R.K., "Advances in Deepwater Pipeline Insulation Techniques and Materials." Deepwater Pipeline Technology Congress, London, UK (Dec. 11–12, 1995).

Oranje, L., Condensate behavior in gas pipelines is predictable. *Oil Gas J.* **39** (1973).

Osman, M.E., and El-Feky, S.A., Design methods for two-phase pipelines compared, evaluated. *Oil Gas J.* **83**(35), 57–62 (1985).

Palermo, T., Argo, C.B., Goodwin, S.P., and Henderson, A., Flow loop tests on a novel hydrate inhibitor to be deployed in North Sea ETAP field. *Ann. N.Y. Acad. Sci.* **912**, 355–365, (2000).

Pan, L., "High Pressure Three-Phase (Gas/Liquid/Liquid) Flow." Ph.D. Thesis, Imperial College, London (1996).

Patault, S., and Tran, Q.H., "Modele et Schema Numerique du Code TACITE-NPW." Technical Report 42415, Institut Francais du Petrole, France (1996).

Pauchon, C., Dhulesia, H., Lopez, D., and Fabre, J., "TACITE: A Comprehensive Mechanistic Model for Two-Phase Flow." Paper presented at the 6th BHRG International Conference on Multiphase Production, Cannes, France (1993).

Pauchon, C., Dhulesia, H., Lopez, D., and Fabre, J., "TACITE: A Transient Tool for Multiphase Pipeline and Well Simulation." Paper presented at the SPE Annual Technical Conference and Exhibition, New Orleans, LA (1994).

Peng, D., and Robinson, D.B., A new two-constant equation of state. *Ind. Eng. Chem. Fundam.* **15**(1), 59–64 (1976).

Petalas, N., and Aziz, K., A mechanistic model for multiphase flow in pipes. *J. Can. Petr. Technol.* **39**, 43–55 (2000).

Polignano, R., Value of glass-fiber fabrics proven for bituminous coatings. *Oil Gas J.* **80**(41), 156–158, 160 (1982).

Pots, B.F.M., *et al.*, "Severe Slug Flow in Offshore Flowline/Riser Systems." Paper presented at the SPE Middle East Oil Technical Conference and Exhibition, Bahrain (March 11–14, 1985).

Rajkovic, M., Riznic, J.R., and Kojasoy, G., "Dynamic Characteristics of Flow Pattern Transitions in Horizontal Two-Phase Flow." Proc. 2nd European Thermal Science and 14th UIT National Heat Transfer Conference, 3, 1403–1408, Edizioni ETS, Pisa, Italy (1996).

Ramachandran, S., Breen, P., and Ray, R., Chemical programs assure flow and prevent corrosion in deepwater facilities and dlowlines. *InDepth* **6**, 1 (2000).

Rhodes, K.I., Pipeline protective coatings used in saudi Arabia. *Oil Gas J.* **80**(31), 123–127 (1982).

Ripmeester, J.A., Tse, J.S., Ratcliffe, C.J., and McLaurin, G.E., *Nature* **135**, 325 (1987).

Sagatun, S.I., Riser slugging: A mathematical model and the practical consequences. *SPE Product. Facil. J.* **19**(3), 168–175 (2004).

Salama, M.M., An alternative to API 14E erosional velocity limits for sand-laden fluids. *ASME J. Energy Res. Technol.* **122**, 71–77 (2000).

Samant, A.K., "Corrosion Problems in Oil Industry Need More Attention." Paper presented at ONGC Library, Oil and Natural Gas Corporation Ltd. (Feb. 2003).

Sandberg, C., "Holmes, J., McCoy, K., and Koppitsch, H., The application of a continues leak detection system to pipelines and associated equipment. *IEEE Transact. Indust. Appl.* **25**, 5 (1989).

Sarica, C., and Shoham, O., A simplified transient model for pipeline/riser systems. *Chem. Engi. Sci.* **46**(9), 2167–2179 (1991).

Sarica, C., and Tengesdal, J.Q., A New Technique to Eliminate Severe Slugging in Pipeline/Riser Systems." Paper presented at the 75th SPE Annual Technical Conference and Exhibition, Dallas, TX (Oct. 1–4, 2000).

Schmidt, Z., "Experimental Study of Two-Phase Slug Flow in a Pipeline-Riser System." Ph.D. Dissertation, University of Tulsa, Tulsa, OK (1977).

Schmidt, Z., Brill, J.P., and Beggs, D.H., Experimental study of severe slugging in a two-phase flow pipeline-riser system. *SPE J.* **20**, 407–414 (1980).

Schmidt, Z., Doty, D.R., and Dutta-Roy, K., Severe slugging in offshore pipeline-riser systems. *SPE J.* 27–38 (1985).

Schweikert, L.E., Tests prove two-phase efficiency for offshore pipeline. *Oil Gas J.* 39–42 (1986).

Scott, S.L., Brill, J.P., Kuba, G.E., Shoham, K.A., and Tam, W., "Two-Phase Flow Experiments in the Prudhoe Bay Field of Alaska." Paper presented at the Multiphase Flow Technology and Consequences for Field Development Conference, 229–251, Stavanger, Norway (1987).

Shoham, O., "Flow Pattern Transitions and Characterization in Gas-Liquid Two Phase Flow in Inclined Pipes." Ph.D. Thesis, Tel-Aviv University, Ramat-Aviv, Israel (1982).

Sinquin, A., Palermo, T., and Peysson, Y., Rheological and flow properties of gas hydrate suspensions. *Oil Gas Sci. Technolo. Rev. IFP.* **59**(1), 41–57 (2004).

Sloan, E.D., Jr., Natural gas hydrates. *J. Petro. Technol.* **43**, 1414 (1991).

Sloan, E.D., Jr., “Clathrate Hydrates of Natural Gases,” 2nd Ed. Dekker, New York (1998).

Sloan, E.D., Jr., “Hydrate Engineering.” SPE Monograph, 21, Society of Petroleum Engineers (SPE) Publications, Richardson, TX (2000).

Smart, J., A method for calculating the corrosion allowance for deepwater pipelines and risers. *J. Pipeline Integrity* **73** (2001).

Soldati, A., Paglianti, A., and Giona, M., Identification of two-phase flow regimes via diffusional analysis of experimental time series. *Exp. Fluids* **21**, 151–160 (1996).

Son, V.K., and Wallace, C., “Reclamation/Regeneration of Glycols Used for Hydrate Inhibition.” Paper presented at the 12th Annual Deep Offshore Technology Conference and Exhibition, New Orleans, LA (Nov. 7–9, 2000).

Sotberg, T., and Bruschi, R., “Future Pipeline Design Philosophy-Framework.” Paper presented at the 11th International Conference on Offshore Mechanics and Arctic Engineering, Calgary, AB, Canada (1992).

Storkaas, E., Alstad, V., and Skogestad, S., “Stabilization of Desired Flow Regimes in Pipelines.” Paper presented at the AIChE Annual Meeting, Reno, NV (2001).

Strommen, R.D., Horn, H., and Wold, K.R., New technique monitors pipeline corrosion, cracking. *Oil Gas J.* **91**(52), 88–92 (1993).

Taitel, Y., Stability of severe slugging. *Int. J. Multiphase Flow* **12**(2), 203–217 (1986).

Taitel, Y., “Flow Pattern Transition in Two-Phase Flow.” Paper presented at the 9th International Heat Transfer Conference, Jerusalem (1990).

Taitel, Y., Barnea, D., and Brill, J.P., Stratified three-phase flow in pipes. *Int. J. Multiphase Flow*. **21**(1), 53–60 (1995).

Taitel, Y., Barnea, D., and Dukler, A. E., Modeling flow pattern transitions for steady upward gas-liquid flow in vertical tubes. *AIChE J.* **26**(3), 345–354 (1980).

Taitel, Y., and Dukler, A.E., A model for predicting flow regime transitions in horizontal and near horizontal gas-liquid flow. *AIChE J.* **22**(1), 47 (1976).

Taitel, Y., Shoham, O., and Brill, J.P., Simplified transient solution and simulation of two-phase flow in pipelines. *Chem. Eng. Sci*. **44**(6), 1353–1359 (1989).

Tek, M.R., Multiphase flow of water, oil, and natural gas through vertical flow strings. Journal of Petroleum Technology (JPT), 1029–1036 (1961).

Tin, V., "Severe Slugging in Flexible Risers." Proceeding of the 5th BHRG International Conference on Multiphase Production, 507–525, Cannes, France (1991).

Tiratsoo, J.N.H., "Pipeline Pigging Technology," 2nd Ed. Gulf Professional Publishing, Houston, TX (1992).

Towler, B.F., and Mokhatab, S., Quickly estimate hydrate formation conditions in natural gases. *Hydrocarb. Proc*. **84**(4), 61–62 (2005).

Trekell, R.E., and Campbell, J.M., "Prediction of the Behavior of Hydrocarbon Clathrate Solutions." Proceedings of the 151st Meeting of American Chemical Society, Petroleum Chemistry Division, 61, Pittsburgh, PA (March 1966).

Uhl, A.E., "Steady Flow in Gas Pipelines." IGT Report No. 10, American Gas Association (AGA), New York, (1965).

Uhlig, H.H., and Revie, R.W., "Corrosion and Corrosion Control," 3rd Ed. Wiley, New York (1985).

Ullah, M.R., A comprehensive assessment of the performance of dry-gas methods for predicting the occurrence of wet-gas flow conditions. *J. Pipelines* **6** (1987).

Van Hout, R., Shemer, L., and Barnea, D., Spatial distribution of void fraction within the liquid slug and some other related slug parameters. *Int. J. Multiphase Flow* **18**, 831–845 (1992).

Varughese, K., In-situ pipeline rehabilitation techniques, equipment improved. *Oil Gas J.* **91**(25), 54–57 (1993).

Verley, R., *et al*., "Wall Thickness Design for High Pressure Offshore Gas Pipelines." Paper presented at the 13th International Conference on Offshore Mechanics and Arctic Engineering, Houston, TX (Feb. 27–March 3, 1994).

Wallis, G.B., "One Dimensional Two-Phase Flow." McGraw-Hill, New York (1969).

Wilkens, R.J., "Prediction of the Flow Regime Transitions in High Pressure, Large Diameter, Inclined Multiphase Pipelines." Ph.D. Thesis, Ohio University, Athens, OH (1997).

Wilkens, R.J., Flow assurance. *In* "Fluid Flow Handbook" (J. Saleh, ed.), Chapter 29. McGraw-Hill, New York (2002).

Xiao, J.J., Shoham, O., and Brill, J.P., "A Comprehensive Mechanistic Model for Two-Phase Flow in Pipelines." Paper presented at the 65th SPE Annual Technical Conference and Exhibition, New Orleans, LA (Sept. 23–26, 1990).

Xiao, J.J., and Shoup, G., Sizing wet-gas pipelines and slug catchers with steady-state multiphase flow simulations. *ASME J. Energy Res. Technol.* **120**, 106–110 (1998).

Yeung, H.C., and Lima, P.C.R., Modeling of pig assisted production methods. *ASME J. Energy Res. Technol.* **124**, 8–13 (2002).

Yocum, B.T., “Offshore Riser Slug Flow Avoidance: Mathematical Model for Design and Optimization.” Paper presented at the SPE London Meeting, London, UK (1973).

Zhang, H.-Q., Wang, Q., Sarica, C., and Brill, J.P., Unified model for gas-liquid pipe flow via slug dynamics. 1. Model development. *ASME J. Energy Res. Technol.* **125**(4), 266–273 (2003).

Zuo, J.Y., and Zhang, D.D., “Gas Hydrate Formation in Aqueous Solutions Containing Methanol and Electrolytes.” Proceedings of the Laurance Reid Gas Conditioning Conference, 191–198, Norman, OK (Feb. 21–24, 1999).

CHAPTER 4

BASIC CONCEPTS OF NATURAL GAS PROCESSING

4.1 INTRODUCTION

Raw natural gas after transmission through the field-gathering network must be processed before it can be moved into long-distance pipeline systems for use by consumers. The objective of gas processing is to separate natural gas, condensate, noncondensable, acid gases, and water from a gas-producing well and condition these fluids for sale or disposal. The typical process operation modules are shown in Figure 4-1. Each module consists of a single piece or a group of equipment performing a specific function. All the modules shown will not necessarily be present in every gas plant. In some cases, little processing is needed; however, most natural gas requires processing equipment at the gas processing plant to remove impurities, water, and excess hydrocarbon liquid and to control delivery pressure. The unit operations used in a given application may not be arranged in the sequence shown in Figure 4-1, although this sequence is typical. The choice of modules to be used and the arrangement of these modules are determined during the design stage of each gas-field development project (Meyer and Sharma, 1980).

Unfortunately, at the individual module level the design may be sound and the operation correct but could result in a poor gas processing facility. The reason is that each module has varying characteristics under varying loads, which can result in a type of internal incompatibility. For instance, a given unit module may require a particular inlet composition to produce

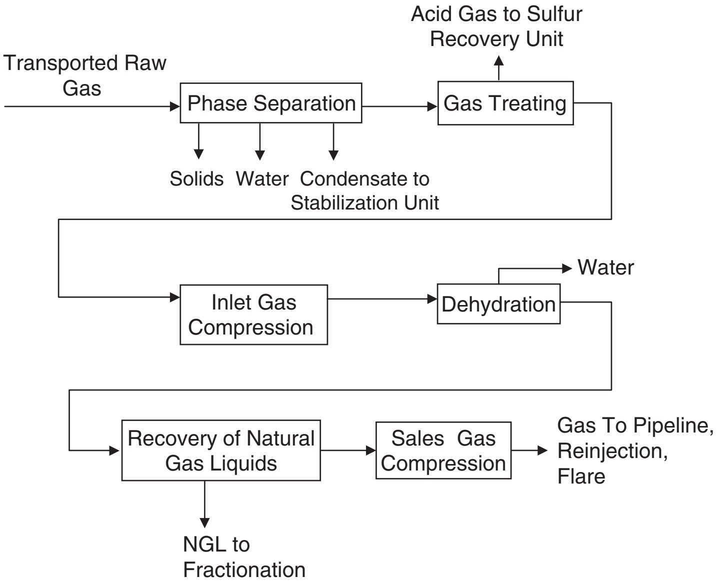

Figure 4-1. Simplified typical onshore treatment process.

the desired output. If a previous unit does not maintain this, the downstream unit may not operate satisfactorily. Thus, the fault might not lie so much with that unit but with total plant design, even though the unit module is usually blamed.

The individual unit modules of Figure 4-1 are briefly reviewed here with greater details to follow in subsequent chapters.

4.2 PROCESS MODULES

The first unit module is the physical separation of the distinct phases, which are typically gas, liquid hydrocarbons, liquid water, and/or solids. Phase separation of the production stream is usually performed in an inlet separator. Inlet gas receiving is complicated by the fact that transmission lines supplying the plant typically operate with two or three phases present and consequently liquid slugging is common. Slugs are normally formed from elevational changes in the inlet supply pipes, changes in gas supply flow rates, and changes in pressure and temperature during transmission. Slug flow may even be encountered in horizontal pipes under steady-state conditions if the flow regime is not properly selected. The arrival of “slugs” at production or processing equipment impacts the operation of production facilities negatively, causing both mechanical problems (due to high velocities and momentum) and process problems (increasing liquid levels, causing surges and trips). In some cases operators can minimize liquid accumulation by managing fields and pipelines in such a way as to create a suitable fluid flow regime (i.e., mist flow regime) in which the gas velocity is high enough to keep liquids dispersed continuously. While it is desirable to design the flow lines to avoid slugging, in practice this can be difficult while maintaining the ability to turn down the pipeline flow rate. In these cases, consideration should be given to providing suitable process equipment to diminish the effect of slugging. Gas pipelines have typically used slug catchers to dissipate the energy of the liquid slugs, to minimize turbulence, to ensure that the gas and liquid flow rates are low enough so that the stratified flow regime and subsequently gravity segregation can occur. The slug catcher is designed to separate gas, hydrocarbon condensate, and inlet water. The gas stream is sent to the inlet separators. The separators usually contain vane elements to aid in the coalescence of liquids. They may also include filters to remove particulates and may be followed by suction scrubbers if compressors are needed to bring low-pressure gas up to high pressure for further processing operations. The liquids that collect in the slug catcher flow to a three-phase separator from which the two liquid phases, hydrocarbon condensate and water/methanol or water/glycol phases, are outputs. Overhead gas from the three-phase separator is recompressed where necessary for use as fuel gas. More detailed information about phase separation is presented in Chapter 5.

Hydrocarbon condensate recovered from natural gas may be shipped without further processing but is typically stabilized to produce a safe-transportable liquid. Unstabilized condensates contain a large percentage of methane and ethane, which will vaporize easily in storage tanks. Stabilization is the full removal of light fractions from the condensate, usually achieved by distillation. Stabilized liquid will generally have a vapor pressure specification (Reid vapor pressure[1] of <10 psi), as the product will be injected into a pipeline or transport pressure vessel, which

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[1] Vapour pressure at 100°F. Reid vapor pressure is a standard indicator of volatility and a key quality control factor, particularly when considering condensate storage.

has definite pressure limitations. Condensate stabilization is discussed in Chapter 6.

The next step in natural gas processing is acid gas treating. In addition to heavy hydrocarbons and water vapor, natural gas often contains other contaminants that may have to be removed. Carbon dioxide (CO_2), hydrogen sulfide (H_2S), and other sulfur-containing species such as mercaptans are compounds that require complete or partial removal. These compounds are collectively known as "acid gases." H_2S when combined with water forms a weak sulfuric acid, whereas CO_2 and water form carbonic acid, thus the term "acid gas." Natural gas with H_2S or other sulfur compounds present is called "sour gas," whereas gas with only CO_2 is called "sweet." Both H_2S and CO_2 are very undesirable, as they cause corrosion and present a major safety risk. Treating processes for receiving these components are covered in Chapter 7.

Depending on the pressure at the plant gate, the next step in processing will either be inlet compression to an "interstage" pressure, typically 300–400 psig (compression is discussed in Chapter 8), or be dew point control and natural gas liquid recovery. Water dew point control is required to meet specifications and to control hydrate formation. Gas hydrate formation is a major concern for engineers in pipeline and natural gas transportation industries as it causes choking/plugging of pipelines and other related problems. Methods of preventing hydrate formation in the plant include lowering the hydrate formation temperature with chemical inhibition or dehydration to remove the water. Gas dehydration is discussed in detail in Chapter 9.

Hydrocarbon dew point or hydrocarbon liquid recovery involves cooling the gas and condensing out the liquids. Hydrocarbon dew point control can be either dehydration followed by cooling/condensation or by a combination of inhibition/cooling/condensation processes. Refrigeration is performed either by autorefrigeration due to a pressure drop across a valve or by an external mechanical refrigeration process. The temperature to which the gas is cooled depends on whether it is necessary to meet a sales gas hydrocarbon dew point specification or whether substantial liquid recovery is desired. Three situations motivate maximum condensate recovery. The first is the desire to maximize condensate production when processing associated gas. The second situation occurs when processing retrograde condensate gas; here the objective is to recover the condensate and reinject the gas into the formation. Third, in some markets the natural gas liquids (NGLs) produced from the condensate may be more valuable as liquid products than as sales gas components, i.e., their recovery will yield a better profit. Whether to leave maximum NGLs in the gas stream (but still attaining sales hydrocarbon dew point specification) or to recover them as liquids is purely an economic decision made by comparing their value as heat versus the equivalent value as liquid chemical feedstock. If the equivalent liquid value is lower than the gas, NGLs should be left in the gas to the extent as possible. However, if the equivalent liquid value is higher than the gas value, then liquid recovery should be maximized. Natural gas processing for liquid recovery is discussed in Chapter 10.

If gas is produced at lower pressures than typical sales pipeline pressure (approximately 700–1000 psig), it is compressed to sales gas pressure. Transport of sales gas is done at high pressure in order to reduce pipeline diameter. Pipelines may operate at very high pressures (above 1000 psig) to keep the gas in the dense phase thus preventing condensation and two-phase flow. Compression typically requires two to three stages to attain sales gas pressure. As stated previously, processing may be done after the first or second stage, prior to sales compression. More details are available in Chapter 11.

Where there is no available gas pipeline, separated associated gas may be flared. The ability to flare depends on regulations as well as the field location. Increasingly in such cases, separated gas is being conserved by compression and reinjection into producing formations for eventual recovery and sales. Also, in gas condensate reservoirs, the gas is often reinjected, or “cycled,” to enable higher net recovery of valuable liquid hydrocarbons from the reservoir.

4.3 SCOPE OF NATURAL GAS PROCESSING

The important factors that usually determine the extent of gas processing include the processing objectives, the type or source of the gas, and the location and size of the gas fields.

4.3.1 Processing Objectives

If the natural gas is transportated by pipeline, the processing installation must be designed to meet either transport or final specifications. Processing of a gas stream may have one of the following three basic objectives (Odello, 1981; Rojey *et al*., 1997).

- To produce a sales gas stream that meets specifications of the type shown in Table 4-1. These specifications are mainly intended to meet pipeline requirements and the needs of industrial and domestic consumers.
- To maximize NGLs production by producing a lean gas stripped of most of the hydrocarbons other than methane.
- To deliver a commercial gas. Such gas must be distinguished by a certain range of gross heating value lying.

4.3.2 Effect of Gas Type in Field Processing

The gas composition of the field is the most important issue in choosing a processing scheme. In other words, depending on the type of reservoir and the composition of the produced gas, the gas processing plant may contain extensive facilities for the processing of the associated liquefiable hydrocarbons. Typically, associated gas is very rich in liquefiable hydrocarbons and must undergo NGL and condensate recovery to meet hydrocarbon dew point or minimum heating value requirements. The gas processing scheme will also be dictated by the format of the sales contract and its specifications. The contract may be totally different for each customer depending on the composition and amount of gas, plant recoveries, and the contractual preferences of the customer (Bullin and Hall, 2000).

Table 4-1
Natural Gas Specifications in the Salable Gas Stream (Goar and Arrington, 1978)

Characteristic	Specification
Water content	4–7 lb/MMscf (max)
Hydrogen sulfide content	1/4 grain/100 scf (max)
Gross heating value	950 Btu/scf (min)
Hydrocarbon dew point	15°F at 800 psig (max)
Mercaptan content	0.2 grain/100 scf (max)
Total sulfur content	1–5 grain/100 scf (max)
Carbon dioxide content	1–3 mole percent (max)
Oxygen content	0–0.4 mole percent (max)
Sand, dust, gums, and free liquid	Commercially free.
Typical delivery temperature	120°F
Typical delivery pressure	714.7 psia

4.3.3 Location of the Gas Field

The productivity of a gas reservoir can vary greatly and depend primarily on type, location, and age. Because the location and output of the wells can vary widely, then not surprisingly, the systems that have been designed to collect and process this output also vary widely (Thorn *et al*., 1999). There are at least two aspects of location that are important: remoteness and local temperature variation. Temperature affects the tendency for hydrate formation in the gas gathering network. Offshore platforms and "outbacks" are examples of remote locations. Even these locales are not strictly comparable because one is sea based vs dry land based. For the sea-based facility the produced fluid from each wellhead flows via a flow line into a manifold and from there to the process facilities located on the platform deck. Ship platforms are extremely limited with respect to size and allowable weight and only those operations absolutely needed are performed. Facilities on the offshore platform will generally process the gas to produce a low water content hydrocarbon stream for export to shore through the subsea pipelines. This process ensures minimal corrosion, as well as minimizing the potential for hydrate formation in the raw gas pipeline. A dry-land outback facility has essentially unlimited area available and can support operations not practical or desirable offshore, such as treating or processing involving fire hazards (Manning and Thompson, 1991).

REFERENCES

Bullin, K.A., and Hall, K.R., "Optimization of Natural Gas Processing Plants Including Business Aspects." Paper presented at the 79th GPA Annual Convention, Atlanta, GA (2000).

Campbell, J.M., "Gas Conditioning and Processing," 3rd Ed. Campbell Petroleum Series, Norman, OK (1992).

Goar, B. G., and Arrington, T. O., Processing sour Gas. 1. Guidelines for handling sour gas. *Oil Gas J*. **76**(26), 160–164 (June 1978).

Huzyk, S.L., Anschutz ranch facilities expansion ups production. *Oil Gas J*. **86**(24), 35–38 (June 1988).

Manning, F.S., and Thompson, R.E., "Oil Field Processing of Petroleum, Vol. 1. Pennwell Publishing Company, Tulsa, OK (1991).

Meyer, P.E., and Sharma, S.C., "Field Production Systems and Oil Processing." Oil, Gas and Petroleum Equipments (Sept and Oct. 1980).

Odello, R., Systematic method aids choice of field gas treatment. *Oil Gas J*. **79**(6), 103–109 (Feb. 1981).

Rojey, A., Jaffret, C., Cornot-Gandolph, S., Durand, B., Jullin, S., and Valais, M., "Natural Gas Production, Processing, Transport." Editions Technip, Paris, France (1997).

Sivalls, C.R., Fundamentals of oil production processing, *In* "Proceedings of the 30th Annual Southwestern Petroleum Short Course," Midland, TX (April 1983).

Thorn, R., Johansen, G.A., and Hammer, E.A., "Three-Phase Flow Measurement in the Offshore Oil Industry: Is There a Place for Process Tomography?" Paper presented at the 1st World Congress on Industrial Process Tomography, Buxton, Greater Manchester (April 14–17, 1999).

CHAPTER 5

PHASE SEPARATION

5.1 INTRODUCTION

Separation of oil and gas is a critical field processing operation. As producing pressure is increased and lighter condensates are produced, efficient separation has become more critical than ever. Moreover, some of the new concepts in separation technology have been applied to advantage on old lease producing oil at moderate or low pressures. As gas transmission lines raise their standards, separation becomes a part of the overall field processing necessary to condition the gas. Several technologies are available to remove liquids and solids from gases. However, selecting gas/liquid separation technologies requires not only knowledge of the process conditions, but a knowledge of the characteristics of the liquid contaminants. Selection should be made based on the droplet size, concentration, and whether the liquid has waxing or fouling tendencies (Brown *et al*., 1994). Before evaluating specific technologies, it is important to understand the mechanisms used to remove liquids and solids from gases. Three principles used to achieve physical separation of gas and liquids or solids are momentum, gravity settling, and coalescing. Any separator may employ one or more of these principles; however, the fluid phases must be immiscible and have different densities for separation to occur. Momentum force is utilized by changing the direction of flow and is usually employed for bulk separation of the fluid phases. The gravitational force is utilized by reducing velocity so the liquid droplets can settle out in the space provided. Gravity segregation is the main force that accomplishes the separation, which means the heaviest fluid settles to the bottom and the lightest fluid rises to the top. However, very small droplets such as mist cannot be separated practically by gravity. These droplets can be coalesced to form larger droplets that will settle by gravity. The purpose of this chapter is to review the principles governing the basic separation process and associated equipment design procedure.

5.2 GRAVITY SEPARATORS

Gravity separators are pressure vessels that separate a mixed-phase stream into gas and liquid phases that are relatively free of each other. In a gravity separator, gravitational forces control separation, and the efficiency of the gas/liquid separation is increased by lowering the gas velocity. Because of the large vessel size required to achieve settling, gravity separators are rarely designed to remove droplets smaller than 250 μm (Talavera, 1990). However, an analysis of this type of separator is given because it is useful to help understand the settling mechanism of other separators.

Gravity separators are often classified by their geometrical configuration (vertical, horizontal) and by their function (two-phase/three-phase separator). In other words, gravity separators are classified as “two phase” if they separate gas from the total liquid stream and “three phase” if they also separate the liquid stream into its crude oil and water-rich phases. Additionally, separators can be categorized according to their operating pressure. Low-pressure units handle pressures of 10 to 180 psi. Medium-pressure separators operate from 230 to 700 psi. High-pressure units handle pressures of 975 to 1500 psi.

Separators are sometimes called “scrubbers” when the ratio of gas rate to liquid rate is very high. These vessels usually have a small liquid collection section and are recommended only for the following items.

- Secondary separation to remove carryover fluids from process equipment such as absorbers and liquid dust scrubbers.
- Gas line separation downstream from a separator and where flow lines are not long.
- Miscellaneous separation where the gas–liquid ratio is extremely high.

In any case, these equipments have the same configuration and are sized in accordance with the same procedure of separators.

5.2.1 General Description

All gravity separators normally have the following components or features (API Spec 12J, 1989).

- A primary gas/liquid separation section with an inlet divertor to remove the bulk of the liquid from the gas.
- A gravity-settling section providing adequate retention time so that proper settling may take place.
- A mist extractor at the gas outlet to capture entrained droplets or those too small to settle by gravity.
- Proper pressure and liquid-level controls.

Gravity separators are designed as either horizontal or vertical pressure vessels. Figure 5-1 is a typical scheme of a three-phase horizontal separator. The fluid enters the separator and hits an inlet diverter. This sudden change in momentum generates the initial bulk separation of liquid and gas. In most designs, the inlet diverter contains a downcomer that directs the liquid flow below the oil/water interface. This forces the inlet mixture of oil and water to mix with the water continuous phase in the bottom of the vessel and rise through the oil/water interface. This process is called "water washing" and promotes the coalescence of water droplets that are

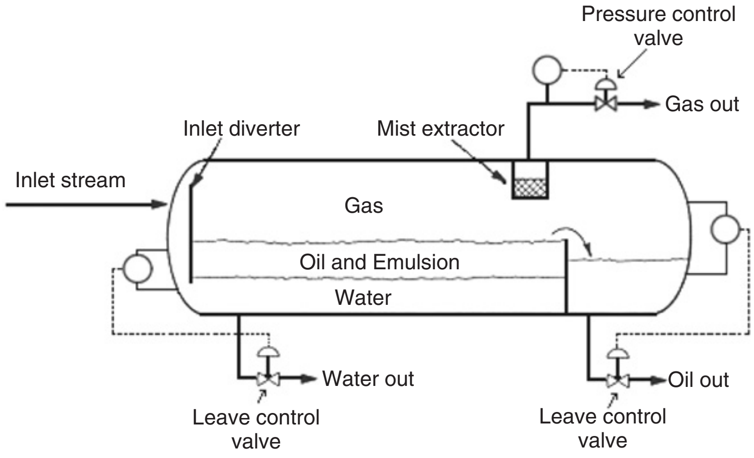

Figure 5-1. A typical scheme of a horizontal three-phase separator.

entrained in the oil continuous phase. The inlet diverter assures that little gas is carried with the liquid, and the water wash assures that the liquid does not fall on top of the gas/oil or oil/water interface, mixing the liquid retained in the vessel and making control of the oil/water interface difficult. The liquid-collecting section of the vessel provides sufficient time so that the oil and emulsion form a layer or "oil pad" at the top. The free water settles to the bottom. The produced water flows from a nozzle in the vessel located upstream of the oil weir. An interface level controller senses the height of the oil/water interface. The controller sends a signal to the water dump valve, thus allowing the correct amount of water to leave the vessel so that the oil/water interface is maintained at the design height.

The gas flows horizontally and outs through a mist extractor (normally known as a demisting device) to a pressure control valve that maintains constant vessel pressure. The level of the gas/oil interface can vary from half the diameter to 75% of the diameter depending on the relative importance of liquid/gas separation and what purpose the separator has. For example, degassers and produced water flash drums have very high liquid/gas interfaces. However, the most common configuration is half full.

Figure 5-2 shows a typical configuration for a vertical three-phase separator. In the vertical separator, the flow enters the vessel through the side as in the horizontal separator and the inlet diverter separates the bulk of the gas. The gas moves upward, usually passing through a mist extractor to remove suspended mist, and then the dry gas flows out. A downcomer is required to transmit the liquid collected through the oil–gas interface so as not to disturb the oil-skimming action taking place.

A chimney is needed to equalize gas pressure between the lower section and the gas section. The spreader or downcomer outlet is located at the oil–water interface. From this point as the oil rises any free water trapped within the oil phase separates out. The water droplets flow countercurrent to the oil. Similarly, the water flows downward and oil droplets trapped in the water phase tend to rise countercurrent to the water flow. It should be clear that the principles of operation (such as oil/water interface level controlling) of three-phase vertical separators are the same as the three-phase horizontal separators described earlier. Essentially, the only difference is that horizontal separators have separation acting tangentially to flow, whereas vertical separators have separation acting parallel to flow. In the vertical separator, level control is not also critical, where the liquid level can fluctuate several inches without affecting operating

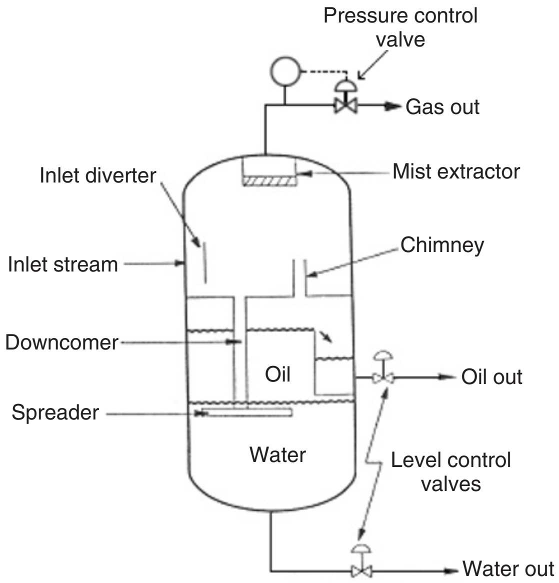

Figure 5-2. A typical scheme of a vertical three-phase separator.

efficiency (GPSA, 1998) However, it can affect the pressure drop for the downcomer pipe (from the demister), therefore affecting demisting device drainage.

5.2.2 Separator Selection

There are no simple rules for separator selection. Sometimes, both configurations should be evaluated to decide which is more economical. The relative merits and common applications of vertical and horizontal separators are summarized by Manning and Thompson (1995).

5.2.2.1 Horizontal Separators

Horizontal separators are used most commonly in the following conditions.

- Large volumes of gas and/or liquids.
- High-to-medium gas/oil ratio (GOR) streams.

- Foaming crudes.
- Three-phase separation.

Advantages and disadvantages of these separators are as follow.

Advantages

- Require smaller diameter for similar gas capacity as compared to vertical vessels.
- No counterflow (gas flow does not oppose drainage of mist extractor).
- Large liquid surface area for foam dispersion generally reduces turbulence.
- Larger surge volume capacity.

Disadvantages

- Only part of shell available for passage of gas.
- Occupies more space unless "stack" mounted.
- Liquid level control is more critical.
- More difficult to clean produced sand, mud, wax, paraffin, etc.

5.2.2.2 Vertical Separators

These separators are used in the following conditions.

- Small flow rates of gas and/or liquids.
- Very high GOR streams or when the total gas volumes are low.
- Plot space is limited.
- Ease of level control is desired.

Advantages and disadvantages of these separators are as follow.

Advantages

- Liquid level control is not so critical.
- Have good bottom-drain and clean-out facilities.
- Can handle more sand, mud, paraffin, and wax without plugging.
- Less tendency for reentrainment.
- Has full diameter for gas flow at top and oil flow at bottom.
- Occupies smaller plot area.

Disadvantages

- Require larger diameter for a given gas capacity, therefore, most competitive for very low GOR or very high GOR or scrubber applications.
- Not recommended when there is a large slug potential.
- More difficult to reach and service top-mounted instruments and safety devices.

5.2.3 Gravity Separation Theory

Vapor/liquid separation is usually accomplished in three stages. The first stage, primary separation, uses an inlet diverter[1] to cause the largest droplets to impinge by momentum and then drop by gravity. The next stage, secondary separation, is gravity separation of smaller droplets as the vapor flows through the disengagement area. Gravity separation can be aided by utilizing distribution baffles that create an even velocity distribution in the fluid, thus allowing enhanced separation. The final stage is mist elimination, where the smallest droplets are coalesced on an impingement device, such as a mist pad or vane pack, followed by gravity settling of the larger formed droplets.

In the gravity-settling section of the separators, the liquid drops will settle at a velocity determined by equating the gravity force (F_B) on the drop with the drag force (F_D) caused by its motion relative to the vapor continuous phase. When the drag force is equal to the buoyancy (gravity) force, the droplet acceleration is zero so that it moves at a constant velocity. This velocity is the terminal or free settling velocity, which is determined with respect to the following equations.

$$F_B = \left(\frac{\pi}{6}\right) D_d^3 \left(\rho_L - \rho_V\right) \left(\frac{g}{g_c}\right) \tag{5-1}$$

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[1] Inlet diverters are very old technology now and are used very rarely. Revamps in the north sea replace these inlet devices all the time, especially where the asset is producing more throughput than originally expected. If sized, incorrectly, inlet devices can cause serious separation issues due to droplet shatter. Schoepentoeters became the popular inlet device for a while, but these were designed for vertical separators and are therefore not always applicable. Cyclonic and distribution baffle inlet devices are more common nowadays.

where D_d is drop diameter, ft; ρ_L is liquid density, lbm/ft^3; ρ_V is vapor density, lbm/ft^3; g is gravitional constant, 32.174 ft/sec^2; and g_c is conversion factor, 32.174 lbm-ft/sec^2-lbf.

Also, the drag force on the droplet is given by

$$F_D = C_D \left(A_p\right) \left(\frac{\rho_V V_d^2}{2g_c}\right) \tag{5-2}$$

where C_D is drag coefficient, dimensionless; A_p is projected drop area, ft^2; $= (\pi/4)\, D_d^2$ (area of circle, not sphere); and V_d is drop velocity, ft/sec.

Therefore, the terminal settling velocity of the liquid droplets (V_t) can be determined by equating Equations (5-1) and (5-2) as follow:

$$V_t = \sqrt{(4/3) D_d (\rho_L - \rho_V) g / (C_D \rho_V)} \tag{5-3}$$

The drag coefficient can also be calculated as follows (Svrcek and Monnery, 1993):

$$C_D = \frac{5.0074}{\ln(x)} + \frac{40.927}{\sqrt{X}} + \frac{484.07}{X} \tag{5-4}$$

and

$$X = \frac{0.95 \times 10^8 D_d^3 \rho_V (\rho_L - \rho_V)}{\mu_V^2} \tag{5-5}$$

where D_d is in ft, densities are in lb/ft^3, and viscosity is in cP.

The droplet-settling velocity equation considers the escape of a drop from the continuous phase (e.g., the escape of an oil drop from the gas phase). For this purpose, the droplet-settling velocity must be greater than the superficial upward bulk vapor velocity, V_V. Typically, the allowable vapor velocity is set between 0.75 V_t and V_t (Svrcek and Monnery, 1993). Equation (5-3) can be rearrenged as a Sauders and Brown (1934) type equation as follows:

$$V_t = K_{SB} \sqrt{(\rho_L - \rho_V)/\rho_V} \tag{5-6}$$

where

$$K_{\mathrm{SB}} = \sqrt{\frac{4gD_{\mathrm{d}}}{3C_{\mathrm{D}}}} \tag{5-7}$$

In practice, the Souders and Brown design coefficient (K_{SB}) depends primarily on the type of mist extractor present, separator geometry, flow rates, and fluid properties. Therefore, K_{SB} is usually determined from experiments. A well-known source of empirical K_{SB} factors for mist pads is the GPSA (1998) Engineering Data Book. The GPSA's K factors have been curve fitted and are given as

$$K_{\mathrm{SB}}\ (P, \mathrm{psig}) = 0.35 - 0.0001\ (P - 100) \tag{5-8}$$

Also, the factor K_{SB} should be adjusted as follows:

1. For most vapors under vacuum, $K_{\mathrm{SB}} = 0.20$.
2. For glycols and amine solutions, multiply K_{SB} values by 0.6–0.8.
3. For compressor suction scrubbers and expander inlet separators, multiply K_{SB} by 0.7–0.8.

Maximum terminal velocities calculated using the K_{SB} factors are for separators normally having a wire-mesh mist extractor and should allow all liquid droplets larger than 10 μm to settle out of the gas. If no mist extractor is present, multiply K_{SB} by 0.5.

It is often necessary to separate two immiscible liquids, light and heavy phases, and a vapor. A typical example in the petroleum industry is the separation of water, and a hydrocarbon liquid and vapor. For this system, the flow of rising light droplets in the heavy liquid phase or setteling heavy droplets in the light liquid phase is considered laminar and is governed by Stokes law (Monnery and Svrcek, 1994):

$$V_{\mathrm{t}} = \frac{K_{\mathrm{s}}\,(\rho_{\mathrm{H}} - \rho_{\mathrm{L}})}{\mu} \tag{5-9}$$

$$K_{\mathrm{s}} = 2.06151 \times 10^{-5} D_{\mathrm{d}}^{2} \tag{5-10}$$

where V_{t} is in inch/min, densities of light and heavy liquid phases ($\rho_{\mathrm{L}}, \rho_{\mathrm{H}}$) are in lb/ft^3, viscosity is in cP, and D_{d} is in μm (1 μm $= 3.28084 \times 10^{-6}$ ft).

As can be seen from Equation (3-14), the settling velocity of a droplet is inversely proportional to the viscosity of the continous phase. Hence, it requires more time for the droplets to settle out of the continuous phase with greater viscosity. In practice, V_t is typically limited to 10 inch/min.

5.2.4 Design Considerations

The following factors must be determined before beginning separator design.

- Gas and liquids flow rates (minimum, average, and peak).
- Operating and design pressures and temperatures.
- Surging or slugging tendencies of the feed streams.
- Physical properties of the fluids, such as density, viscosity, and compressibility.
- Designed degree of separation (e.g., removing 100% of particles greater than 10 μm).

Consideration for the future life of the field should also be included. For example, most north sea separators were designed for high oil cuts, but are now high water cuts, which produce a lot of nozzle problems. In the separator design, it is also worthwhile to clarify two definitions: holdup and surge times. Holdup is defined as the time it takes to reduce the liquid level from normal (NLL) to low (LLL) while maintaining a normal outlet flow without feed makeup. Surge time is defined as the time it takes for the liquid level to rise from normal (NLL) to high (HLL) while maintaining a normal feed without any outlet flow. Holdup time is based on the reserve required to maintain good control and safe operation of downstream facilities, whereas surge time is usually based on requirements to accumulate liquid as a result of upstream or downstream variations or upsets, for e.g., slugs. In the absence of specific requirements, surge time may be taken as one-half of holdup time. Table 5-1 shows typical values of holdup time, t_H, and surge time, t_S (Svrcek and Monnnery, 1993).

5.2.5 Design Procedure

The initial design and calculation of gravity separators are discussed in many books and basic references (Kumar, 1987; Campbell, 1992; Arnold and Stewart, 1998). However, a more accurate and detailed sizing of two- and three-phase separators can be performed using the design methods developed by Svrcek and Monnery (1993) and Monnery and

Table 5-1
Typical Values of Holdup and Surge Times

Service	t_H, min	t_S, min
A. Unit feed drum	10	5
B. Separators		
1. Feed to column	5	3
2. Feed to other drum or tankage with pump or through exchange	5	2
without pump	2	1
3. Feed to fired heater	10	3

Svrcek (1994), which have been well received by the industry worldwide. These procedures are a result of a review of literature sources and accepted industrial design guidelines and allow the production facility engineer to choose the detailed sizing parameters of two- and three-phase separators. This section attempts to address the basics of three-phase separator design and to provide step-by-step procedures for three-phase vapor/liquid/liquid separator design. To add a degree of conservatism to the design, the volume available in the heads is ignored. It is reasonably common nowadays for vessel sizing to be subcontracted out either directly to the vessel supplier or sometimes to the internals supplier.

5.2.5.1 Vertical Separators

For a three-phase vertical separator, the total height can be broken into different sections, as shown in Figure 5-3. The separator height is then calculated by adding the heights of these sections. If a mist eliminator pad is used, additional height is added.

The calculations of diameter and height are detailed as follow.

1. Calculate the vertical terminal velocity using Equation (5-6) and set $V_V = 0.75\ V_t$ for a conservative design.
2. Calculate the vapor volumetric flow rate:

$$Q_V = \frac{\dot{m}_V}{3600\rho_V} \tag{5-11}$$

3. Calculate the vessel internal diameter, D_i:

$$D_i = \left(\frac{4Q_V}{\pi V_V}\right)^{0.5} \tag{5-12}$$

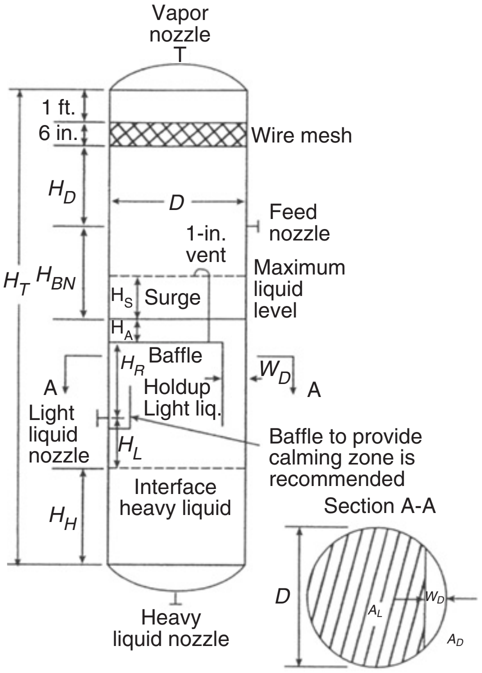


Figure 5-3. Basic design of three-phase vertical separator (Monnery and Svrcek, 1994).

If there is a mist eliminator, add 3–6 inch to D_i to accommodate a supporting and round up to the next 6-inch increment to obtain D; if there is no mist eliminator, $D = D_i$.

4. Calculate the settling velocity of the heavy liquid out of the light liquid using Equation (5-9) (with $\mu = \mu_L$).
5. Calculate the rising velocity of the light liquid out of the heavy liquid phase using Equation (5-9) (with $\mu = \mu_H$).
6. Calculate the light and heavy liquid volumetric flow rates, Q_{LL} and Q_{HL}:

$$Q_{LL} = \frac{\dot{m}_{LL}}{60\rho_L} \tag{5-13}$$

$$Q_{HL} = \frac{\dot{m}_{HL}}{60\rho_H} \tag{5-14}$$

7. Assume $H_L = 1$ ft (minimum) and $H_H = 1$ ft (minimum) and calculate the settling times for the heavy liquid droplets to settle ($t_{s,HL}$) and for the light liquid droplets to rise ($t_{s,LL}$) through this section, respectively, as

$$t_{s,HL} = \frac{12H_L}{V_{HL}} \tag{5-15}$$

$$t_{s,LL} = \frac{12H_H}{V_{LL}} \tag{5-16}$$

   where the settling velocity of heavy liquid droplets out of light liquid (V_{HL}) and the rising velocity of light liquid droplets out of heavy liquid (V_{LL}) are in inch/min.

8. If there is a baffle plate, calculate the area, A_L, as

$$A_L = A - A_D \tag{5-17}$$

   where A is vertical vessel cross-sectional area, ft^2 ($= \frac{\pi D^2}{4}$) and A_D is downcomer cross-sectional area, ft^2.

   In the equation just given, the larger value of A_D that can be calculated from two following equations should be used.

$$A_D = \left(\frac{7.48\ \text{gal}}{\text{ft}^3}\right)\left(\frac{60\ \text{min}}{1\ \text{hr}}\right)\left(\frac{Q_{LL} + Q_{HL}}{\text{G}}\right) \tag{5-18}$$

   where baffle liquid load (G) will be obtained from Figure 5-4.

$$\begin{aligned}\frac{A_D}{A} = &\left\{-4.755930 \times 10^{-5} + 0.174875\left(\frac{W_D}{W}\right) + 5.668973\left(\frac{W_D}{W}\right)^2\right.\\ &\left.\quad -4.916411\left(\frac{W_D}{W}\right)^3 - 0.145348\left(\frac{W_D}{W}\right)^4\right\}\Big/\\ &\left\{1.0 + 3.924091\left(\frac{W_D}{W}\right) - 6.358805\left(\frac{W_D}{W}\right)^2\right.\\ &\left.\quad +4.018448\left(\frac{W_D}{W}\right)^3 - 1.801705\left(\frac{W_D}{W}\right)^4\right\}\end{aligned} \tag{5-19}$$

   where downcomer chord width (W_D) is assumed 4 inch.

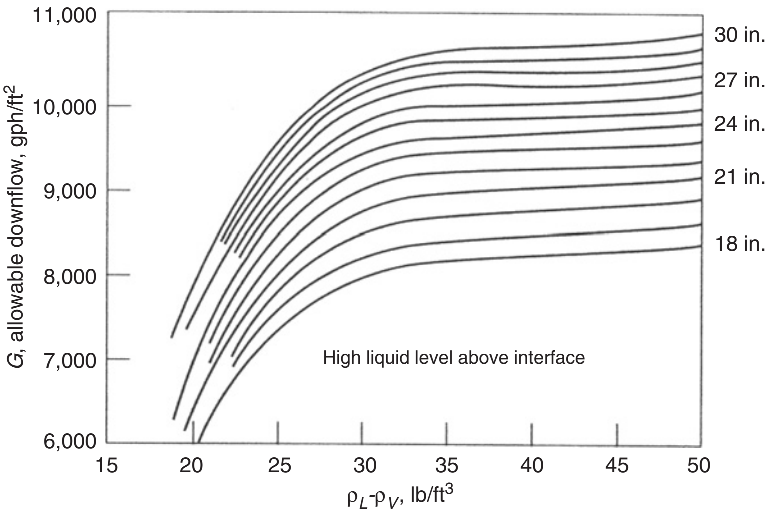


Figure 5-4. Determining the downcomer allowable flow (Monnery and Svrcek, 1994).

In Figure 5-4 the high liquid level above interface is equal to $H_L + H_R$, where the height from light liquid nozzle to baffle (H_R) is assumed 9 inch as a minimum.

9. Calculate the residence time (t_r) of each phase based on the volumes occupied by the light and heavy phases as

$$t_{r,LL} = \frac{H_L A_L}{Q_{LL}}, \text{ min} \tag{5-20}$$

$$t_{r,HL} = \frac{H_H A_H}{Q_{HL}}, \text{ min} \tag{5-21}$$

If $t_{r,LL} \langle t_{s,HL}$ or $t_{r,HL} \langle t_{s,LL}$, increase the diameter and repeat the procedure from step 7 (liquid separation is controlling). Note that $A_H = A$.

10. Calculate H_R based on the required holdup time (t_H) as

$$H_R = \frac{Q_{LL} t_H}{A_L} \tag{5-22}$$

Check this value with that assumed in step 8 to ensure that the assumed value is reasonable. If surge is not specified, calculate the surge height (H_S) based on surge time (t_S):

$$H_S = \frac{t_S\,(Q_{LL} + Q_{HL})}{A} \tag{5-23}$$

where the minimum H_S is 6 inch.

11. Calculate the vessel total height (H_T) as

$$H_T = H_H + H_L + H_R + H_A + H_{BN} + H_D \tag{5-24}$$

where H_A is liquid level above baffle, which is 6 inch (minimum), and HBN is liquid height from above baffle to feed nozzle, ft

$$\text{HBN} = \frac{1}{2}\,(D_N, \text{ft}) + \text{greater of}\,(2\text{ ft or } H_S + 0.5\text{ ft})$$

where the nozzle diameter (D_N) is calculated using the following criterion:

$$D_N \geq \left[\frac{4Q_M}{\pi 60/\sqrt{\rho_M}}\right]^{0.5} \tag{5-25}$$

Q_M and ρ_M are volumetric flow rate and no-slip density of vapor/liquid mixture, respectively.

H_D is disengagement height (ft) = 0.5D or a minimum of

36 inch + ½ (D_N); without mist eliminator

24 inch + ½ (D_N); with mist eliminator

If a mist eliminator pad is used, additional height is added as shown in Figure 5-3.

5.2.5.2 Horizontal Separator

There are different types of horizontal three-phase separators. However, the most common type includes a boot, as shown in Figure 5-5, which is a better design for the system of a small amount of heavy liquid (<15–20%

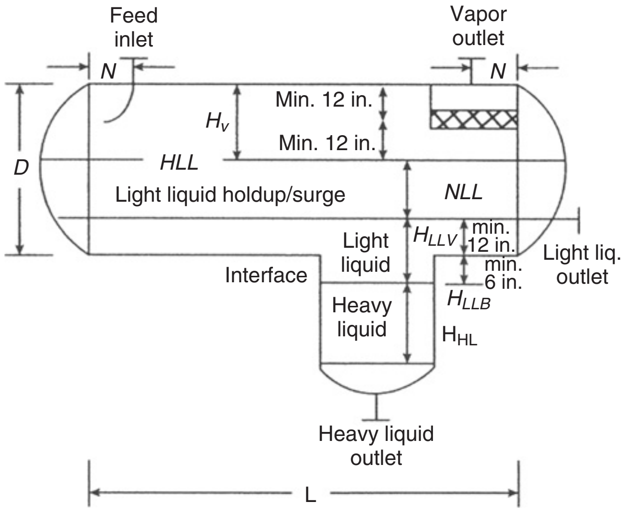


Figure 5-5. Basic design of horizontal three-phase separator with boot (Monnery and Svrcek, 1994)

of total liquid by weight) and a large amount of vapor. This section only presents a design procedure for this type of separator. Readers are referred to the original paper of Monnery and Svrcek (1994) for a detailed design of other types of horizontal three-phase separators.

The horizontal design procedure incorporates optimizing the diameter and length by minimizing the approximate weight of the shell and heads. A stepwise procedure for designing horizontal three-phase separators with a boot is as follows (Monnery and Svrcek, 1994).

1. Calculate the vapor volumetric flow rate, Q_V, using Equation (5-11).
2. Calculate the light and heavy liquid volumetric flow rates, Q_{LL} and Q_{HL}, per Equations (5-13) and (5-14).
3. Calculate the vertical terminal velocity, V_t, using Equation (5-6) and set $V_V = 0.75\ V_t$.
4. Select holdup and surge times from Table 5-1 and calculate the holdup and surge volumes, V_H and V_S, from the following equations (unless surge is otherwise specified, such as slug volume):

$$V_H = t_H \times Q_L \tag{5-26}$$

$$V_S = t_S \times Q_L \tag{5-27}$$

Table 5-2
L/D Ratio Guidelines

Vessel operating pressure (psig)	L/D
$0<P\leq 250$	1.5–3.0
$250<P\leq 500$	3.0–4.0
$P>500$	4.0–6.0

5. Obtain L/D from Table 5-2 and initially calculate the diameter according to

$$D=\left[\frac{4\,(V_H+V_S)}{0.6\pi\left(\frac{L}{D}\right)}\right]^{\frac{1}{3}},\ \text{ft} \tag{5-28}$$

   Then calculate the total cross-sectional area, $A=\frac{\pi D^2}{4}$.
6. Set the vapor space height, H_V, to the larger of 0.2D or 2 ft (1 ft if there is no mist eliminator). Using $\frac{H_V}{D}$ in Equation (5-19) (with replacing $\frac{W_D}{W}$) obtain $\frac{A_V}{A}$ and calculate A_V.
7. Set the light liquid heights in the vessel and boot, H_{LLV} and H_{LLB}.
8. Calculate the cross-sectional area of the light liquid above the bottom of the vessel, A_{LLV}, using $\frac{H_{LLV}}{D}$ (instead of $\frac{W_D}{W}$) in Equation (5-19).
9. Calculate the minimum length to accommodate the liquid holdup/surge:

$$L_1=\frac{V_H+V_S}{A-A_V-A_{LLV}} \tag{5-29}$$

10. Calculate the liquid dropout time, ϕ, using the following equation:

$$\phi=\frac{H_V}{V_V},\ \text{sec} \tag{5-30}$$

11. Calculate the actual vapor velocity, V_{AV}, as

$$V_{AV}=\frac{Q_V}{A_V},\ \text{ft/sec} \tag{5-31}$$

12. Calculate the minimum length required for vapor/liquid separation, L_2, as

$$L_2 = V_{AV} \times \phi, \text{ft} \tag{5-32}$$

13. If $L_1 < L_2$, then set $L_1 = L_2$ (vapor/liquid separation controls). This simply results in some extra holdup and residence time. If $L_1 \ll L_2$, increase H_V and recalculate A_V and then repeat from step 9. If $L_1 > L_2$, the design is acceptable for vapor/liquid separation. If $L_1 \gg L_2$, liquid holdup controls. L_1 can only be reduced and L_2 increased if H_V is reduced. H_V may only be reduced if it is greater than the minimum specified in step 6. With reduced H_V, recalculate A_V and repeat from step 9.
14. Calculate the settling velocity of the heavy liquid out of the light liquid phase, V_{HL}, as

$$V_{HL} = \frac{K_S\,(\rho_H - \rho_L)}{\mu_L} \tag{5-33}$$

    where K_S is obtained from Equation (5-10).
15. Calculate the settling time of the heavy liquid out of the light liquid phase as

$$t_{s,HL} = \frac{12(H_{LLB} + D - H_V)}{V_{HL}} \tag{5-34}$$

16. Calculate the residence time of the light liquid as

$$t_{r,LL} = \frac{(A - A_V)\,L}{Q_{LL}} \tag{5-35}$$

    This volume of light liquid ignores the light liquid volume in the boot.
17. If $t_{r,LL} < t_{s,HL}$, then increase the vessel length (liquid separation controls) as

$$L = \frac{t_{s,HL} \times Q_{LL}}{A - A_V} \tag{5-36}$$

18. Calculate *L/D*. If $L/D \ll 1.5$, then decrease D (unless it is already at a minimum) and if $L/D \gg 6.0$, then increase D; repeat from step 5.

Table 5-3
Wall Thickness, Surface Area, and Approximate Vessel Weight[a]

Component	Wall thickness (inch)	Surface area (ft^2)
Shell	$\frac{PD}{2SE - 1.2P} + \sigma_C$	πDL
2:1 elliptical heads	$\frac{PD}{2SE - 0.2P} + \sigma_C$	$1.09D^2$
Hemispherical heads	$\frac{PD}{4SE - 0.4P} + \sigma_C$	$1.571D^2$
Dished heads	$\frac{0.885PD}{SE - 0.1P} + \sigma_C$	$0.842D^2$
Approximate vessel weight	$W = \left(\frac{490 lb}{ft^3}\right)\left(\frac{\sigma}{12}\right)(A_{Shell} + 2A_{Head})$	

[a] The design pressure (P, psig) is typically either the operating pressure with 15 to 30 psi added to it or the operating pressure plus 10%, whichever is greater. For the allowable stress, S, refer to ASME Pressure Vessel Code (1986). The joint efficiency, E, ranges from 0.6 to 1.0. The corrosion allowance (σ_C) typically ranges from 1/16 to 1/8 inch.

19. Calculate the wall thickness and surface area of the shell and heads, and approximate vessel weight according to Table 5-3.
20. Increase or decrease the vessel diameter by 6-inch increments and repeat the calculations until *L/D* ranges from 1.5 to 6.0.
21. With the optimum vessel size (minimum weight), calculate the normal and high liquid levels (H_{NLL} and H_{HLL}) as

$$H_{HLL} = D - H_V \tag{5-37}$$

$$A_{NLL} = A_{LLV} + \frac{V_H}{L} \tag{5-38}$$

Determine H_{NLL} using Equation (5-19) from $\frac{A_{NLL}}{A}$ (replacing W_D with H_{NLL} and A_{NLL} with A_D).

22. Design the heavy liquid boot. Set the height of the heavy liquid, H_{HL}; calculate the rising velocity of the light liquid out of the heavy liquid phase (V_{LL}) as

$$V_{LL} = \frac{K_s(\rho_H - \rho_L)}{\mu_H} \tag{5-39}$$

Set boot velocity (V_B) to 0.75 V_{LL} and calculate the heavy liquid boot diameter, D_B, as

$$D_B = \left(\frac{4 \times 12 Q_{HL}}{\pi V_B} \right)^{0.5}, \text{ft} \tag{5-40}$$

Then calculate the settling time of the light liquid ($t_{s,LL}$) out of the heavy liquid phase using Equation (5-16).

23. Calculate the residence time of the heavy liquid ($t_{r,HL}$) in the boot as

$$\left(t_{r,HL}\right)_B = \frac{\left(\pi D_B^2\right) H_{HL}}{4 Q_{HL}} \tag{5-41}$$

If $\left(t_{r,HL}\right)_B < t_{s,LL}$, then increase the boot diameter.

5.2.6 Practical Separator Design

The biggest development in recent years is the widespread recognition that the actual performance of a separator may fall far short of the theoretical performance due to the actual flow patterns within the vessel being far from the ideal (although strictly this is a "rediscovery"). It has, however, been helped by two visualization techniques — computational fluid dynamics (CFD) and physical modeling, which vividly show what can go wrong and how to correct it.

The most important areas to ensure a separator performs to design are as follow.

- Correct inlet nozzle sizing and a good inlet device (momentum breaker). The two main sorts are the "vane" inlet (schoepentoeter, or the many proprietary versions of this) and the cyclonic inlet.
- Primary fluid distribution–distribution plates to translate the reduced but still high velocities from the inlet device into quiescent flows in a liquid–liquid separator body, or distribution plates either side of a vane pack (downstream is best as upstream ones shatter droplets unnecessarily) or other gas demister.
- Intermediate fluid distribution when necessary (e.g., to align flow before entry to a plate pack or where a separator has an unusual shape or aspect ratio).
- Exit devices: vortex breakers and antiliquid–pickup details.

The methods of verifying performance “before” and “after” the introduction of good practice such as these internals are CFD and physical modeling, which uses dimensionless number similarity to model at a reduced scale, with air–water, air–oil–water, or gas–oil–water. The more advanced centers have pressure flow loops that can run on oil and gas at pressure, but have the drawback that one cannot view what is happening inside the vessel in the way a perspective model allows.

5.2.7 Operating Problems

The following problems are occasionally occurred in the operation of separators (Arnold and Stewart, 1998).

5.2.7.1 Foamy Crudes

The major cause of foam in crude oil is the appearance of impurities, other than water, that are impractical to remove before the stream reaches the separator. Foam presents no problem within a separator if the internal design assures adequate time or sufficient coalescing surface for the foam to “break.” Foaming in a separating vessel is a threefold problem:

1. Mechanical control of liquid level is aggravated because any control device must deal with three liquid phases, an emulsion is the third phase, instead of two-phases.
2. Foam has a large volume-to-weight ratio. Therefore, it can occupy much of the vessel space that would otherwise be available in the liquid-collecting or gravity-settling sections.
3. In an uncontrolled foam bank, it becomes impossible to remove separated gas or degassed oil from the vessel without entraining some of the foamy material in either the liquid or the gas outlets. Essentially as the foam is dispersed, it creates very small liquid droplets, which carry over.

It should be noted that the amount of foam is dependent on the pressure drop to which the inlet liquid is subjected, as well as the characteristics of the liquid at separator conditions. In some cases, the effect of temperature may be significant. Foam depressants will often be effective in increasing the capacity of a given separator. Foam can be reduced by (1) using a defoaming pack, (2) using defoaming chemicals, and (3) utilizing heat to break it down.

5.2.7.2 Paraffin

Coalescing plates in the liquid section and mesh pad mist extractors in the gas section are particularly prone to clogging by accumulations of paraffin waxes. Manways, handholes, and nozzles should be provided to allow steam, solvent, or other types of cleaning of the separator internals. Also, the bulk temperature of the liquid should always be kept above the cloud point of the crude oil to prevent parraffin wax formation in the separators.

5.2.7.3 Sand

Sand can be very troublesome in separators by causing cutout of valve trim, plugging of separator internals, and accumulation in the bottom of the separator, thus leading to level control problems. Traditionally, sand has only been removed once it has collected in the main production separators. However, removal of sand upstream of these separators reduces sand problems to a minimum, giving substantial operational benefits. To meet these needs, the Mozley Wellspin desander has been developed to remove sand effectively in simple, compact systems based on solid/liquid hydrocyclones, which remove the sand before it enters the separator (NATCO, 2002b). It should be noted that sand problems may be solved by using a filter or desanding cyclone before the separator; however, filters will quickly block in sandy service and are not often used.

5.2.7.4 Liquid Carryover and Gas Blowby

Liquid carryover occurs when free liquid escapes with the gas phase and can indicate high liquid level, damage to vessel internals, foam, improper design, plugged liquid outlets, or a flow rate that exceeds the design rate of the vessel. Gas blowby occurs when free gas escapes with the liquid phase and can be an indication of low liquid level, vortexing, or level control failure.

5.2.7.5 Emulsions

Emulsions can be particularly troublesome in the operation of three-phase separators. Over a period of time an accumulation of emulsified materials and/or other impurities usually will form at the interface of the water and oil phases. In addition to adverse effects on the liquid level control, this accumulation will also decrease the effective oil or water retention time in the separator, with a resultant decrease in water–oil separation efficiency. The addition of chemicals and/or heat often minimizes this difficulty. Frequently, it is possible to appreciably lower the settling time necessary for oil–water separation by either the application of heat in the liquid section of the separator (heat can be added through recycling) or the addition of deemulsifying chemicals.

5.3 MULTISTAGE SEPARATION

To achieve good separation between gas and liquid phases and maximizing hydrocarbon liquid recovery, it is necessary to use several separation stages at decreasing pressures in which the well stream is passed through two or more separators arranged in series. The operating pressures are sequentially reduced, hence the highest pressure is found at the first separator and the lowest pressure at the final separator. In practice, the number of stages normally ranges between two and four, which depends on the gas/oil ratio (GOR) and the well stream pressure, where two-stage separation is usually used for low GOR and low well stream pressure, three-stage separation is used for medium to high GOR and intermediate inlet pressure, and four-stage separation is used for high GOR and a high pressure well stream. Note that three-stage separation usually represents the economic optimum, where it allows 2 to 12% higher liquid recovery in comparison with two-stage separation and, in some cases, recoveries up to 25% higher (Rojey *et al.*, 1997). To recover the gases fractions produced in the separators operating at medium pressure and low pressure, it is necessary to recompress them to the pressure of the high-pressure separator. However, for an associated gas, recompression is sometimes considered too costly; hence the gas produced from the low-pressure separator may be flared.

It should be noted that the main objective of stage separation is to provide maximum stabilization to the resultant phases (gas and liquid) leaving the final separator, which means that the considerable amounts of gas or liquid will not evolve from the final liquid and gas phases, respectively. The quantities of gas and liquid recovered at a given pressure are determined by equilibrium flash calculations using an appropriate equation of state (EOS). This helps optimize the value of pressure that is set for each separator. The pressures are often staged so that the ratio of the pressures in each stage is constant. Therefore, if the pressure in the first separator (which is normally fixed by specification or economics) and the pressure in the final separator (which will be near the atmospheric pressure) are known, the pressure in each stage can be determined.

5.4 CENTRIFUGAL SEPARATORS

In centrifugal or cyclone separators, centrifugal forces act on droplet at forces several times greater than gravity as it enters a cylindrical separator. This centrifugal force can range from 5 times gravity in large, low-velocity units to 2000 times gravity in small, high-pressure units. Generally, centrifugal separators are used for removing droplets greater than 100 μm in diameter, and a properly sized centrifugal separator can have a reasonable removal efficiency of droplet sizes as low as 10 μm.

Centrifugal separators are also extremely useful for gas streams with high particulate loading (Talavera, 1990). Such equipment has already been studied, however, a simple; compact, and lightweight gas–liquid cylindrical cyclone (GLCC) separator has been developed by Tulsa University that requires little maintenance and is easy to install and operate. The compact dimensions, smaller footprint, and lower weight of the GLCC have a potential for cost savings to the industry, especially in offshore applications. Also, the GLCC reduces the inventory of hydrocarbons significantly, which is critical to environmental and safety considerations. The GLCC separator, used mainly for bulk gas/liquid separation, can be designed for various levels of expected performance. Typical performance levels from the GLCC separator are 0.5–2.0 gallons of liquid per MMscf in the gas outlet and 0–5% gas in the liquid outlet (NATCO, 2002a). More information on the design, control system studies, experimental investigations, and field applications of GLCC separators is discussed in greater detail by Gomez *et al.* (2000), Mohan and Shoham (2003), and Wang *et al.* (2003).

5.5 TWISTER SUPERSONIC SEPARATOR

The Twister supersonic separator is a unique combination of known physical processes, combining expansion, cyclonic gas/liquid separation, and recompression process steps in a compact, tubular device to condense and separate water and heavy hydrocarbons from natural gas. Condensation and separation at supersonic velocity are key to achieving step-change reductions in both capital and operating costs. The residence time inside the Twister supersonic separator is only milliseconds, allowing hydrates no time to form and avoiding the requirement for hydrate inhibition chemicals. Elimination of the associated chemical regeneration systems avoids harmful benzene, toluene, and xylene emissions to the environment or the expense of chemical recovery systems. The simplicity and reliability of a static device, with no rotating parts, which operates without chemicals, ensure a simple facility with a high availability suitable for unmanned operation in harsh and/or offshore environments. In addition, the compact and low weight Twister system design enables debottlenecking of existing space and weight-constrained platforms. The first commercial application at the Shell-operated B11 platform offshore Malaysia proves the technical and commercial maturity of a Twister supersonic separator, which is a major milestone toward the industry acceptance of this innovative technology (Brouwer *et al.*, 2004). However, the feasibility and benefits of this technology must be studied for a specific gas development. More information about the system design of the Twister supersonic separator is discussed in greater detail by Okimoto and Brouwer (2002) and Brouwer and Epsom (2003).

5.6 SLUG CATCHERS

Slug catchers are used at the terminus of offshore pipelines to catch large slugs of liquid in pipelines, to hold these slugs temporarily, and then to allow them to follow into downstream equipment and facilities at a rate at which the liquid can be handled properly. Slug catchers may be either a vessel or constructed of pipes. Pipe-type slug catchers are frequently less expensive than vessel type slug catchers of the same capacity due to thinner wall requirements of smaller pipe diameter. The manifold nature of multiple pipe-type slug catchers also makes possible the later addition of additional capacity by laying more parallel pipes. A schematic of a

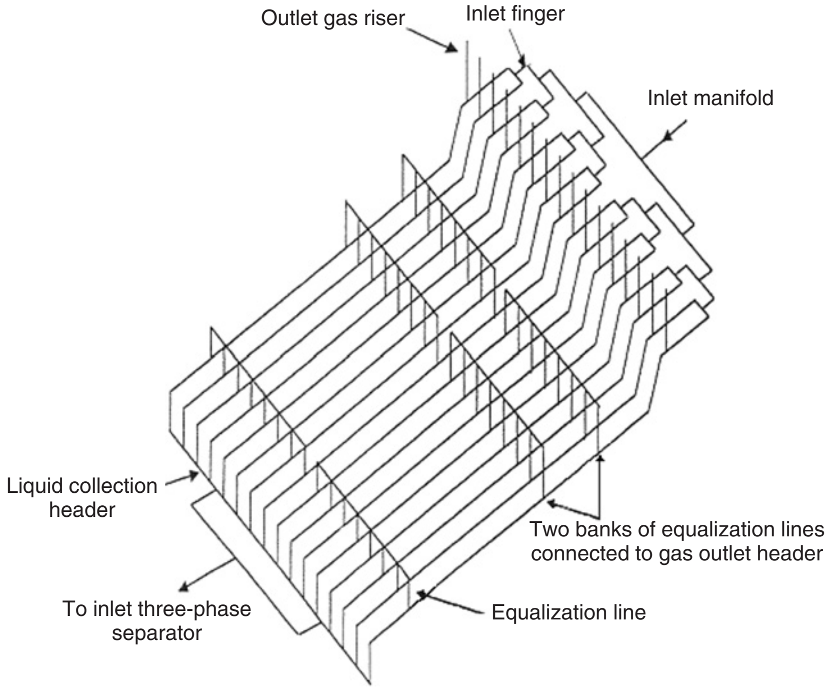


Figure 5-6. Three-dimensional rendering of finger-type slug catcher.

pipe-type slug catcher appears in Figure 5-6. The general configuration consists of the following parts.

- Fingers with dual slope and three distinct sections: gas/liquid separation, intermediate, and storage sections.
- Gas risers connected to each finger at the transition zone between separation and intermediate sections.
- Gas equalization lines located on each finger. These lines are located within the slug storage section.
- Liquid header collecting liquid from each finger. This header will not be sloped and is configured perpendicular to the fingers.

Note that it has been assumed that all liquids (condensate and water) are collected and sent to an inlet three-phase separator, although it is possible to separate condensate and water at the fingers directly. When doing condensate/water separation at the slug catcher itself, we have to allow separately for the maximum condensate slug and the maximum water slug in order to ensure continuous level control.

Separation of gas and liquid phases is achieved in the first section of the fingers. The length of this section promotes a stratified flow pattern and permits primary separation to occur. Ideally, liquid droplets, 600 μm and below, are removed from the gas disengaged into the gas risers, which are located at the end of this section. The length of the intermediate section is minimal such that there is no liquid level beneath the gas riser when the slug catcher is full, i.e., storage section completely full. This section comprises a change in elevation between the gas risers and the storage section that allows a clear distinction between liquid and gas phases. The length of the storage section ensures that the maximum slug volume can be retained without liquid carryover in the gas outlet. During normal operations, the normal liquid level is kept at around the top of the riser from each finger into the main liquid collection header, which is equivalent to approximately a 5-min operation of the condensate stabilization units at maximum capacity.

As the finger type of slug catcher is defined as a piping configuration rather than a pressure vessel, it is not constrained to the same requirements as a normal vessel; however, due to its size, it will contain the majority of high-pressure hydrocarbon gas on the site. It is recommended that the slug catcher be depressurized automatically (for preventation of fire) as quickly as possible without imposing unusually high flow rates on the flare system.

Slug catcher design is dependent on several factors, of which the most important are pigging operation and changes in flow rates (Burke and Kashou, 1996). Pigging can also be used as a means of limiting the required slug catcher size, where by pigging at frequent intervals, liquid inventory buildup in a pipeline can be reduced and the maximum slug size can be limited (GPSA, 1998). However, a slug catcher size should be chosen on the basis of balancing the cost of frequent pigging operations and the capital reduction of smaller slug catchers (Xiao and Shoup, 1998). Pigging requirements and ramp-up periods are determined by transient analysis once the steady-state liquid holdup of the pipeline is understood. Generally, ramp-up rates are determined over the range of operating flow rates. Turndown flow rates are normally predetermined by the client with an overall processing philosophy. Detail information about the complete design of the finger-type slug catchers and other types of slug-catching configurations are discussed regularly with specific vendors.

5.7 HIGH-EFFICIENCY LIQUID–GAS COALESCERS

Aerosols in gas streams can often be less than 5 μm in size and require the use of special separation equipment. High-efficiency liquid–gas coalescers have been applied effectively for the removal of fine aerosols in gas production, processing, and transmission. Coalescers are typically constructed as cartridges that use pleated glass fiber media supported by a metal core. The coalescer cartridges are then placed in a housing that controls the inlet/outlet gas velocities to ensure good separation and prevent any reentrainment of liquids. Coalescer media contain a much finer pore structure and larger surface area as compared to traditional separators that often use mesh pads or vane pack internals.

5.7.1 Aerosols

Aerosols are formed by three mechanisms: condensation, atomization, and entrainment. The relative sizes produced by these formation mechanisms are given in Figure 5-7.

Aerosols formed by condensation of a vapor into a liquid are the smallest and most difficult to remove contaminants, having a size distribution in the range of 0.2 to 5 μm. Atomization creates aerosol drops by breaking up larger liquid drops through mechanical shear, such as passing through a constriction in a valve under a high velocity. Atomization forms

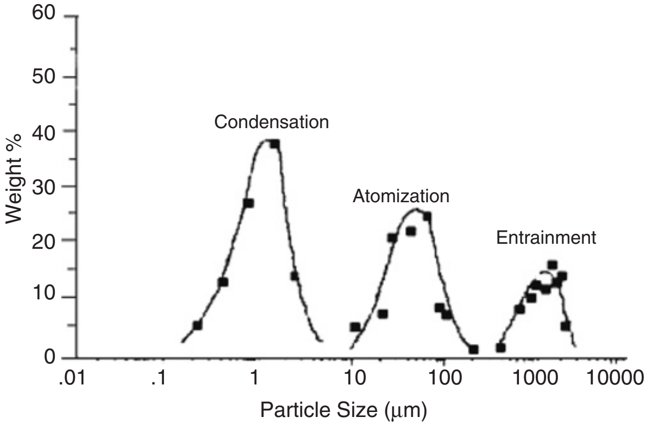


Figure 5-7. Aerosol types (Brown *et al.*, 1994).

aerosols in the size range of 10 to 200 μm. Entrainment involves the movement of liquid slugs along pipelines, where the liquid drop sizes are very large, from 500 to 5000 μm. All three types of aerosol liquids are commonly found in gas systems. High-efficiency liquid–gas coalescers can effectively remove the fine aerosols created by the condensation mechanism.

5.7.2 Coalescer Construction/Operation Principles

High-efficiency liquid–gas coalescers are generally constructed from glass fibers, as this material allows for a fine porous structure with fiber diameters of a few micrometers. The small pore size is needed to achieve greater capture and separation of these small aerosols. The primary rationale for the use of high-efficiency coalescers is that significant aerosol contaminant exists in plants that are in the submicrometer and low micrometer size range (Brown *et al.*, 1994).

This type of liquid–gas coalescer can operate at significantly lower flow rates than the initial design flow rate and therefore can tolerate a high turndown ratio. This is due to the separation mechanisms for coalescing, which are primarily diffusion and direct interception, unlike vane separators and mesh pads that rely heavily on inertial separation principles. This allows the high-efficiency liquid/gas coalescer systems a greater degree of flexibility and they can operate at peak performance even for high turndown ratios (reduced flow rates) that can occur during commonly encountered partial plant shutdowns and upset conditions.

Use of a surface treatment (Miller *et al.*, 1988) on high-performance vertical liquid/gas coalescer cartridge systems has been proven to enhance performance significantly by allowing higher flow rates or smaller housing diameters compared to untreated coalescers. The surface treatment alters the properties of the coalescer medium so that it will not wet out with oil- or water-based fluids. The treatment has also been found to extend the service of the coalescer by reducing fouling and also to lower the saturated pressure drop. A Pall vertical high-efficiency liquid/gas coalescer system is depicted in Figure 5-8.

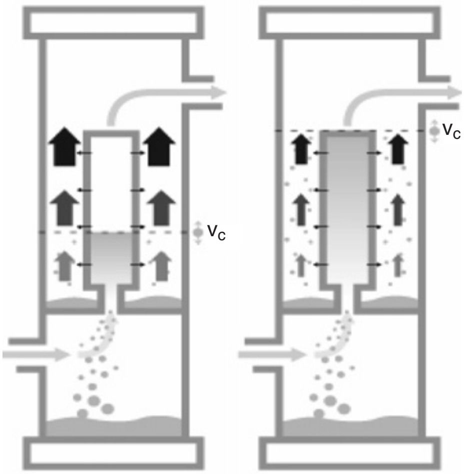
The inlet gas with liquid aerosol contamination first enters at the bottom of the housing into a first-stage knockout section. Here any slugs or larger size droplets (approximately >300 μm) are removed by gravitational settling. The gas then travels upward through a tube sheet and flows radially from the inside of the cartridges through the coalescer

Figure 5-8. High-efficiency vertical liquid–gas coalescer system (Schlotthauer and Hashemi, 1991; Wines, 2004).

medium to the annulus. The inlet aerosol distribution is in the size range of 0.1–300 μm and, after passing through the coalescer medium, is transformed to enlarged coalesced droplets in the size range of 0.5–2.2 mm. The advantage of flowing from the inside to the outside of the coalescer cartridge is that the gas velocity can be adjusted more easily in the annulus by selecting the optimum housing diameter to prevent reentrainment of coalesced droplets.

As the gas leaves the coalescer cartridge and travels upward in the annulus it contributes to the total flow, thereby increasing the annular velocity. The annular velocity is modeled as a linear function with vertical distance, and the annular velocity is zero at the bottom of the cartridge and increases to a maximum value at the top of the cartridge.

Once the coalesced droplets are formed, they immediately drain vertically downward in the coalescer medium pack. The surface treatment enhances this drainage greatly, and as a direct consequence of the treatment, the coalesced droplets are shielded from the upward gas flow in the annulus in the upper two-thirds section of the coalescer cartridge. The coalesced droplets are first exposed to the annular gas flow when they appear



Surface Treated Coalescer

Untreated Coalescer

Figure 5-9. Effect of surface treatment on coalescer drainage (Wines, 2004).

on the external face of the coalescer medium pack at the bottom third of the coalescer cartridge (see Figure 5-9). Once the coalesced droplets are released to the annular space they are subject to the force of the upward-flowing gas. The trajectory of the coalesced droplets is modeled on a force balance between gravity settling and the drag force created by the gas flow past the droplets. This analysis leads to the calculation of a critical annular velocity for reentrainment.

Due to the surface treatment, there are minimal coalesced droplets present in the annulus above the drainage point at the bottom third of the coalescer cartridge. For a coalescer cartridge that is not specially surface treated, the coalesced liquids are present throughout the length of the coalescer in the annulus space, and the critical annular velocity for reentrainment is given for the top of the element (see Figure 5-9). For the treated coalescer, it is allowable to have annular velocities greater than the critical value for reentrainment in the portion of the annulus space where there are no liquids present. This allows the maximum annular velocity at the top of the coalescer cartridge to be about three times the critical reentrainment value needed at the vertical position of the lower one-third of the cartridge height where liquids are present.

Therefore, the maximum annular velocity at the top of the coalescer cartridge is found to be about three times greater than the value for an untreated coalescer. The annular area is determined using the maximum allowable annular velocity and is designed to be of sufficient size to prevent reentrainment and as small as possible to minimize the housing diameter.

5.7.3 Modeling the Liquid/Gas Coalescer

The modeling of the liquid/gas coalescer system can be divided into two basic aspects for performance: media velocity and annular velocity. The other consideration to be taken into account is pressure drop. The pressure drop for a given system can be decreased by using more coalescer elements.

5.7.3.1 Media Velocity

The media velocity (V_{med}) is defined as the actual flow rate divided by the coalescer filter area (Wines, 2004):

$$V_{\mathrm{med}} = \frac{Q_{\mathrm{a}}}{N A_{\mathrm{med}}} \tag{5-42}$$

where N is number of coalescers, A_{med} is media area for one coalescer, and Q_{a} is actual system flow rate at system conditions and is obtained from the standard system flow rate (Q_{s}) as

$$Q_{\mathrm{a}} = \frac{Q_{\mathrm{S}}\ (\mathrm{SG})\rho_{\mathrm{Air,stp}}}{\rho_{\mathrm{G}}} \tag{5-43}$$

where SG is gas-specific gravity, $\rho_{\mathrm{Air,stp}}$ is density of air at standard temperature and pressure, and ρ_{G} is density of gas at system conditions.

The media velocity is not the actual velocity through the open pores of the media, but rather an average by convention over the combined pore area and solid matrix area in the spatial plane normal to the flow direction. The maximum media velocity for a coalescer construction is related to a number of factors intrinsic to the particular coalescer design and to the physical properties of the system. Four steps have been identified with the mechanism of the formation and removal of droplets in the coalescer medium: (1) capture, (2) coalescing, (3) release, and (4) drainage.

Formation of the coalesced droplets first involves the capture of the small aerosols onto the fibers of the coalescer medium. The actual coalescing or merging of the fine droplets is believed to take place on the fibers, especially at fiber intersections. The coalesced droplets are then released from the fiber due to the drag force of the gas flow exceeding the adsorption energy. This process is repeated through the depth of the coalescer medium until the coalescing process is completed and the largest possible stable droplet size is achieved. During the coalescing stages, the growing droplets are also draining downward inside the media pack due to the force of gravity.

The surface treatment allows the release and drainage process to proceed at a faster rate, which in turn frees up more coalescing sites on the fibers and allows the coalescer to process higher inlet liquid aerosol concentrations than the untreated coalescer medium.

5.7.3.2 Effect of System Conditions on Media Velocity

The ability of the coalescer medium to perform effectively also depends on the system environment. While different coalescer constructions will exhibit quantitative differences, they will follow the same qualitative behavior. The media velocity has been determined to depend on system parameters such as inlet aerosol concentration, aerosol density, gas density, and gas viscosity.

At low aerosol concentrations, the maximum media velocity is constant and is unaffected by aerosol levels. Under these conditions, media are limited by the capture mechanism and are not affected by drainage. At higher levels of aerosol concentration, the coalescer medium becomes limited by drainage and is inversely proportional to the aerosol concentration. The effect of the surface treatment on this process is to enhance the drainage and allow for higher maximum media velocities under the same aerosol loading when limited by drainage.

5.7.3.3 Annular Velocity

The annular velocity (V_{ann}) is defined as the actual flow rate divided by the annulus area (Wines, 2004):

$$V_{\mathrm{ann}} = \frac{Q_{\mathrm{a}}}{A_{\mathrm{ann}}} \tag{5-44}$$

where A_{ann} is cross-sectional annular area defined as the cross-sectional area of the housing without coalescers minus the area of the coalescer end caps:

$$A_{\text{ann}} = \pi R_{\text{h}}^2 - N\pi R_{\text{c}}^2 \tag{5-45}$$

where R_{h} is radius of the housing, R_{c} is radius of coalescer end cap, and N is number of coalescers.

The enlarged droplets leaving the coalescer media pack can be assumed to be as large as possible for the given flow conditions when complete coalescence has occurred. Therefore, the coalesced droplet diameter will be the same for any specific design of the coalescer cartridge as long as complete coalescence has been achieved. If complete coalescence is not achieved, the calculation of the coalesced droplets must take into account the degree of coalescence.

In most industrial applications, the coalesced droplets will range in size from 0.5 to 2.2 mm and will be mostly influenced by the interfacial tension, which is significantly affected by the liquid density, system temperature, and system pressure. As the pressure is increased, the gas density will increase while the liquid density is affected only slightly. The solubility of the gas in the liquid is enhanced with increasing pressure. This leads to a substantial decrease in interfacial tension with increasing pressure and consequently to significantly smaller coalesced droplets at the higher pressures.

Once the coalesced droplet size has been estimated, the next step is to determine the maximum annular velocity that can be sustained without reentrainment. In general, the coalesced droplets will produce Reynolds numbers (Re) outside of the creeping flow regime (<0.1) and Stokes law. Instead, a force balance is used between the liquid droplets settling by gravity and the drag force of the gas flowing upward in the opposite direction.

5.7.3.4 Determination of Minimum Housing Diameter

The housing diameter is determined from the area of the annulus and the area of the coalescer end caps. The maximum annular velocity at the top of the coalescer cartridges is used to determine the annular area required. The value of the maximum annular velocity ($V_{\text{ann,max}}$) at the top of the coalescer cartridges is dependent on the critical annular velocity for reentrainment (V_c) and the vertical location at which the coalesced droplets are present in the free annulus space. This relationship can be described as follows (Wines, 2004):

$$V_{\mathrm{ann,max}} = k_{\mathrm{a}} V_{\mathrm{c}} \tag{5-46}$$

where k_{a} is the annular velocity enhancement factor due to drainage.

For the untreated coalescer medium the coalescer cartridge is completely wetted and coalesced droplets are present in the annulus space up to the top of the annulus where the annular velocity is highest. There is no drainage enhancement and $k_{\mathrm{a}} = 1$. The maximum annular velocity to prevent reentrainment is then equal to the critical value for reentrainment:

$$\text{Untreated coalescer:} \qquad V_{\mathrm{ann,max}} = V_{\mathrm{c}} \tag{5-47}$$

The effect of the surface treatment is to increase the drainage greatly, and the annular velocity at the top of the coalescer cartridge can now be significantly higher than the critical value because there are no coalesced droplets present in the annulus except in the bottom third of the cartridge. The maximum annular velocity is now determined with $k_a = 3.1$ as follows:

$$\text{Surface Treated-Coalescer:} \qquad V_{\mathrm{ann,max}} = 3.1 V_{\mathrm{c}} \tag{5-48}$$

Convincing evidence for the enhanced maximum annular velocity given by Equation (5-46) has been demonstrated by laboratory tests (Williamson *et al.*, 1988; Murphy, 1984) and is presented in Figure 5-9. Visual observations during these tests also confirm that liquids are present on the outside of the coalescer pack only at the bottom third for the surface-treated coalescer and are present throughout the length of the wetted untreated coalescer.

5.7.4 Coalescer Performance/Operational Limits

Generally, the high-efficiency liquid/gas coalescers are used for inlet aerosol concentrations of a few thousand ppmw or less and are placed downstream of other bulk removal separators as the final purification stage. Under these conditions, typical service life for liquid/gas coalescers is 1–2 years. Coalescer systems are usually sized for a clean differential pressure (DP) of 2–5 psi, and when this DP reaches 15 psi, they are replaced with new elements.

Outlet concentrations for these high efficiency liquid/gas coalescers are as low as 0.003 ppmw (Williamson *et al.*, 1988; Murphy, 1984). The experimental apparatus for measuring coalescer efficiency is given in Figure 5-10. This test was developed by Pall Corporation and is known as the liquid aerosol separation efficiency test. At this time, there is no standard test for liquid–gas coalescers that the industry has universally adopted. Some of the important aspects of measuring coalescer performance addressed in this test include the following.

- Difficult aerosol challenge with drops ranging from 0.1 to 1 μm created in an atomizer — representative of condensation aerosols.
- Test operated under pressurized conditions — not vacuum.
- Coalescer evaluated after it is saturated with liquids — representative of field operation.
- Outlet contaminant collected by a full flow sampler that collects any aerosols or liquids flowing along the walls of the pipe.

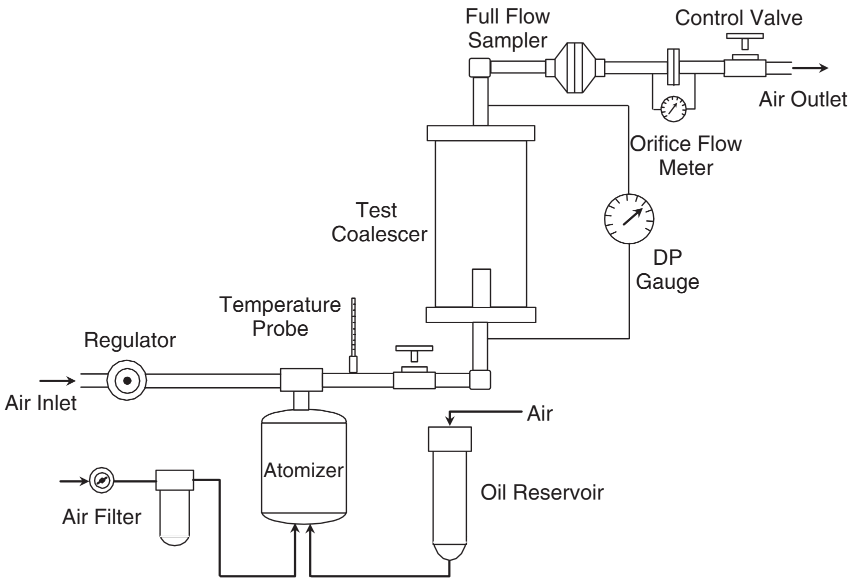


Figure 5-10. Liquid aerosol separation efficiency test.

5.7.5 Liquid/Gas Coalescer Applications

The separation of liquid aerosol contamination with high-performance liquid/gas coalescer cartridge systems has found widespread acceptance in refinery and gas plants in recent years for a number of applications, including protection of compressors, turbo equipment, low NOx burner nozzles, amine and glycol contactors, molecular sieve beds, and well head hydrate inhibition. This has largely been the result of traditional separation approaches, including knockout vessels, centrifugal separators, mesh pads, or vane separators, not meeting the end user's requirements for aerosol reduction. A brief description of some of the main applications is given here.

5.7.5.1 Compressor Protection

Contaminants in the inlet gas can have a severe effect on compressor reliability. At least 20% of all reciprocating compressor failures can be attributed to inlet gas contaminants (Smith *et al*., 1997). Even when the compressors are not having catastrophic failure, a costly maintenance schedule may be followed where the compressor is shut down every 6 months or less for inspection and minor repairs. A preventative system using a high-performance liquid/gas coalescer system, however, has been found to provide protection of the compressor for a 2-year period (Wines and Lorentzen, 1999).

Gas contaminants are made up of solid particulates and liquid aerosols, often including a high percentage in the submicrometer size range. Solids are usually corrosion products (iron oxides or iron sulfides), salts, or silt. Liquids can be hydrocarbons (refinery products, lube oil, condensate), aqueous (water, alcohol, dissolved salts, caustic, acid), or a combination of both. Contaminants affect reciprocating and centrifugal compressors in different ways.

Reciprocating compressors are affected by aerosol contaminant primarily in the piston cylinder intake/exhaust. A buildup of contaminants on the valves can lead to sticking valves or valves with partial bypass resulting in a reduced compression ratio and increased power consumption. The buildup of contaminants inside the cylinder can also damage the piston rings, pistons, and the cylinder wall.

Centrifugal compressors impart a dynamic head to the inlet gas by use of high-speed impellers that are confined in a casing containing stationary diffusers. While centrifugal compressors are less sensitive to inlet gas contaminants, they are still affected adversely. For centrifugal compressors, the primary contaminant-related problem is the buildup of foulants on the rotating blades (also known as salting). This can lead to a partial blockage in the flow path causing increased power consumption and can create an imbalance in the blades leading to serious mechanical failures due to resultant vibrations if left unchecked.

5.7.5.2 Amine/Glycol Contactor Protection

Gas-treating processes that use liquid–gas contactors can be prone to foaming when contaminants build up in the solvent systems. The most common gas-treating solvent systems are alkanol amines used to remove H_2S or CO_2 and triethylene glycol used to remove water vapor. Activated carbon and antifoam chemical additive are often used to alleviate foaming issues. These approaches are not effective for significant contaminant ingression, as the activated carbon becomes saturated past its capacity at a rapid rate. The use of antifoam treats a symptom rather than addressing the root cause by removing the contaminant. Upsets in contaminant levels can often trigger foaming, as it is difficult to gauge the respective antifoam level required. High levels of antifoam additive can also have the opposite effect and lead to foam promotion in some instances. A total fluid management approach of using high-efficiency liquid/gas coalescers on the incoming gas and 10-μm absolute filters on an amine loop was found to be an effective approach to minimizing foaming and amine losses (Pauley *et al.*, 1988).

5.7.5.3 Well Head Hydrate Inhibition

Hydrates, solid water/methane complexes, can form in natural gas under high pressures even at ambient temperatures. Hydrates can wreck havoc on gas processing and transmission systems by plugging off pipelines and process equipment. Methanol and/or glycol injection is often used to inhibit hydrate formation. Removing water aerosols right at the well head is an even more effective way to prevent issues with hydrates forming in the downstream pipeline and will reduce chemical injection costs for methanol and glycol.

5.7.5.4 Molecular Sieve Protection

Molecular sieves (zeolites) are used to dehydrate gas to achieve dew points well below saturation and are effective at removing water vapor. When free water in the form of aerosols is present, the molecular sieves can reach maximum water capacity rapidly, leading to frequent regeneration. Regeneration is accomplished thermally and will have an associated cost for this heating value. Hydrocarbon and solid aerosols can lead to fouling of the molecular sieve bed, thereby reducing the number of active sites to bind to water vapor. The use of high efficiency liquid/gas coalescers is an effective and economic means to protect molecular sieve beds.

5.7.5.5 Low and Ultra Low NOx Burner Protection

Burners are used extensively in the refining and petrochemical industries. Applications include furnaces used for crude oil heating, cracking, reforming, and coking, as well as boilers, gas turbines driers, and incinerators. To meet environmental regulations, advanced burner technology including low and ultralow NOx burner tips is being applied increasingly. These advanced burner designs are more intricate in design and have finer orifices that are more prone to fouling than older designs.

Contamination in the fuel gas can lead to process problems with the furnace operation. The nozzles can become plugged, leading to poor furnace performance and, in extreme cases, damage the convective sections of the furnace. At some plants, furnace maintenance has become a costly task, requiring burner tip replacement or cleaning every few days. Fuel gas can contain contaminants from various sources. Corrosion products can form in the process piping and lead to fouling by inorganic materials. Liquid hydrocarbons can result in coking at the burner tip, leading to fouling by organic materials. Some of the contaminants found in fuel gas include iron sulfides, iron oxides, amine, glycol, water, and hydrocarbon liquids.

High-efficiency liquid–gas coalescers have been applied successfully to protect low and ultralow NOx burners (Wines, 2004). They can be installed either at the point of use directly upstream of a bank of furnaces or at a central location conditioning the feed to multiple banks of burners. The piping should be cleaned thoroughly prior to installing the coalescers and any piping downstream of the coalescers should be either heat traced or insulated to prevent temperature drops that could cause liquids to condense.

5.8 HIGH-EFFICIENCY LIQUID–LIQUID COALESCERS

Liquid–liquid separations may require the use of special equipment when the drop sizes are small, typically in the range of 1 to 50 μm in size. These fluid systems are classified as stable emulsions, and often conventional bulk separators with mist pads or plate-type internals will not be effective. High-efficiency liquid–liquid coalescers have been developed to break these emulsions and provide improved separation.

5.8.1 Emulsions

Emulsions consist of the three components: oil (representing hydrocarbon or organic liquids), water (including any aqueous mixtures), and surfactants. Depending on the ratio of these components, oil-in-water emulsions or water-in-oil emulsions can exist. The structure of the oil-in-water or water-in-oil emulsions is well defined with spherical droplets of the dispersed phase surrounded by a bulk continuous phase and surfactant sheathing the droplets. Surfactants contain both hydrophilic (water loving) and hydrophobic (water fearing) portions in the same molecule. This unique structure allows them to associate at water–oil interfaces and helps them stabilize the droplet shape. A spherical drop shape is formed to minimize the surface area between the oil and the water, which also minimizes the free energy required to make this surface. In order to make an emulsion the system must be subjected to shear or mixing to allow the three components to break up into the droplet structure.

Emulsions are inherently unstable and will spontaneously separate out into two bulk phases. This process requires that the small droplets merge together or coalesce repeatedly until they form increasing drop sizes that eventually merge with a bulk phase until all of the drops are gone. Depending on the nature of the emulsion, the separation can occur in a matter of seconds or months. Many of the same factors that affect emulsion stability also influence coalescer performance.

5.8.1.1 Surfactants

Surfactants can be broadly classified into three groups: cationic, anionic, and nonionic. All surfactants consist of polar or hydrophilic groups joined to nonpolar or hydrophobic hydrocarbon chains. Cationic surfactants contain polar head groups that have a positive charge whereas anionic surfactants have polar groups that have negative charges. Nonionic surfactants have polar groups that are neutral and are typically made up of ethylene oxide groups.

Surfactants can have a wide array of configurations, including a polar head and nonpolar tail structure, branched tails, and random placement of polar groups within a nonpolar hydrocarbon chain. The ability of surfactants to aid in the emulsification process will depend on the ratio of the polar to nonpolar groups, the charge density and size of the polar group, and the volume occupied by the nonpolar groups (branching and length). Surfactants can also be influenced by other species such as cosurfactants (alcohols) or aqueous contaminants (salts).

The sources of surfactants found in industrial processes include intentional additives, surfactants found in nature, and surfactants created inadvertently through reaction processes. Some examples of surfactants classified this way are given.

Additives: corrosion inhibitors, deemulsifiers, scale inhibitors, flocculents
Natural: petroleum naphtha sulfonates, naphthenic acids and mercaptides in crude oil (Hughes, 1997)
Reaction products: oxidation/caustic treatment of hydrocarbons in refineries leading to formation of cresylic acids and phenol homologs (Suarez, 1996)

5.8.2 Coalescer Principles and Materials of Construction

Coalescers are typically manufactured as either pads or cartridge filters that have been designed especially to take small droplets in an emulsion and grow them into large drops that are separated more easily. This process is accelerated over natural coalescing by the fibers present in coalescer media that force the contact of small droplets, thereby promoting the coalescing process. The pore gradient of coalescer medium is constructed so that the inlet medium has fine pore sizes that increase in size with the flow direction (see Figure 5-11).

Coalescers have been primarily constructed with glass fiber media until recently when polymer and fluoropolymer materials were adopted. Glass fiber works adequately for emulsions with interfacial tensions >20 dyne/cm. It is known to disarm and lose efficiency in the presence of surfactants (Hughes, 1997). These coalescers are widely used to dewater jet fuel for the aviation industry.

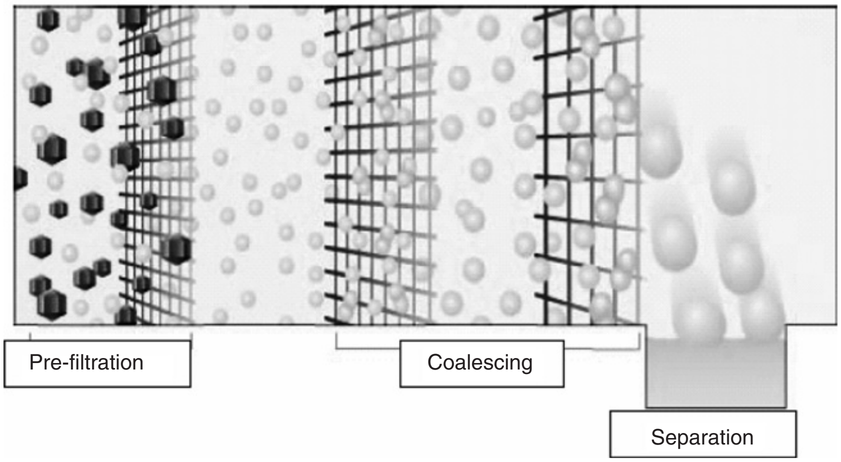


Figure 5-11. Coalescing in media.

High-efficiency liquid–liquid coalescers are the newest generation of coalescers, incorporating the latest in coalescer technology. They are constructed from polymer and fluoropolymer materials that have been optimized to separate the most difficult emulsions with interfacial tensions as low as 0.5 dyne/cm. This coalescer can be used with a broad range of applications. It can process aggressive chemicals and handle demanding operating conditions while providing the highest level of performance.

5.8.3 Coalescer Mechanism of Operation

The liquid–liquid coalescing system operates in three stages: separation of solids/preconditioning, coalescence, and separation of coalesced drops.

5.8.3.1 Separation of Solids/Preconditioning of fluid

Solids can increase the stability of an emulsion and removing solids can make coalescing easier. Generally, this step can be achieved by a separate cartridge filter system or by a regenerable backwash filter system for high levels of solids. In addition, the filtration stage protects the coalescer and increases service life. This step also initiates the coalescence of the hydrocarbon droplets, thereby enhancing the separation capabilities of the system.

5.8.3.2 Coalescence

The next step in the process is primary coalescence. In this stage, pore dimensions begin with a very fine structure and then become more open to allow for void space for the coalescing droplets. In the primary coalescence zone, the inlet dispersion containing fine droplets in the size range of 0.2 to 50 μm is transformed into a suspension of enlarged droplets in the size range of 500 to 5000 μm.

The coalescence mechanism can be described by the following steps.

1. Droplet adsorption to fiber
2. Translation of droplets to fiber intersections by bulk flow
3. Coalescence of two droplets to form one larger droplet
4. Repeated coalescence of small droplets into larger droplets at fiber intersections
5. Release of droplets from fiber intersections due to increased drag on adsorbed droplets caused by bulk flow
6. Repeat of steps 1–5 with progressively larger droplet sizes and more open media porosity

Based on this mechanism, we can predict that a number of factors will influence the coalescence performance. The specific surface properties of the coalescer fibers are critical in influencing the adsorption of droplets as well as the ultimate release after coalescing. There is a balancing act between increasing the attraction or adsorption characteristics of the fibers against the release mechanism, which strong adsorption would inhibit. The necessary condition that droplet–fiber adsorption occurs for coalescing has been supported by a number of sources (Jeater *et al.*, 1980; Basu, 1993).

5.8.3.3 Separation of Coalesced Droplets

Once the droplets have been coalesced they are now as large as possible for the given flow conditions. The separation stage can be achieved in one of two ways.

5.8.3.3.1 Horizontal Configuration The coalescer housing contains a settling zone that relies on the difference in densities between the coalesced droplets and the bulk fluid (see Figure 5-12). This configuration can be

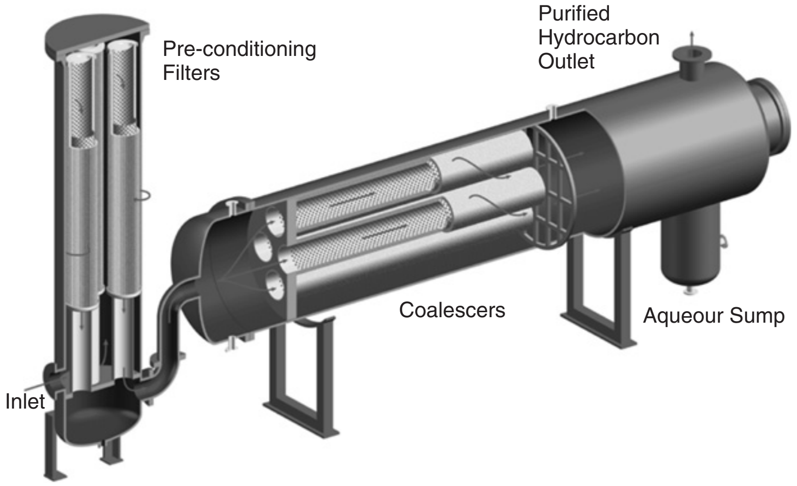


Figure 5-12. Horizontal liquid–liquid coalescer configuration (Katona *et al.*, 2001).

used for both hydrocarbon from water and water from hydrocarbon separation, but the location of the collection sump and outlet nozzle will need to be reversed. For the case of removal of hydrocarbon from water, a collection sump is located at the top of the housing and the purified water leaves at the bottom outlet nozzle. The sump can be drained manually on a periodic basis or equipped with an automatic level control and drain system. Estimation of the coalesced drop size and required settling zone are best determined through pilot scale tests at field conditions.

5.8.3.3.2 Vertical Configuration Once the droplets have been coalesced they are now as large as possible for the given flow conditions, in the range of 0.5 to 2 mm in diameter. The separation stage is achieved using hydrophobic separator cartridges that provide an effective barrier to aqueous coalesced drops, but allow hydrocarbon to pass through them. The separator cartridges can be stacked below the coalescers for the most efficient utilization of the separator medium. This configuration only applies to the separation of water or aqueous contaminants from hydrocarbons (see Figure 5-13).

After leaving the coalescing stage, the large aqueous coalesced drops and hydrocarbon then flow axially in a downward direction and the flow direction is from the outside of the separator to the inside. The large

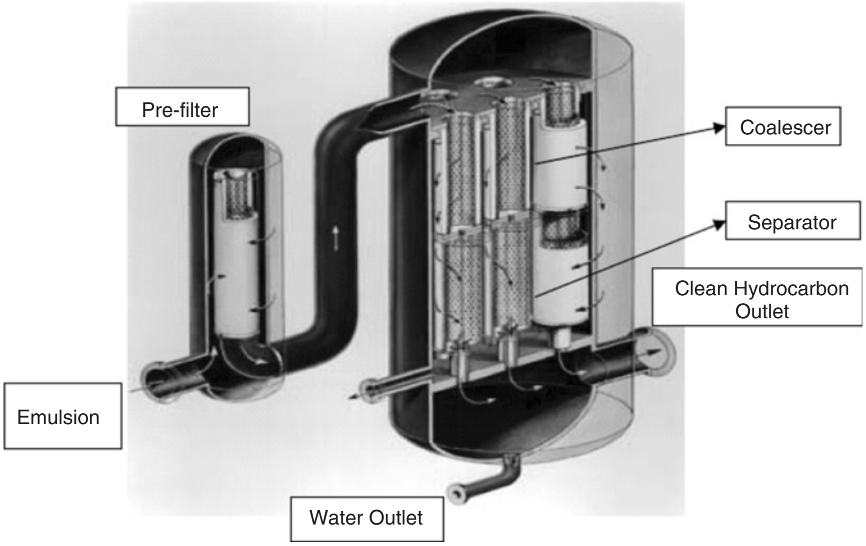


Figure 5-13. Vertical liquid–liquid coalescer configuration (Hampton *et al.*, 2001).

coalesced drops are repelled by the separators and are collected in the bottom sump. The purified hydrocarbon passes through the separators and exits at the bottom of the housing. The aqueous phase in the collection sump can be drained manually on a periodic basis or equipped with an automatic level control and drain system.

5.8.4 Liquid/Liquid Coalescer Performance

Properly designed and sized high-efficiency coalescer systems can process inlet discontinuous phase concentrations up to 10% and reduce them to ppm levels in the outlet for interfacial tensions as low as 0.5 dyne/cm. For water from hydrocarbon separations, coalescer outlets below 15 ppmv per the AquaGlo (registered trademark of Gammon Corp.) method (ASTM D3240) can be achieved, and for hydrocarbons from water, concentrations below 20 ppmw per the oil and grease method have been demonstrated.

The use of polymers and fluoropolymers in the coalescer materials of construction allows for the expanded use of coalescers over earlier conventional types so that they can withstand an array of aggressive chemical applications over a wide range of temperatures from −40°F up to 300°F.

5.8.5 Limitations of Using Coalescers

While liquid–liquid coalescers have many benefits in breaking tough emulsions, there are some limitations to consider. Solids can become problematic at higher concentrations and lead to excessive change out of disposable prefilters. Generally, the solids range that liquid–liquid coalescers can operate economically with disposable filters is <10 ppm. Above this level of solids, further pretreatment will be required, such as backwash cartridge filters, mixed media packed beds, or hydrocyclones for solids removal.

Operational limits of the coalescer for the removal of free liquids must also be understood. If the clarified stream leaving the coalescer is then cooled, condensation of a previously dissolved contaminant can occur, leading to a hazy fluid at the lower temperature. The coalescer will not be able to remove contaminants that are dissolved in solution and, therefore, the location to place the coalescer and any subsequent change in process conditions after the coalescer must be considered carefully.

Coalescers typically will have a service life of 1 to 2 years when protected adequately by prefiltration. Despite the long life, the coalescers will eventually require disposal and replacement; however, this will be minimal given the low frequency of replacement.

For liquid–liquid coalescers constructed from glass fiber medium, the problem of surfactant disarming must also be considered and for low IFT emulsion systems (<20 dyne/cm) they will not operate efficiently for separation. For these conditions, nondisarming fluoropolymer or polymer coalescers should be considered. These types of materials also have wider compatibility for chemical streams and a wider temperature range.

5.8.6 Applications

High-efficiency liquid–liquid coalescers are finding increasing applications in industry where problematic emulsions exist. They are used to protect equipment, to recover valuable streams, and to meet environmental discharge limits. Some examples are given.

5.8.6.1 Pipeline Condensate

In gas production, associated light-end hydrocarbon condensates and water are produced along with the gas. In many instances, chemical additives such as methanol and glycol are used to prevent hydrate formation. These chemicals can increase greatly the tendency for the water–hydrocarbon condensate to form stable emulsions. High-efficiency liquid–liquid coalescers have been applied to separate these difficult emulsions to purify both water and hydrocarbon phases.

5.8.6.2 Produced Water

Discharge limits on produced water are becoming more stringent with current levels in many areas at 29 ppm oil and grease. To treat produced water to achieve this target often requires multiple separation options, including mixed sand beds, dissolved air flotation, and hydrocyclones. Despite these efforts, emulsified oil can still penetrate through to the effluent, and the use of high efficiency liquid–liquid coalescers offers a reliable way to meet environmental limits.

5.8.6.3 Petrochemical Final Products

In refinery operations, steam stripping is often used to remove volatile components (hydrogen sulfide and light hydrocarbon fractions) from petroleum products, including gasoline, diesel, and kerosene. This can lead to water condensation and off specification hazy final products.

5.8.6.4 Caustic Treating

In the production of gasoline, organic sulfur compounds such as mercaptans often need to be removed. One of the more well-known processes for the removal of mercaptans is known as the Merox (registered trademark of UOP) process, which is licensed by UOP. In the reactor, mercaptans are extracted into the caustic and are then converted to disulfide oils by oxidation and catalytic action. Phenols are also extracted into the caustic phase. The reactor effluent flows to a three-phase separator where air, disulfides, and regenerated caustic are separated. The regenerated caustic is then recycled to the reactor.

The reactor effluent typically contains a quantity of carried over caustic that results in hazy gasoline product, high costs of caustic make-up, and corrosion of downstream piping. The hazy gasoline is problematic and has to be blended off or reprocessed, as it would cause the product to be off-specification in terms of sodium. Installing a high-efficiency liquid–liquid coalescer system was found to be an effective solution to recover the caustic carryover (Katona *et al.,* 2001).

5.8.6.5 Recover Liquid Catalysts

Liquid or homogeneous catalysts are used in a number of processes where the catalyst and the reactants are insoluble and form emulsions. Recovery of the catalyst is important to reduce the need for make up catalyst and to protect downstream equipment from fouling by the catalyst.

REFERENCES

API Spec 12J, “Specification for Oil and Gas Separators,” 7th Edition. American Petroleum Institute, Washington, DC (Oct. 1989).

ASTM D3240,”Standard Test Method for Undissolved Water in Aviation Turbine Fuels,” ASTM International.

Arnold, K., and Stewart, M., “Surface Production Operations, Vol. 1: Design of Oil-Handling Systems and Facilities,” 2nd Ed. Gulf Professional Publishing, Houston, TX (1998).

ASME Pressure Vessel Code, Section VIII, Division 1, Table UCS-23, PP.270-271, ASME, New York (1986).

Basu, S., A study on the effect of wetting on the mechanism of coalescence. *J. Colloid Interface Sci.* **159** (1993).

Brouwer, J.M., Bakker, G., Verschoof, H-J., and Epsom, H.D., “Supersonic Gas Conditioning: First Commercial Offshore Experience,” Paper presented at 83rd Annual GPA Convention, New Orleans, LA (March 14–17, 2004).

Brouwer, J.M., and Epsom, H.D., “Twister Supersonic Gas Conditioning for Unmanned Platforms and Subsea Gas Processing,” SPE 83977. Paper presented at Offshore Europe Conference, Aberdeen, UK (Sept. 2–5, 2003).

Brown, R.L., Malbrel, C., and Wines, T.H., “Recent Developments in Liquid/Gas Separation Technology.” Paper presented at the 44th Laurance Reid Gas Conditioning Conference, Norman, OK (Feb. 28, 1994).

Burke, N.E., and Kashou, S.F., Slug-sizing/slug-volume prediction: State of the art review and simulation. *SPE Prod. Facilities J.* **11,** 166–172 (1996).

Campbell, J.M., “Gas Conditioning and Processing,” 7th Ed. Campbell Petroleum Series, Norman, OK (1992).

Gomez, L.E., Mohan, R.S., Shoham, O., and Kouba, G.E., Enhanced mechanistic model and field application design of gas-liquid cylindrical cyclone separators. *SPE J.* **5**(2), 190–198 (2000).

GPSA Engineering Data Book, Vol. 1, Section 7, 11th Ed. Gas Processors Suppliers Association, Tulsa, OK (1998).

Hampton, P., Darde, T., James, R., and Wines, T.H., “Liquid-Liquid Separation Technology Improves IFPEXOL Process Economics,” Oil & Gas Journal, 99, 16, 54–57 (2001).

Hughes, V. B., “Aviation Fuel Handling: New Mechanism Insight into the Effect of Surfactants on Water Coalescer Performance.” Paper presented at the 2nd International Filtration Conference, San Antonio (April 1997).

Jeater, P., Rushton, E., and Davies, G.A., “Coalescence in Fibre Beds.” Filtration and Separation (March/April 1980).

Katona, Q., Darde, T., and Wines, T.H., “Improve Haze Removal for FCC Gasoline.” Hydrocarbon Processing, 103–108 (August 2001).

Kumar, S., “Gas Production Engineering.” Gulf Publishing Company, Houston, TX (1987).

Manning, F.S., Thompson, R.E., “Oil Field Processing, Vol. 2: Crude Oil.” Pennwell Publishing Company, Tulsa, OK (1995).

Miller, J.D., Koslow, R.R., and Williamson, K.W., U.S. Patent 4, 676, 807 (June 1987); id. U. S. Patent 4, 759, 782 (July 1988).

Mohan, R.S., and Shoham, O., “Design and Development of Gas-liquid Cylindrical Cyclone: Compact Separators for Three-phase Flow.” Final Technical Report submitted to the U.S. Department of Energy (June 25, 2003).

Monnery, W.D., and Svrcek, W.Y., “Successfully Specify Three-Phase Separators.” Chemical Engineering Progress, 29–40 (Sept. 1994).

Murphy, W.L., “Practical In-Service Simulation Tests for Rating of High Efficiency Aerosol Coalescing Performance.” PEDD-FSR-101a, Pall Corporation Equipment Development (November 1984).

NATCO, “Gas-Liquid Cylindrical Cyclone (GLCC) for Compact Gas-Liquid Separations.” NATCO Group Products Manual, Issue 1, PP. 1243-A1/A2 (July 2002a).

NATCO, “Mozley Wellspin Extensively Developed with Maximum Resistance to Wear and Robustness in Operation.” NATCO Group Products Manual, Issue 1, PP. 305 A1-A4 (Aug. 2002b).

Okimoto, F.T., and Brouwer, J.M., Supersonic gas conditioning. *World Oil.* **223**(8), (2002).

Pauley, C. R., Hashemi, R., and Caothien, S., "Analysis of Foaming Mechanisms in Amine Plants." Paper presented at the AIChE Summer Meeting, Denver, Colorado (August 1988).

Rojey, A., Jaffret, C., Cornot-Gandolphe, S., Durand, B., Jullian, S., and Valais, M., "Natural Gas Production, Processing, Transport." Editions Technip, Paris, France (1997).

Schlotthauer, M., and Hashemi, R., "Gas Conditioning: A Key To Success In Turbine Combustion Systems Using Landfill Gas Fuels." Paper presented at the 14th Annual Landfill Gas Symposium, GRCDA/SWANA, San Diego, California (March 27, 1991).

Smith, T.J., Schultheis, S.M., and Hanifan, M., "Millennium Petrochemicals Inc. Morris Plant Improves Compressor Reliability and Onstream Time Using Reciprocating Compressor Monitoring." Paper presented at the NPRA Maintenance Conference, New Orleans, Louisiana (May 1997).

Souders, M., and Brown, G.G., Design and fractionating columns. 1. Entrainment and capacity. *Ind. Eng. Chem.* **26**(1), 98 (1934).

Suarez, F.J., "Pluses and minuses of caustic treating," Hydrocarbon Processing" **75**(10) 117–123 (October 1996).

Svrcek, W.Y., and Monnery, W.D., Design two-phase separators within the rigth limits. *Chem. Engin. Prog.* **89**(10), 53–60 (1993).

Talavera, P.G., "Selecting gas/liquid separators," Hydrocarbon Processing." 81–84 (June 1990).

Wang, S., Gomez, L.E., Mohan, R.S., Shoham, O., and Kouba, G.E., Gas-liquid cylindrical cyclone (GLCC) compact separators for wet gas applications. *ASME J. Energy Res. Technol.* **125**(1), 43–50 (2003).

Williamson, K., Tousi, S., and Hashemi, R., "Recent Developments in Performance Rating of Gas/Liquid Coalescers." Paper presented at the First Annual Meeting of the American Filtration Society, Ocean City, Maryland (March 1988).

Wines, T.H., and Lorentzen, C., "High Performance Liquid/Gas Coalescers for Compressor Protection." Compressor Workshop, Lambton College, Sarnia, Ontario (April 1999).

Wines, T.H., Optimize NO_x reductions facility-wide. *Hydrocarb. Proc* **83**(4), 53–59 (2004).

Xiao, J.J., and Shoup, G., Sizing wet-gas pipelines and slug catchers with steady-state multiphase flow simulations. *ASME J. Energy Res. Technol.* **120,** 106–110 (1998).

CHAPTER 6

CONDENSATE STABILIZATION

6.1 INTRODUCTION

Hydrocarbon condensate recovered from natural gas may be shipped without further processing but is stabilized often for blending into the crude oil stream and thereby sold as crude oil. In the case of raw condensate, there are no particular specifications for the product other than the process requirements. The process of increasing the amount of intermediates (C_3 to C_5) and heavy (C_6^+) components in the condensate is called "condensate stabilization." This process is performed primarily in order to reduce the vapor pressure of the condensate liquids so that a vapor phase is not produced upon flashing the liquid to atmospheric storage tanks. In other word, the scope of this process is to separate the very light hydrocarbon gases, methane and ethane in particular, from the heavier hydrocarbon components (C_3^+). Stabilized liquid, however, generally has a vapor pressure specification, as the product will be injected into a pipeline or transport pressure vessel, which has definite pressure limitations. Condensates may contain a relatively high percentage of intermediate components and can be separated easily from entrained water due to its lower viscosity and greater density difference with water. Thus, some sort of condensate stabilization should be considered for each gas well production facility. The purpose of this chapter is to describe the basic processes used to condensate stabilization and associated equipment design procedure.

6.2 STABILIZATION PROCESSES

Stabilization of condensate streams can be accomplished through either flash vaporization or fractionation.

6.2.1 Flash Vaporization

Stabilization by flash vaporization is a simple operation employing only two or three flash tanks. This process is similar to stage separation utilizing the equilibrium principles between vapor and condensate phases. Equilibrium vaporization occurs when the vapor and condensate phases are in equilibrium at the temperature and pressure of separation. Figure 6-1 shows a typical scheme of condensate stabilization through the flash vaporization process. As shown, condensate from the inlet separator after passing through the exchanger enters to the high-pressure flash tank, where the pressure is maintained at 600 psia. A pressure drop of 300 psia is obtained here, which assists flashing of large amounts of lighter ends, which join the sour vapor stream after recompression. The vapor can either be processed further and put into the sales gas or be recycled into the reservoir and used as gas lift to produce more crude oils. The bottom liquid from the high-pressure tank flows to the middle pressure flash tank operated at 300 psia. Additional methane and ethane are released in this tank. The bottom product is withdrawn again to the low-pressure tank

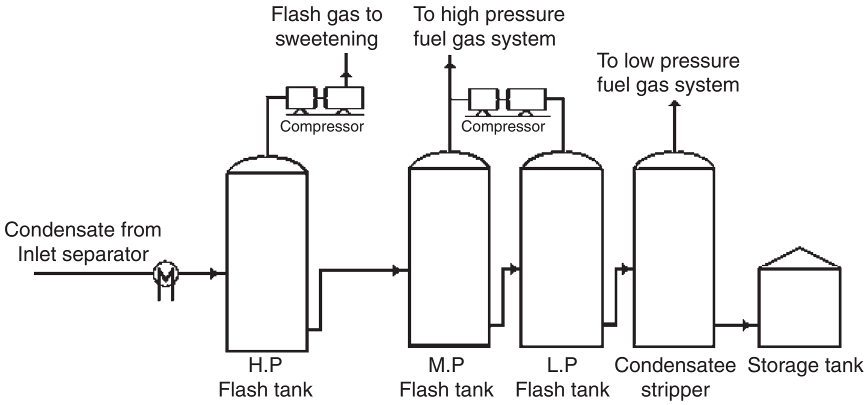


Figure 6-1. Schematic of condensate stabilization through flash vaporization process. H.P., high pressure; M.P., middle pressure; L.P., low pressure.

operated at 65 psia. To ensure efficient separation, condensate is degassed in the stripper vessel at the lowest possible pressure prior to storage. This reduces excess flashing of condensate in the storage tank and reduces the inert gas blanket pressure required in it.

Note that flash vaporization as a condensate stabilization method is old technology and is not used in a modern gas plant. However, variations of flash technology might also be found on oil production facilities stabilizing crude oil.

6.2.2 Stabilization by Fractionation

Stabilization by fractionation is a detailed process, very popular in the industry and precise enough to produce liquids of suitable vapor pressure. During the operation, light fractions such as methane–ethane–propane and most of the butanes are removed and recovered. The finished product from the bottom of the column is composed mainly of pentanes and heavier hydrocarbons, with small amounts of butane. The process actually makes a cut between the lightest liquid component (pentane) and the heaviest gas (butane). The bottom product is thus a liquid free of all gaseous components able to be stored safely at atmospheric pressure. Stabilization by fractionation is a modern operation and economically attractive next to flash vaporization. It is a single tower process, as only one specification product is required. The bottom product of the column is capable of meeting any kind of rigid specifications with the proper operating conditions.

6.2.2.1 Process Description

Figure 6-2 shows a schematic condensate stabilization process. The liquid hydrocarbon (condensate) is brought into the system from the inlet separator and preheated in the stabilizer feed/bottoms exchanger before entering the stabilizer feed drum.[1] Liquid from the feed drum is fed to the stabilization tower at approximately 50 to 200 psi depending on whether they are sour (sour stabilization is carried out at the low end of the range and sweet stabilization at the high end of the range).

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[1] Sometimes the liquid is flashed down to a feed drum at pressure slightly above the tower pressure. This flashes off vapor so that the stabilization tower can often be a smaller diameter.

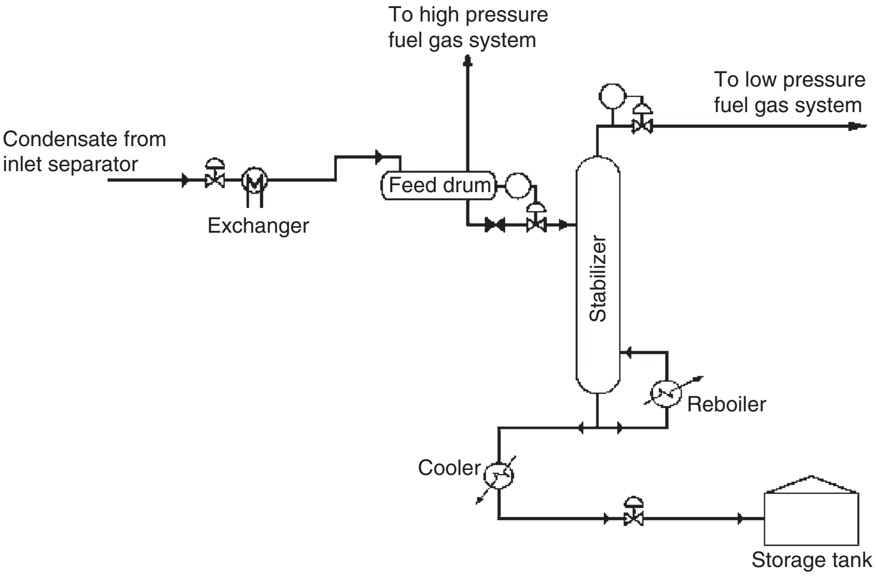

Figure 6-2. Schematic of a condensate stabilization system.

The condensate stabilizer reduces vapor pressure of the condensate by removing the lighter components. The stabilization is typically carried out in a reboiled absorber, with tray type internals. However, if a better separation is required, typically the column is changed from a top feed reboiled absorber to a refluxed distillation tower. As the liquid falls into the column, it becomes leaner in light ends and richer in heavy ends. At the bottom of the tower some of the liquid is circulated through a reboiler to add heat to the tower. As the gas goes up from tray to tray, more and more of the heavy ends get stripped out of the gas at each tray and the gas becomes richer in the light ends and leaner in the heavy ends. Overhead gas from the stabilizer, which would seldom meet market specifications for the natural gas market, is then sent to the low-pressure fuel gas system through a back-pressure control valve that maintains the tower pressure to set point. Liquids leaving the bottom of the tower have undergone a series of stage flashes at ever-increasing temperatures, driving off the light components, which exit the top of the tower. These liquids must be cooled to a sufficiently low temperature to keep vapors from flashing to atmosphere in the condensate storage tank.

Most gas processing plants are implementing advanced process control[2] (APC) on their condensate production systems in order to maximize condensate yields, improve stability of the condensate stabilization process, and ensure that product quality limits are adhered to at all times. However, most often, the main reason to implement APC on this unit is product quality and control, where the gas suppliers sell the condensate product with a Reid vapor pressure (RVP) defined by the customer. In this instance, a distributed control system is used by the APC system to maintain RVP within quality limits and push this to a higher specification; therefore, generating greater revenues as condensate throughput is increased (Hotblack, 2004).

From a control standpoint, some optimization schemes include:

- Nonlinear level control concepts on the feed drum can smooth the flow to the stabilizer column. This allows the feed drum to be truly used as a capacitance. The feed rate to the stabilizer is not changed if the level in the feed drum is within a dead band and not changing too quickly. As the level approaches the dead band limits, flow is gradually changed. As the level moves outside the dead band, then the feed rate is moved more aggressively.
- Predictive models can be employed to anticipate the effects of feed rate and compositional effects on the bottom composition.
- An inferential property for bottom RVP can be determined based on trays with sensitive pressure-compensated temperatures (PCT) to improve the control of the product quality. Alternatively, the most sensitive tray temperature or PCT can be cascaded to the flow of heat input to the reboiler.
- Separation efficiency is improved as tower pressure is lowered. This can reduce heat requirements for reboiling while making a better separation between light and heavy key component in the tower. The pressure can be lowered subject to maximum valve opening of the pressure control valve or column flooding as indicated by differential pressure

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[2] APC technology uses a multivariable control technique based on a linear dynamic process model. Using this predictive model, the controller is able to calculate an optimum set of manipulated variable moves, which minimize the error between actual and desired process behavior subject to process constraints. The controller is able to take account of process interactions and overcome process disturbances to reduce the standard deviation of key controlled variables.

measurement across the tower. The tower is more susceptible to flooding at lower pressures. If an overhead compressor is employed to boost the tower overhead up to fuel system pressure, then the pressure can be lowered subject to the maximum speed of the compressor driver, rod load limitations for a reciprocating compressor, or surge considerations for a centrifugal compressor.

6.2.2.2 Design Considerations of Stabilization Column

In most cases of lease operation, the stabilization column will operate as a nonrefluxed tower. This type of operation is simpler but less efficient than the refluxed tower operation. Because the nonrefluxed tower requires no external cooling source, it is particularly applicable to remote locations. A condensate stabilization column with reflux will recover more intermediate components from the gas than a cold-feed stabilizer. However, it requires more equipment to purchase, install, and operate. This additional cost must be justified by the net benefit of the incremental liquid recovery, less the cost of natural gas shrinkage and loss of heating value, over that obtained from a cold-feed stabilizer. When a condenser is used in a stabilization column, it will always be a partial condenser because of the quantities of methane and ethane that must be removed from the tower feed.

The stabilization tower pressure depends on the amount of liquid to be stabilized and whether it is sweet or sour. For sweet stabilization, the pressure should be as high as possible to minimize overhead vapor recompression, as this gas is remixed with the separator vapor. This also tends to decrease the cost of reflux cooling, if it is used. However, relative volatility of the components also decreases with pressure and, as stated previously, driving H_2S overhead requires a relatively low pressure.

Figure 6-3 shows the maximum recommended feed temperature to a stabilizer as a function of operating pressure of the stabilizer. An exception to this may be the case where either sour or small quantities of liquid are being handled and where first cost therefore is very critical. In these cases, many times a 40- to 70- psia working pressure for a nonrefluxed tower may show an economic advantage (Campbell, 1992).

In some cases the reboiler for the stabilizer will be an indirect salt bath heater or a steam-fired heat exchanger. Figure 6-4 shows suggested bottom (reboiler) temperatures for producing a specified Reid vapor pressure product. In fact, the temperature used on the bottom is limited by

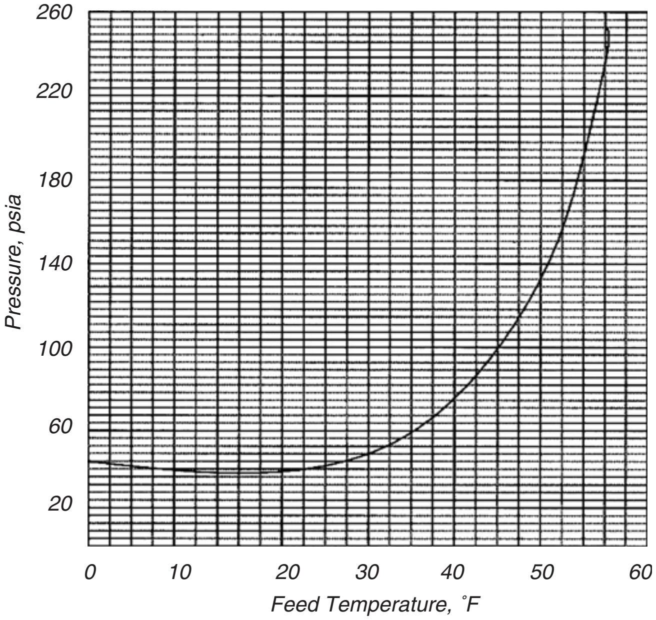

Figure 6-3. Maximum recommended feed temperature to a cold-feed stabilizer (Campbell, 1992).

the thermal breakdown characteristics of the condensate and, of course, product specs.

After the pressure has been chosen and the operating temperatures have been established through use of Figures 6-3 and 6-4, the split in the tower must be predicted. There are several methods in which this can be done, but one of the most convenient manual methods involves utilization of pseudo-equilibrium constant (K) values for each component between the top and the bottom of the tower. Using this concept, the separation that can be achieved across a nonrefluxed stabilizer can be estimated by use of the pseudo K values and a simple flash calculation. The vapor from the flash calculation will be the composition of the overhead product, and the liquid from the flash calculation will be the composition of the bottom liquid. Today, distillation and absorption/stripping calculations are done most often with process simulation computer software.

Note that for estimating the desired composition of the bottom liquid if a split of nC_4 (normal butane) is assumed, the mole fraction of each

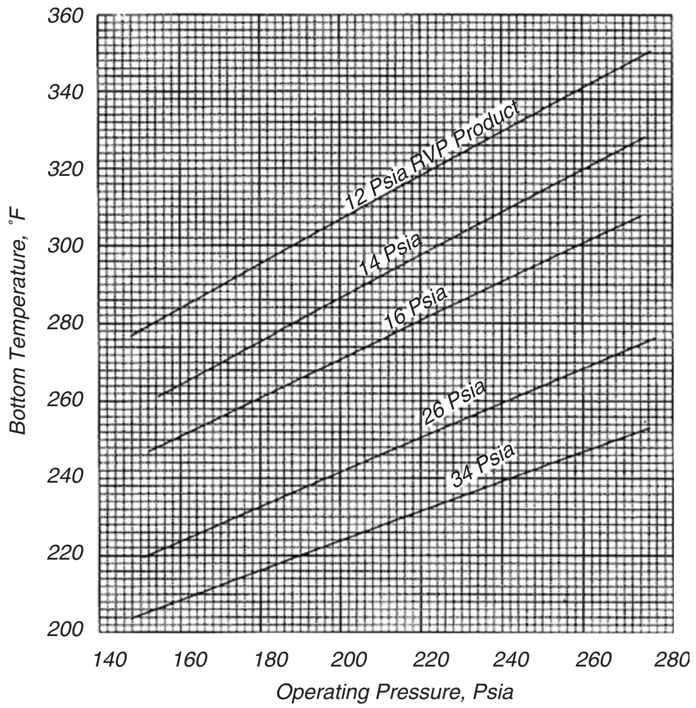

Figure 6-4. Estimation of proper bottom temperature of a nonrefluxed stabilizer (Campbell, 1992).

component in the liquid can be estimated from the following equations:

$$L_i = \frac{F_i(\text{nC}_4 \text{ split})}{\text{RV}_i} \tag{6-1}$$

$$X_i = \frac{L_i}{\sum_{i=1}^{n} L_i} \tag{6-2}$$

where X_i is the mole fraction of component i in the liquid, F_i is total number of moles of component i in the feed, L_i is total number of moles of component i in the bottom liquid, nC_4 split is assumed moles of component nC_4 in the bottom liquid divided by moles of nC_4 in the feed, RV_i is relative volatility of component i from Table 6-1, and n is number of components in the bottom liquid.

Table 6-1
RVP and Relative Volatility of Various Components (Reid *et al.*, 1977)

Component	RVP psia	Relative volatility
C_1	5000	96.9
C_2	800	15.5
C_3	190	3.68
$i\text{-}C_4$	72.2	1.40
$n\text{-}C_4$	51.6	1.00
$i\text{-}C_5$	20.4	0.40
$n\text{-}C_5$	15.6	0.30
C_6	5.0	0.10
C_7^+	≈0.1	0.00
CO_2	—	Infinite
N_2	—	Infinite
H_2S	394	7.64

The vapor pressure is the primary property used to make this split. It is assumed that the mole fraction of each component times its vapor pressure represents the contribution of that component to the total mixture vapor pressure. The total mixture vapour pressure can then be computed from Equation (6-3):

$$P_V = \sum_{i=1}^{n} (P_{Vi} \times X_i) \tag{6-3}$$

where P_V is vapor pressure of mixture, psia, and P_{Vi} is vapor pressure of component i, psia.

If the vapor pressure of the mixture is higher than the desired RVP of the bottom liquid, choose a lower number for the nC_4 split. If the calculated vapor pressure is lower than the desired RVP, choose a higher number for the nC_4 split. Iterate until the calculated vapor pressure equals the desired RVP of the bottom liquid.

The bottom liquid temperature can also be determined by calculating the bubble point of the liquid described by the previous iteration at the chosen operating pressure in the tower. This is done by choosing a temperature, determining pseudo K values from Figure 6-5, and computing

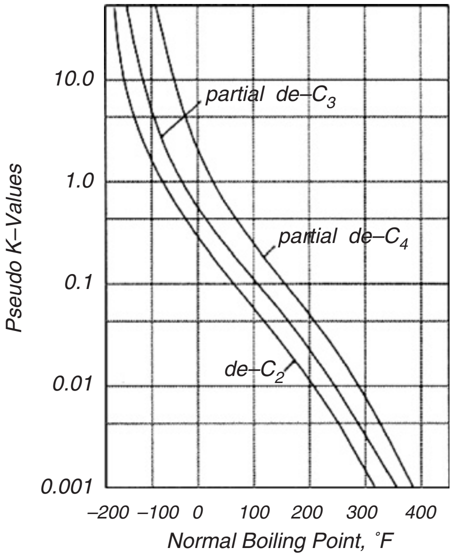

Figure 6-5. Pseudo *K* values for cold feed stabilizers (Campbell, 1992).

parameter C by the following equation:

$$C = \sum_{i=1}^{n} (L_i \times K_i) \tag{6-4}$$

If C is greater than 1.0, the assumed temperature is too high. If C is lower than 1.0, the assumed temperature is too low. By iteration a temperature can be determined where $C = 1.0$. Typically, bottoms temperatures will range from 200 to 400°F depending on operating pressure, bottoms composition, and vapor pressure requirements. Temperatures should be kept to a minimum to decrease the heat requirements, limit salt buildup, and prevent corrosion problems.

6.3 CONDENSATE STORAGE

Condensate is stored between production and shipping operations in condensate storage tanks, which are usually of floating roof type (external and internal). If the condensate does not meet the specifications, the off-specification condensate may be routed to an off-specification condensate storage fixed roof tank (vertical and horizontal) until it is recycled to the condensate stabilization unit by the relevant recycle pump if the latter is available at the plant. The primary quality criterion for the condensate is its RVP, which is affected by atmospheric pressure (plant elevation) and maximum abmient temperature. To store the condensate in floating roof storage tanks, it is very crucial to control the RVP at the desired level (especially in warm seasons). Emissions from condensate storage tanks are normally categorized as occurring from breathing losses (standing storage losses) or working losses. The term breathing loss refers to those emissions that result without any corresponding change in the liquid level within the tank. Most likely, these types of emissions result from hydrocarbon vapors that are released from the tank by expansion or contraction caused by changes in either temperature or pressure. Working loss represents those emissions that occur due to changes in the liquid level caused by either filling or emptying the tank itself (U.S. EPA's AP-42 manual). For floating roof tanks, breathing losses are a result of evaporative losses through rim seals, deck fittings, and deck seam losses. Withdrawal losses occur as the level drops, and thus the floating roof is lowered. Some liquid remains on the inner tank wall surface and evaporates when the tank is emptied. For an internal floating roof tank that has a column-supported fixed roof, some liquid also clings to the columns and evaporates. Evaporative loss occurs until the tank is filled and the exposed surfaces are again covered (SCAQMD supplemental manual, 2003).

The working pressure required to prevent breathing and thereby save standing storage losses depends on the vapor pressure of the product, the temperature variations of the liquid surface and the vapor space, and the setting of the vacuum vent. When these factors are known, the storage pressure required to eliminate venting can be computed by the Equation (6-5) (GPSA, 1998):

$$P_S = P_{V,\max} + \left(P_T - P_{V,\min}\right)\left[\frac{T_{\max} + 460}{T_{\min} + 460}\right] - P_{\text{atm}} \tag{6-5}$$

where P_S is required storage pressure, psia; $P_{V,\max}$ is true vapor pressure of liquid at maximum surface temperature, psia; $P_{V,\min}$ is true vapor pressure of liquid at minimum surface temperature, psia; $T_{\max}$ is maximum average temperature of vapor, °F; $T_{\min}$ is minimum average temperature of vapour, °F; P_{atm} is atmospheric pressure, psia; and P_T is absolute internal tank pressure at which vacuum vent opens, psia.

For the condition where $P_{V,\min}$ is greater than P_T, Equation (6-5) can be represented as

$$P_S = P_{V,\max} - P_{\mathrm{atm}} \tag{6-6}$$

With the situation represented by Equation (6-5), gas is admitted to the vapor space through the vacuum vent, where a gas vent system should be employed for preventing safety problems. However, for the situation represented by Equation (6-6), gas may be purged and kept out of the tank (Campbell, 1992).

Maximum liquid surface temperatures vary from 85 to 115°F. Sufficient accuracy will generally result from the assumption that it is 10°F higher than the maximum temperature of the body of the liquid in a tank at that location.

Note that true vapor pressure (TVP) may be the most difficult term in the aforementioned equation to calculate. A nomograph has been devised that relates TVP to both the Reid vapor pressure and the storage temperature (T_S). Numerically, the relationship between TVP, RVP and storage temperature can be expressed as follows (TRW Environmental, Inc., 1981).

$$\mathrm{TVP} = (\mathrm{RVP})\,\mathrm{EXP}\left\{C_0\left(C_1 - \frac{1}{559.69}\right)\right\} \tag{6-7}$$

where C_0 is constant dependent on the value of RVP, $C_1 = 1/(T_\mathrm{S} + 460)$, and T_S is temperature of the stored liquid , °F.

In the aforementioned equation, the term C_0 is dependent on the given value of RVP as shown in Table 6-2 (TRW Environmental, Inc., 1981).

Once the parameters C_0, T_S, and RVP have been determined, one is technically able to calculate a value for TVP. It should be noted, however, that an error was discovered in the API nomograph-calculated values of TVP, such that the RVP was not equal to the TVP at 100°F as should be expected, given the general definition of RVP. Using linear regression methods, a correction factor (C_F) was developed that should be added to the calculated values of TVP in order to obtain correct TVP values as follows (TRW Environmental, Inc., 1981).

$$\text{Corrected TVP} = \text{Calculated TVP} + C_\mathrm{F} \tag{6-8}$$

Table 6-2
C_0 for Different RVP Values

RVP	C_0	RVP	C_0
$2 < \text{RVP} < 3$	−6439.2	$\text{RVP} = 7$	−6109.8
$\text{RVP} = 3$	−6255.9	$7 < \text{RVP} < 8$	−6238.9
$3 < \text{RVP} < 4$	−6212.1	$\text{RVP} = 8$	−6367.9
$\text{RVP} = 4$	−6169.2	$8 < \text{RVP} < 9$	−6477.5
$4 < \text{RVP} < 5$	−6177.9	$\text{RVP} = 9$	−6587.0
$\text{RVP} = 5$	−6186.5	$9 < \text{RVP} < 10$	−6910.5
$5 < \text{RVP} < 6$	−6220.4	$\text{RVP} = 10$	−7334.0
$\text{RVP} = 6$	−6254.3	$10 < \text{RVP} < 15$	−8178.0
$6 < \text{RVP} < 7$	−6182.1	$\text{RVP} > 15$	−9123.2

The correction factor was found to be dependent on RVP according to the following equations:

$$\text{RVP} < 3 : C_\text{F} = 0.04(\text{RVP}) + 0.1 \tag{6-9}$$

$$\text{RVP} > 3 : C_\text{F} = \text{EXP}\{2.345206 \log(\text{RVP}) - 4.132622\} \tag{6-10}$$

It is important to mention that the methods used for calculating each tank emissions are described in detail in the U.S. EPA's AP-42 manual. In this manual, equations are developed to calculate emissions for fixed roof tanks. By assuming that open top tanks and tanks with open holes or roof openings do not have emissions greater than those for fixed roof tanks, the State California Air Resource Board was able to use the AP-42 equations to calculate emissions for all of the storage tanks in a particular oil production field.

REFERENCES

Campbell, J.M., "Gas Conditioning and Processing," 3rd Ed. Campbell Petroleum Series, Norman, OK (1992).

GPSA Engineering Data Book, 11th Edition, Vol. 1, Section 6, Gas Processors Suppliers Association, Tulsa, OK (1998).

Hotblack, C., BG Tunisia's advanced process control improves condensate product stability. *World Oil* **225**, 9 (2004).

Reid, R., Prausnitz, J.M., and Sherwood, T., "The Properties of Gases and Liquids," 3rd Ed. McGraw-Hill, New York (1977).

SCAQMD Supplemental Manual, "Supplemental Instructions for Liquid Organic Storage Tanks and References." Annual Emissions Reporting Program, South Coast Air Quality Management District, CA (June 2003).

TRW Environmental, Inc., "Background Documentation for Storage of Organic Liquids." EPA Contract No. 68-02-3174, NC (May 1981).

U.S. EPA's AP-42 Manual, "Compilation of Air Pollutant Emission Factors," Chapter 7, Vol. 1, 5th Ed. U.S. Environmental Protection Agency, Washington, DC.

CHAPTER 7

ACID GAS TREATING

7.1 INTRODUCTION

Natural gas, while ostensibly being hydrocarbon in nature, contains large amounts of acid gases, such as hydrogen sulfide and carbon dioxide. Natural gas containing hydrogen sulfide or carbon dioxide is referred to as *sour*, and natural gas free from hydrogen sulfide is referred to as *sweet*. The corrosiveness nature of hydrogen sulfide and carbon dioxide in the presence of water (giving rise to an acidic aqueous solution) and because of the toxicity of hydrogen sulfide and the lack of heating value of carbon dioxide, natural gas being prepared for sales is required to contain no more than 5 ppm hydrogen sulfide and to have a heating value of no less than 920 to 980 Btu/scf. The actual specifications depend on the use, the country where the gas is used, and the contract. However, because natural gas has a wide range of composition, including the concentration of the two acid gases, processes for the removal of acid gases vary and are subject to choice based on the desired end product.

There are many variables in treating natural gas. The precise area of application of a given process is difficult to define. Several factors must be considered: (1) types and concentrations of contaminants in the gas, (2) the degree of contaminant removal desired, (3) the selectivity of acid gas removal required, (4) the temperature, pressure, volume, and composition of the gas to be processed, (5) the carbon dioxide–hydrogen sulfide ratio in the gas, and (6) the desirability of sulfur recovery due to process economics or environmental issues.

In addition to hydrogen sulfide and carbon dioxide, gas may contain other contaminants, such as mercaptans and carbonyl sulfide. The presence of these impurities may eliminate some of the sweetening processes, as some processes remove large amounts of acid gas but not to a sufficiently low concentration. However, there are those processes that are not designed to remove (or are incapable of removing) large amounts of acid gases. These processes are also capable of removing the acid gas impurities to very low levels when the acid gases are there in low to medium concentrations in the gas.

Process selectivity indicates the preference with which the process removes one acid gas component relative to (or in preference to) another. For example, some processes remove both hydrogen sulfide and carbon dioxide; other processes are designed to remove hydrogen sulfide only. It is important to consider the process selectivity for, say, hydrogen sulfide removal compared to carbon dioxide removal that ensures minimal concentrations of these components in the product, thus the need for consideration of the carbon dioxide to hydrogen sulfide in the gas stream.

The focus on this chapter is the treatment of natural gas to remove the acid gases, carbon dioxide and hydrogen sulfide. However, because the processes are integrated within the concept of gas treating or acid gas removal or acid gas treating, some consideration of the other processes for natural gas purification (preparation for sales) must also be made.

7.2 ACID GAS REMOVAL PROCESSES

The processes that have been developed to accomplish gas purification vary from a simple once-through wash operation to complex multistep recycling systems (Speight, 1993). In many cases, process complexities arise because of the need for recovery of the materials used to remove the contaminants or even recovery of the contaminants in the original, or altered, form (Kohl and Riesenfeld, 1985; Newman, 1985).

There are two general processes used for acid gas removal: adsorption and absorption (Speight, 1993). Adsorption is a physical–chemical phenomenon in which the gas is concentrated on the surface of a solid or liquid to remove impurities. Usually, carbon is the adsorbing medium, which can be regenerated upon desorption [Speight (1993, 1999) and references cited therein]. The quantity of material adsorbed is proportional to the surface area of the solid and, consequently, adsorbents are usually granular solids with a large surface area per unit mass. Subsequently, the captured gas can be desorbed with hot air or steam either for recovery or for thermal destruction. Adsorbers are widely used to increase a low gas concentration prior to incineration unless the gas concentration is very high in the inlet air stream. Adsorption is also employed to reduce problem odors from gases. There are several limitations to the use of adsorption systems, but it is generally felt that the major one is the requirement for minimization of particulate matter and/or condensation of liquids (e.g., water vapor) that could mask the adsorption surface and reduce its efficiency drastically. Absorption differs from adsorption in that it is not a physical–chemical surface phenomenon, but an approach in which the absorbed gas is ultimately distributed throughout the absorbent (liquid). The process depends only on physical solubility and may include chemical reactions in the liquid phase (chemisorption). Common absorbing media used are water, aqueous amine solutions, caustic, sodium carbonate, and nonvolatile hydrocarbon oils, depending on the type of gas to be absorbed. Usually, the gas–liquid contactor designs that are employed are plate columns or packed beds.

Absorption is achieved by dissolution (a physical phenomenon) or by reaction (a chemical phenomenon). Chemical adsorption processes adsorb sulfur dioxide onto a carbon surface where it is oxidized (by oxygen in the flue gas) and absorbs moisture to give sulfuric acid impregnated into and on the adsorbent.

As currently practiced, acid gas removal processes involve the chemical reaction of the acid gases with a solid oxide (such as iron oxide) or selective absorption of the contaminants into a liquid (such as ethanolamine) that is passed countercurrent to the gas. Then the absorbent is stripped of the gas components (regeneration) and recycled to the absorber. The process design will vary and, in practice, may employ multiple absorption columns and multiple regeneration columns.

Liquid absorption processes [which usually employ temperatures below 50°C (120°F)] are classified either as physical solvent processes or as chemical solvent processes. The former processes employ an organic solvent, low temperatures, or high pressure. In chemical solvent processes, absorption of the acid gases is achieved mainly by use of alkaline solutions such as amines or carbonates (Kohl and Riesenfeld, 1985). Regeneration (desorption) can be brought about by the use of reduced pressures and/or high temperatures, whereby the acid gases are stripped from the solvent.

Amine washing of natural gas involves chemical reaction of the amine with any acid gases with the liberation of an appreciable amount of heat and it is necessary to compensate for the absorption of heat. Amine derivatives such as ethanolamine (monoethanolamine), diethanolamine, triethanolamine, methyldiethanolamine, diisopropanolamine, and diglycolamine have been used in commercial applications (Kohl and Riesenfeld, 1985; Speight, 1993; Polasek and Bullin, 1994).

7.2.1 Batch Tyoe Processes

The most common type of process for acid gas removal is the batch-type process and may involve a chemical process in which the acid gas reacts chemically with the cleaning agent, usually a metal oxide. These processes are not merely physical separation processes in which the acid gas is removed by a physical phenomenon, such as adsorption. Thus, the batch-type processes have the common requirement that the process be operated as a batch system where, at the end of the cycle, the chemical agent must be changed or regenerated in order to continue treating.

Batch processes are limited to removing small amounts of sulfur, i.e., low gas flow rates and/or small concentrations of hydrogen sulfide. These processes are described in great detail as follows.

7.2.1.1 Metal Oxide Processes

These processes scavenge hydrogen sulfide and organic sulfur compounds (mercaptans) from gas streams through reactions with solid-based media. They are typically nonregenerable, although some are partially regenerable, losing activity upon each regeneration cycle. Most dry sorption processes are governed by the reaction of a metal oxide with H_2S to form a metal sulfide compound. For regenerable reactions, the metal sulfide compound can then react with oxygen to produce elemental sulfur and a regenerated metal oxide. The primary metal oxides used for dry sorption processes are iron oxide and zinc oxide.

Dry sorption processes can be categorized into two subgroups: oxidation to sulfur and oxidation to oxides of sulfur. Because these processes rely on oxidation, gas constituents that cannot be oxidized under the process conditions will not be removed (Kohl and Riesenfeld, 1985). This is advantageous when dealing with biogas, as only hydrogen sulfide, mercaptans, and, in some cases, carbon dioxide will be removed, with miniscule losses of methane due to adsorption. The main product of sulfur oxidation to oxides of sulfur is sulfur dioxide. Because this is a controlled exhaust gas, poses a threat to equipment due to corrosion and poisoning of fuel cell membranes, and requires additional gas processing to achieve air discharge standards, it will not be addressed as a gas processing option.

7.2.1.1.1 Iron Sponge Process The iron sponge process or dry box process is the oldest and still the most widely used batch process for sweetening of natural gas and natural gas liquids (Duckworth and Geddes, 1965; Anerousis and Whitman, 1984; Zapffe, 1963). The process was implemented during the 19th century. Large-scale, commercial operations have, for the most part, discontinued this process due to the high labor costs of removing packed beds. However, its simplicity, low capital costs, and relatively low media cost continue to make the process an ideal solution for hydrogen sulfide removal.

The process is usually best applied to gases containing low to medium concentrations (300 ppm) of hydrogen sulfide or mercaptans. This process tends to be highly selective and does not normally remove significant quantities of carbon dioxide. As a result, the hydrogen sulfide stream from the process is usually high purity. Use of the iron sponge process for sweetening sour gas is based on adsorption of the acid gases on the surface of the solid sweetening agent followed by chemical reaction of ferric oxide (Fe_2O_3) with hydrogen sulfide:

$$2Fe_2O_3 + 6H_2S \rightarrow 2Fe_2S_3 + 6H_2O \tag{7-1}$$

The reaction requires the presence of slightly alkaline water and a temperature below 43°C (110°F) and bed alkalinity should be checked regularly, usually on a daily basis. A pH level on the order of 8–10 should be maintained through the injection of caustic soda with the water. If the gas does not contain sufficient water vapor, water may need to be injected into the inlet gas stream.

The ferric sulfide produced by the reaction of hydrogen sulfide with ferric oxide can be oxidized with air to produce sulfur and regenerate the ferric oxide:

$$2Fe_2S_3 + 3O_2 \rightarrow 2Fe_2O_3 + 6S \tag{7-2}$$

$$S_2 + 2O_2 \rightarrow 2SO_2 \tag{7-3}$$

The regeneration step, i.e., the reaction with oxygen, is exothermic and air must be introduced slowly so the heat of reaction can be dissipated.

If air is introduced quickly, the heat of reaction may ignite the bed. Some of the elemental sulfur produced in the regeneration step remains in the bed. After several cycles this sulfur will cake over the ferric oxide, decreasing the reactivity of the bed. Typically, after 10 cycles the bed must be removed and a new bed introduced into the vessel.

In some designs the iron sponge may be operated with continuous regeneration by injecting a small amount of air into the sour gas feed. The air regenerates ferric sulfide while hydrogen sulfide is removed by ferric oxide. This process is not as effective at regenerating the bed as the batch process and requires a higher pressure air stream (Arnold and Stewart, 1999).

In the process as shown in Figure 7-1, the sour gas should pass down through the bed. In the case where continuous regeneration is to be utilized, a small concentration of air is added to the sour gas before it is processed. This air serves to regenerate the iron oxide continuously, which has reacted with hydrogen sulfide, which serves to extend the onstream life of a given tower but probably serves to decrease the total amount of sulfur that a given weight of bed will remove.

The number of vessels containing iron oxide can vary from one to four. In a two-vessel process, one of the vessels would be onstream removing hydrogen sulfide from the sour gas while the second vessel would either be in the regeneration cycle or having the iron sponge bed replaced.

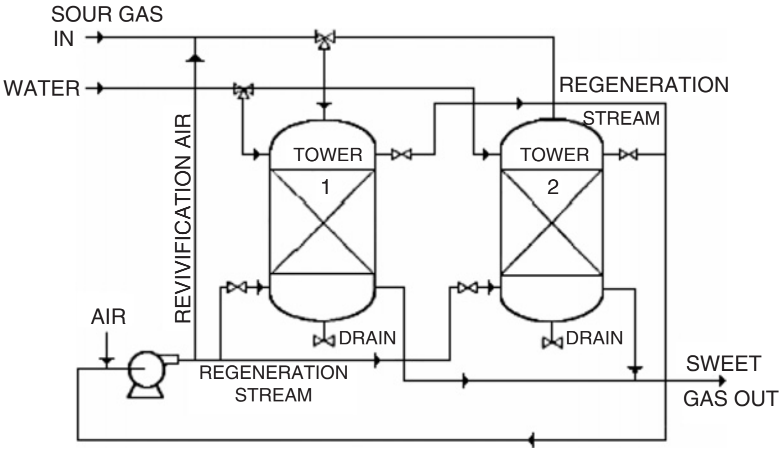

Figure 7-1. Typical iron oxide process flow sheet (Maddox, 1974).

When periodic regeneration is used, a tower is operated until the bed is saturated with sulfur and hydrogen sulfide begins to appear in the sweetened gas stream. At this point the vessel is removed from service and air is circulated through the bed to regenerate the iron oxide. Regardless of the type of regeneration process used, a given iron oxide bed will lose activity gradually and eventually will be replaced. For this reason the vessels in Figure 7-1 should be designed to minimize difficulties in replacing the iron sponge in the beds. The change out of the beds is hazardous. Exposure to air when dumping a bed can cause a sharp rise in temperature, which can result in spontaneous combustion of the bed. Care must be exercised in opening the tower to the air. The entire bed should be wetted before beginning the change out operation.

Iron oxide reactors have taken on a number of configurations from conventional box vessels to static tower purifiers. The selection depends on the process application. Static tower purifiers are used in high-pressure applications, as a longer bed depth gives a greater efficiency and the total pressure drop is a smaller fraction of the available pressure. Static tower purifiers are also fitted with trays to eliminate compaction for lower pressure applications (Gollmar, 1945).

Conventional boxes consist of large rectangular vessels, either built into the ground or supported on legs to save footprint. They are composed of several layers, with a typical bed depth of at least 2 feet.

Iron oxide suspensions, such as iron oxide slurries, rely on hydrated ferric oxide as the active regenerable agent. However, iron oxide suspensions react in a basic environment with an alkaline compound, followed by the reaction of the hydrosulfide with iron oxide to form iron sulfide (Kohl and Nielsen, 1997). The iron is then regenerated by aeration.

7.2.1.1.2 Zinc Oxide Process Zinc oxide is also used for hydrogen sulfide removal from the gas stream. The zinc oxide media particles are extruded cylinders 3–4 mm in diameter and 4–8 mm in length (Kohl and Nielsen, 1997). This uniform sizing allows for relatively accurate pressure drop calculations in designing reactors. The general reaction of zinc oxide with hydrogen sulfide is

$$ZnO + H_2S \rightarrow ZnS + H_2O \tag{7-4}$$

At increased temperatures (400 to 700°F), zinc oxide has a rapid reaction rate, therefore providing a short mass transfer zone, resulting in a short length of unused bed and improved efficiency. At operating temperatures, the zinc oxide sorbent has a maximum sulfur loading of 0.3 to 0.4 kg sulfur/kg sorbent. In large industrial plants, this process achieves a high degree of efficiency, as the spent zinc oxide bed can be used for metal and sulfur recovery (Schaack and Chan, 1989).

Generally, the iron oxide process is suitable only for small to moderate quantities of hydrogen sulfide. Approximately 90% of the hydrogen sulfide can be removed per bed, but bed clogging by elemental sulfur occurs and the bed must be discarded, and the use of several beds in series is not usually economical. Removal of larger amounts of hydrogen sulfide from gas streams requires a continuous process, such as the ferrox process or the Stretford process. The *ferrox process* is based on the same chemistry as the iron oxide process except that it is fluid and continuous. The *Stretford process* employs a solution containing vanadium salts and anthraquinone disulfonic acid (Maddox, 1974).

7.2.1.2 Slurry Processes

Slurry processes were developed as alternatives to iron sponge. Slurries of iron oxide have been used to selectively absorb hydrogen sulfide (Fox, 1981; Kattner *et al.,* 1986). The results have been inconsistent; problems include obtaining stoichiometric use, premature hydrogen sulfide breakthrough, and foaming. Also, high corrosion rates have been reported. This has been circumvented by coating the contact vessels with phenolic or epoxy resins (Samuels, 1988). The chemical cost for these processes is higher than that for iron sponge process, but this is partially offset by the ease and lower cost with which the contact tower can be cleaned out and recharged. Also, because liquids freeze and foam, antifreeze must be added in winter, and an antifoamant agent should be available. Finally, obtaining approval to dispose of the spent chemicals, even if they are non-hazardous, is time-consuming. Two different slurry processes are described here in great detail.

7.2.1.2.1 Chemsweet Process The Chemsweet process is a batch process for the removal of hydrogen sulfide from natural gas (Manning, 1979). Chemicals used are a mixture of zinc oxide, zinc acetate, water, and a dispersant to keep the zinc oxide particles in suspension. When one part is mixed with five parts of water the acetate dissolves and provides a controlled source of zinc ions that react instantaneously with the bisulfide and sulfide ions that are formed when hydrogen sulfide dissolves in water. The zinc oxide replenishes the zinc acetate. The following reactions are performed in a Chemsweet process.

$$\text{Sweetening:}\quad ZnAc_2 + H_2S \leftrightarrow ZnS + 2HAc \tag{7-5}$$

$$\text{Regeneration:}\quad ZnO + 2HAc \leftrightarrow ZnAc_2 + H_2O \tag{7-6}$$

$$\text{Overall:}\quad ZnO + H_2S \leftrightarrow ZnS + H_2O \tag{7-7}$$

The presence of carbon dioxide in the natural gas is of little consequence as the pH of the slurry is low enough to prevent significant absorption of carbon dioxide, even when the ratio of CO_2 to H_2S is high (Manning and Thompson, 1991).

The Chemsweet process can treat gas streams with a high hydrogen sulfide concentration and has been operated between pressures of 89 and 1415 psia (Kutsher and Smith, 1966). Mercaptan concentrations in excess of 10% of the hydrogen sulfide concentration in the gas stream can be a problem. Some of the mercaptans will react with the zinc oxide and be removed from the gas. The resulting zinc mercaptides [Zn(OH)RH] will form a sludge and possibly cause foaming problems (GPSA, 1998).

7.2.1.2.2 Sulfa-Check Process The Sulfa-Check process selectively removes hydrogen sulfide and mercaptans from natural gas in the presence of carbon dioxide (Bhatia and Allford, 1986; Dobbs, 1986). The process uses sodium nitrite ($NaNO_2$).

Gas streams with elevated oxygen levels with the Sulfa-Check process will produce some nitrogen oxides in the gas stream (Schaack and Chan, 1989). Removal of hydrogen sulfide is not affected under short contact times, as the reaction is almost instantaneously (Hohlfeld, 1979). Sodium hydroxide and sodium nitrite are consumables in the processes and cannot be regenerated.

This process is accomplished in a one-step single vessel design using an aqueous solution of sodium nitrite buffered to stabilize the pH above 8. Also, there is enough strong base to raise the pH of the fresh material to 12.5.

The claimed reaction with hydrogen sulfide forms elemental sulfur, ammonia, and caustic soda as follows:

$$NaNO_2 + 3H_2S \leftrightarrow NaOH + NH_3 + 3S + H_2O \tag{7-8}$$

Other reactions forming the oxides of nitrogen do occur (Burnes and Bhatia, 1985), and carbon dioxide in the gas reacts with the sodium hydroxide to form sodium carbonate and sodium bicarbonate. The spent solution is a slurry of fine sulfur particles in a solution of sodium and ammonium salts (Manning and Thompson, 1991).

7.2.2 Amine Processes

Chemical absorption processes with aqueous alkanolamine solutions are used for treating gas streams containing hydrogen sulfide and carbon dioxide. However, depending on the composition and operating conditions of the feed gas, different amines can be selected to meet the product gas specification. Amines are categorized as being primary, secondary, and tertiary depending on the degree of substitution of the central nitrogen by organic groups. Primary amines react directly with H_2S, CO_2, and carbonyl sulfide (COS). Examples of primary amines include monoethanolamine (MEA) and the proprietary diglycolamine agent (DGA). Secondry amines react directly with H_2S and CO_2 and react directly with some COS. The most common secondry amine is diethanolamine (DEA), while diisopropanolamine (DIPA) is another example of a secondary amine, which is not as common anymore in amine-treating systems. Tertiary amines react directly with H_2S, react indirectly with CO_2, and react indirectly with little COS. The most common examples of tertiary amines are methyldiethanolamine (MDEA) and activated methyldiethanolamine.

Processes using ethanolamine and potassium phosphate are now widely used. The ethanolamine process, known as the *Girbotol process*, removes acid gases (hydrogen sulfide and carbon dioxide) from liquid hydrocarbons as well as from natural and from refinery gases. The Girbotol treatment solution is an aqueous solution of ethanolamine, which is an organic alkali that has the reversible property of reacting with hydrogen sulfide under cool conditions and releasing hydrogen sulfide at high temperatures. The ethanolamine solution fills a tower called an absorber through which the sour gas is bubbled. Purified gas leaves the top of the tower, and the ethanolamine solution leaves the bottom of the tower with the absorbed acid gases. The ethanolamine solution enters a reactivator tower where heat drives the acid gases from the solution. Ethanolamine solution, restored to its original condition, leaves the bottom of the reactivator tower to go to the top of the absorber tower, and acid gases are released from the top of the reactivator.

Depending on the application, special solutions such as mixtures of amines; amines with physical solvents, such as sulfolane and piperazine; and amines that have been partially neutralized with an acid such as phosphoric acid may also be used (Bullin, 2003).

The proper selection of the amine can have a major impact on the performance and cost of a sweetening unit. However, many factors must be considered when selecting an amine for a sweetening application (Polasek and Bullin, 1994). Considerations for evaluating an amine type in gas treating systems are numerous. It is important to consider all aspects of the amine chemistry and type,[1] as the ommision of a single issue may lead to operational issues. While studying each issue, it is important to understand the fundamentals of each amine solution.

While many of the recent published papers concerning amine selection or amine conversions have focused on the utilization of MDEA over the older generic amines (Bullin *et al.,* 1990; Polasek *et al.,* 1992), there are many recent cases where these older generic amines have been the best and, even perhaps, the only choice for recent new plant design (Jenkins and Haws, 2002). MEA and DEA have found the most general application in the sweetening of natural gas streams. Even though a DEA system may not be as efficient as some of the other chemical solvents are, it may be less expensive to install because standard packaged systems are readily available. In addition, it may be less expensive to operate and maintain (Arnold and Stewart, 1999).

MEA is a stable compound and, in the absence of other chemicals, suffers no degradation or decomposition at temperatures up to its normal boiling point. MEA reacts with H_2S and CO_2 as follow:

$$2(RNH_2) + H_2S \leftrightarrow (RNH)_2S \tag{7-9}$$

$$(RNH_3)_2S + H_2S \leftrightarrow 2(RNH_3)HS \tag{7-10}$$

$$2(RNH_2) + CO_2 \leftrightarrow RNHCOONH_3R \tag{7-11}$$

These reactions are reversible by changing the system temperature. MEA also reacts with carbonyl COS and carbon disulfide (CS_2) to form heat-stable salts that cannot be regenerated.

DEA is a weaker base than MEA and therefore the DEA system does not typically suffer the same corrosion problems but does react with hydrogen

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[1] Physical properties of the most common amines are given in Appendix 3.

sulfide and carbon dioxide:

$$2R_2NH + H_2S \leftrightarrow (R_2NH)_2S \qquad (7\text{-}12)$$

$$(R_2NH_2)_2S + H_2S \leftrightarrow 2(R_2NH_2)HS \qquad (7\text{-}13)$$

$$2R_2NH + CO_2 \leftrightarrow R_2NCOONH_2R_2 \qquad (7\text{-}14)$$

DEA also removes carbonyl sulfide and carbon dislduflide partially as its regenerable compound with COS and CS_2 without much solution losses.

One key difference among the various specialty amines is selectivity toward hydrogen sulfide. Instead of removing both hydrogen sulfide and carbon dioxide, as generic amines such as MEA and DEA do, some products readily remove hydrogen sulfide to specifications, but allow controlled amounts of carbon dioxide to slip through. When the MDEA process was developed in the mid-1970s, it was principally destined for the sweetening of gases that did not require complete CO_2 removal (Blanc *et al*., 1981) or required the removal of only a controlled part of the CO_2. The selectivity of MDEA-based products can lead to more energy savings. For example, allowing carbon dioxide to remain in the treated gas reduces the amount of acid gas in the amine that needs to be regenerated, thus reducing the amount of energy required.

The use of MDEA for selective H_2S removal is based on the fact that unlike DEA, MDEA does not react directly with CO_2. This particularity, which is used to advantage for selective H_2S and controlled CO_2 removal, becomes a handicap for complete acid gas removal, especially when a substantial quantity of CO_2 has to be removed. To overcome this kinetic obstacle, activators were sought among secondary amines having high speeds of reaction with CO_2 to blend into the otherwise desirable MDEA solvent. Some advantages of the activated MDEA are (Lallemand and Minkkinen, 2001) as follows.

- Chemical solvent processes generally have a higher energy requirement than physical solvent processes, but are very selective toward hydrocarbons. However, among these processes, the activated methyl diethanolamine process, which removes hydrogen sulfide completely and carbon dioxide as required, has a low energy requirement.

- With the exception of acid gas dryness and pressure, the activated methyl diethanolamine process meets all of the necessary quality characteristics for a reinjection scheme; most remarkably the liberation of acid gases without hydrocarbons, where the presence of hydrocarbons reduces the density and water solubility of the pressurized acid gas fluids.

The only disadvantage of the activated methyl diethanolamine process is liberation of acid gases at low pressure.

A series of chemical activators used with methyl diethanolamine offers the most cost-effective answer to complete or controlled acid gas removal from sour to very sour natural gases. However, there are limitations of even the most advanced *activated methyl diethanolamine*-only based gas treatment technologies in handling very highly acid gas-loaded natural or associated oil field gases, especially for bulk acid gas removal when the acid gases are destined for cycling and/or disposal by reinjection. The Elf-activated MDEA process from TotalFinaElf is probably the most cost-effective solution today to meet the widest range of applications from complete CO_2 removal to bulk H_2S and/or CO_2 removal even for acid gas reinjection projects (Lallemand and Minkkinen, 2001).

7.2.2.1 Process Description

The general process flow diagram for an amine-sweetening plant varies little, regardless of the aqueous amine solution used as the sweetening agent (Figure 7-2). The sour gas containing H_2S and/or CO_2 will nearly always enter the plant through an inlet separator (scrubber) to remove any free liquids and/or entrained solids. The sour gas then enters the bottom of the absorber column and flows upward through the absorber in intimate countercurrent contact with the aqueous amine solution, where the amine absorbs acid gas constituents from the gas stream. Sweetened gas leaving the top of the absorber passes through an outlet separator and then flows to a dehydration unit (and compression unit, if necessary) before being considered ready for sale.

In many units the rich amine solution is sent from the bottom of the absorber to a flash tank to recover hydrocarbons that may have dissolved or condensed in the amine solution in the absorber. The rich solvent is then preheated before entering the top of the stripper column. The amine–amine heat exchanger serves as a heat conservation device and lowers total heat

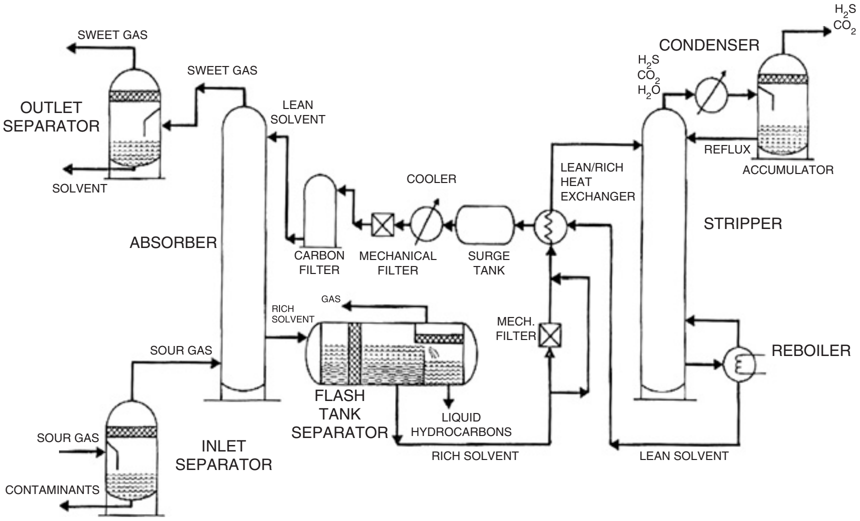

Figure 7-2. Schematic of amine gas-sweetening process flow diagram.

requirements for the process. A part of the absorbed acid gases will be flashed from the heated rich solution on the top tray of the stripper. The remainder of the rich solution flows downward through the stripper in countercurrent contact with vapor generated in the reboiler. The reboiler vapor (primarily steam) strips the acid gases from the rich solution. The acid gases and the steam leave the top of the stripper and pass overhead through a condenser, where the major portion of the steam is condensed and cooled. The acid gases are separated in the separator and sent to the flare or to processing. The condensed steam is returned to the top of the stripper as reflux.

The lean amine solution from the bottom of the stripper column is pumped through an amine–amine heat exchanger and then through a cooler before being introduced to the top of the absorber column. The amine cooler serves to lower the lean amine temperature to the 100°F range. Higher temperatures of the lean amine solution will result in excessive amine losses through vaporization and also lower acid gas-carrying capacity in the solution because of temperature effects.

Experience has shown that amine gas treatment is a fouling service. Particulates formed in the plant as well as those transported into the plant can be very bothersome. A filtration scheme of mechanical and activated carbon filters is therefore important in maintaining good solution control. Mechanical filters such as cartridge filters or precoat filters remove particulate material while carbon filters remove chemical contaminants such as entrained hydrocarbons and surface-active compounds.

7.2.2.2 Design Considerations

This section describes some design aspects for the major equipment used in both MEA and DEA systems.

7.2.2.2.1 Amine Absorber Amine absorbers use countercurrent flow through a trayed or packed tower to provide intimate contact between the amine solvent and the sour gas so that the H_2S and CO_2 molecules can transfer from the gas phase to the solvent liquid phase. In tray columns, a liquid level is maintained on each tray by a weir usually 2 or 3 inches high. The gas passes up from underneath the trays through openings in the trays such as perforations, bubble caps, or valves and disperses into bubbles through the liquid, forming a froth. The gas disengages from the froth, travels through a vapor space, providing time for entrained amine solution to fall back down to the liquid on the tray, and passes through the next tray above. In packed columns the liquid solvent is dispersed in the gas stream by forming a film over the packing, providing a large surface area for CO_2 and H_2S transfer from the gas to the liquid solvent. The degree of sweetening achieved is largely dependent on the number of trays or the height of packing available in the absorber.

Twenty valve-type trays (spaced 24 inches apart) or the equivalent height in packing column are common and are often a standard design. Typically, small-diameter towers use packing, whereas larger towers use stainless steel trays. The cross-sectional area of the contactor is sized for the gas and amine flow rates, where the maximum gas superficial velocity is obtained from the Souders and Brown (1932) equation as

$$V_{\mathrm{SG}} = 0.25\left[\frac{\rho_{\mathrm{amine}} - \rho_{\mathrm{gas}}}{\rho_{\mathrm{gas}}}\right]^{0.5}, \mathrm{ft/sec} \qquad (7\text{-}15)$$

It may be necessary to reduce the gas velocity by 25 to 35% to avoid jet flooding and by 15% to allow for foaming. Use an amine velocity of 0.25 ft/sec in the downcomer (Khan and Manning, 1985).

In most cases a mist eliminator pad is installed near the gas outlet of the absorber (the distance between the top tay and the mist pad is 3 to 4 feet) to trap entrained solvent, and an outlet knockout drum, similar to the inlet separator for the gas feed, is provided to collect solvent carryover. Some contactors have a water wash consisting of two to five trays at the top of the absorber to minimize vaporization losses of amine, which is often found in low-pressure monoethanolamine systems.

Absorbers will usually have multiple feed points, allowing the option of introducing the lean amine lower in the column or at multiple trays. If carbon dioxide absorption is desired, all of the lean amine should, in general, be fed on the top tray, thus utilizing all available stages. For carbon dioxide slip, multiple feed points are better in some cases (Bullin, 2003).

7.2.2.2.2 Amine Pumps There are several different amine pumps in each of the processes. The amine booster and reflux pumps are centrifugal, preferably in-line or horizontal. Selection of the circulation pump depends on the contactor pressure and the amine circulation rate. Normally, reciprocating pumps are preferred only if the absorber pressure is very high. However, centrifugal pumps are used for low pressures (e.g., 100 psig) and multistage horizontal centrifugal pumps for high pressures (e.g., 700 psig) or high circulation rates (e.g., 300 gal/min). In sizing and rating pumps use a low positive suction pressure of 3 to 10 psig.

The circulation flow rates for amine systems can be determined from the acid gas flow rates by selecting a solution concentration and an acid gas loading. For this purpose, the following equation can be used (Arnold and Stewart, 1999):

$$Q = \frac{k(Q_{\mathrm{G}})(\mathrm{MF})}{\rho(\mathrm{WF})(\mathrm{AG})} \tag{7-16}$$

where Q is circulation rate for amine systems, gal/min; K is constant (112 for MEA system and 192 for DEA system); Q_{G} is gas flow rate, MMscfd; MF is total acid–gas fraction in inlet gas, moles acid gas/mole inlet gas; WF is amine weight fraction, lb amine/lb solution; and ρ is solution density, lb/gal at 60°F; and AG is acid gas loading, mole acid gas/mole amine.

The rich solution acid gas loading depends on the acid gas partial pressure and corrosiveness of solution. The normal range of this parameter is 0.45 to 0.52 lbmole acid-gas/lbmole amine for MEA and 0.43 to 0.73 lbmole acid-gas/lbmole amine for DEA systems.

For design, the following solution strengths and loading are recommended to provide an effective system without an excess of corrosion (GPSA, 1998):

MEA system: WF = 20 wt.%
AG = 0.33 mole acid gas/mole MEA
DEA system: WF = 35 wt.%
AG = 0.5 mole acid gas/mole DEA

For the recommended concentrations, the densities at 60°F are

$$20\%\ \text{MEA} = 8.41\ \text{lb/gal} = 0.028\ \text{mole MEA/gal}$$

$$35\%\ \text{DEA} = 8.71\ \text{lb/gal} = 0.029\ \text{mole DEA/gal}$$

Using these design limits, Equation (7-16) can be simplified to the following equation (Arnold and Stewart, 1999):

$$Q = k'(Q_G)(MF) \tag{7-17}$$

where k' is constant (201 for MEA system and 126 for DEA system).

The circulation rate determined with these equations should be increased by 10–15% to supply an excess of amine.

The solvent circulation rate should always be reduced to the minimum required to meet the treating needs for the unit subject to the constraints such as maximum loading of the rich amine. Since the reboiler duty is almost always tied directly to the circulation rate, lower circulation rates reduce the overall energy requirements. Lower circulation rates also tend to increase the CO_2 slip and can improve the quality of the feed to the sulfur recovery unit (Bullin, 2003).

7.2.2.2.3 Flash Tank The rich amine solution from the absorber enters a flah tank, allowing the lightest of the hydrocarbons to flash. A small percentage of acid gases will also flash when the pressure is reduced. The heavier hydrocarbons remain as a liquid, but separate from the aqueous amine, forming a separate liquid layer. Because the hydrocarbons have a lower density than the aqueous amine, they form the upper liquid layer and can be skimmed off the top. Therefore, a provision should be made to remove these liquid hydrocarbons. Typically the flash tanks are designed for 2 to 3 minutes of retention time for the amine solution while operating half-full (Arnold and Stewart, 1999).

7.2.2.2.4 Amine Reboiler The amine reboiler provides the heat input to an amine stripper, which reverses the chemical reactions and drives off the acid gases. Two different reboiler designs are used in amine plants: thermosiphon and kettle. Thermosiphon reboilers return the heated amine solution and steam to the regenerator tower by the same pipe. Kettle reboilers return the heated amine solution and steam to the regenerator tower in different pipes.

The reboiler heat duty includes the (1) sensible heat required to raise the temperatures of the rich amine feed, the reflux, and the makeup water to the temperature of the reboiler, (2) heat of reaction to break chemical bonds between the acid gas molecules and the amine, and (3) heat of vaporization of water to produce a stripping vapor of steam. The heat duty and transfer area of the amine reboiler can be determined as follows (Jones and Perry, 1973):

$$H_{\mathrm{R}} = 432000 \times Q \tag{7-18}$$

$$A = 11.30 \times Q \tag{7-19}$$

where is Q is amine circulation flow rate, gal/min; H_{R} is heat duty of amine reboiler, Btu/min; and A is heat transfer area of reboiler, ft^2.

The reboiler duty should be maintained as low as possible, but must be adequate to regenerate the amine solution sufficiently to meet the sweet gas requirements and to ensure that the CO_2 loadings in the reboiler do not cause excessive corrosion. Higher reboiler duties do not reduce circulation rates to any degree and just consume energy (Bullin, 2003).

Reboiler temperature is dependent on solution concentration, flare/vent line back pressure, and/or residual CO_2 content required. It is good practice to operate the reboiler at as low a temperature as possible. The normal operating range for reboiler temperature is 225 to 260°F for MEA and 230 to 250°F for DEA systems.

7.2.2.2.5 Amine Stripper Amine strippers use heat and steam to reverse the chemical reactions with CO_2 and H_2S. The steam acts as a stripping gas to remove the CO_2 and H_2S from the liquid solution and to carry these gases to the overhead. Like the absorber, the stripper is either a tray or a packed column with approximately 20 trays or the equivalent height in packing. To minimize amine vaporization loss, there may be a water wash section at the top of the column with an additional four to six trays (Kohl and Riesenfeld, 1985).

The rich amine feed is introduced on the third or fourth tray from the top. The lean amine is removed at the bottom of the stripper and acid gases are removed from the top.

Liquid flow rates are greatest near the bottom tray of the tower where the liquid from the bottom tray must provide the lean amine flow rate from the tower plus enough water to provide the steam generated by the reboiler. The lean amine circulation rate is known, and from the reboiler duty, pressure, and temperature, the amount of steam generated and thus the amount of water can be calculated.

The vapor flow rate within the tower must be studied at both ends of the stripper. The higher of these vapor rates should be used to size the tower for vapor. At the bottom of the tower the vapor rate equals the amount of steam generated in the reboiler. Near the top of the tower, the vapor rate equals the steam rate overhead plus the acid gas rate. The steam overhead can be calculated from the steam generated in the reboiler by subtracting the amount of steam condensed by raising the lean amine from its inlet temperature to the reboiler temperature and the amount of steam condensed by vaporizing the acid gases.

7.2.2.2.6 Amine Condensers Amine-stripper condensers are typically overhead air-cooled and fin-fan exchangers. The amine reflux condenser is required to cool the overhead gases and condense the overhead steam to water. The inlet temperature to the cooler can be found using the partial pressure of the overhead steam to determine the temperature from steam tables. The cooler outlet temperature is typically 130 to 145°F depending on the ambient temperature. The heat duty and transfer area of an amine reflux condenser can be determined as follows (Jones and Perry, 1973):

$$H_C = 18 \times 10^5 \times Q \tag{7-20}$$

$$A = 5.20 \times Q \tag{7-21}$$

where H_C is heat duty of amine reflux condenser, Btu/min; and Q is amine circulation flow rate, gal/min.

The reflux accumulator is a two-phase separator used to separate the acid gases from the condensed water. The water is accumulated and pumped back to the top of the stripper as reflux.

7.2.2.2.7 Lean/Rich Amine Exchanger The lean/rich amine heat exchanger preheats the rich amine solution and reduces the duty of the reboiler. It also cools the lean amine and reduces the duty of the aerial cooler. Both shell and tube and plate and frame exchangers are used. The rich solution is passed through the tubes, which are usually made of stainless steel, with a low inlet velocity (2 to 3.5 ft/sec) to minimize corrosion. Typically the temperature change for both streams is 70 to 100°F and the pressure drops 2 to 5 psi. Using two or more exchangers in series is common practice to accommodate a temperature cross; for example, the exit rich amine stream is hotter than the exit lean amine stream (Manning and Thompson, 1991).

The heat duty and transfer area of a shell and tube type amine exchanger can be determined as follows (Jones and Perry, 1973):

$$H_{\mathrm{E}} = 27 \times 105 \times Q \tag{7-22}$$

$$A = 11.25 \times Q \tag{7-23}$$

where is H_{E} is heat duty of amine heat exchanger, Btu/min; A is Heat transfer area, ft^2; and Q is amine circulation flow rate, gal/min.

7.2.2.2.8 Amine Cooler The amine cooler is typically an air-cooled, fin-fan cooler, which lowers the lean amine temperature before it enters the absorber. The lean amine entering the absorber should be approximately 10°F warmer than the sour gas entering the absorber. Lower amine temperatures may cause the gas to cool in the absorber and thus condense hydrocarbon liquids. Higher temperatures would increase the amine vapor pressure and thus increase amine losses to the gas. The duty for the cooler can be calculated from the lean amine flow rate, with the lean amine temperature leaving the rich/lean exchanger and the sour gas inlet temperature.

7.2.2.2.9 Amine Reclaimer Due to side reactions and/or degradation, a variety of contaminants will begin to accumulate in an amine system. The method of removing these using an amine reclaimer depends on the amine involved. When MEA is used in the presence of COS and CS_2, they react to form heat-stable salts. Therefore, MEA systems usually include a reclaimer. The reclaimer is a kettle-type reboiler operating on a small side stream of lean amine solution. The temperature in the reclaimer is maintained such that the water and MEA boil to the overhead and are piped back to the stripper. The heat-stable salts remain in the reclaimer until the reclaimer is full. Then the reclaimer is shut in and dumped to a waste disposal. Thus, the impurities[2] are removed but the MEA bonded to the salts is also lost.

Diethanolamine (DEA) has a higher boiling temperature than monoethanolamines, requiring other methods of reclaiming such as vacuum distillation in order to prevent thermal degradation of the amine. Moreover, diethanolamine has a slow degradation rate. Consequently, in most cases it is not practical, economical, or necessary to reclaim DEA solutions. Solution purification is maintained by mechanical and carbon filtration and by caustic or soda ash addition to the system to neutralize the heat-stable amine salts.

Moisture may be removed from hydrocarbon gases at the same time as hydrogen sulfide is removed. Moisture removal is necessary to prevent harm to anhydrous catalysts and to prevent the formation of hydrocarbon hydrates (e.g., C_3H_8 $18H_2O$) at low temperatures. A widely used dehydration and desulfhurization process is the glycol/amine process, in which the treatment solution is a mixture of ethanolamine and a large amount of glycol. The mixture is circulated through an absorber and a reactivator in the same way as ethanolamine is circulated in the Girbotol process. The glycol absorbs moisture from the hydrocarbon gas passing up the absorber; the ethanolamine absorbs hydrogen sulfide and carbon dioxide. The treated gas leaves the top of the absorber; the spent ethanolamine glycol mixture enters the reactivator tower, where heat drives off the absorbed acid gases and water.

7.2.3 Carbonate Washing and Water Washing

Carbonate washing is a mild alkali process for emission control by the removal of acid gases (such as carbon dioxide and hydrogen sulfide) from gas streams (Speight, 1993) and uses the principle that the rate

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[2] Typical contaminants removed from the MEA solution by the reclaimer are the degradation products, 1-(2-hydroxyethyl) imidazolidone-2 and N-(2-hydroxyethyl) ethylenediamine, and nonvolatiles, such as inorganic ions, iron sulfide, high boiling hydrocarbons, and heat-stable salts.

of absorption of carbon dioxide by potassium carbonate increases with temperature. It has been demonstrated that the process works best near the temperature of reversibility of the reactions:

$$K_2CO_3 + CO_2 + H_2O \rightarrow 2KHCO_3 \tag{7-24}$$

$$K_2CO_3 + H_2S \rightarrow KHS + KHCO_3 \tag{7-25}$$

Water washing, in terms of the outcome, is analogous to washing with potassium carbonate (Kohl and Riesenfeld, 1985), and it is also possible to carry out the desorption step by pressure reduction. The absorption is purely physical and there is also a relatively high absorption of hydrocarbons, which are liberated at the same time as the acid gases.

7.2.4 Methanol Based Processes

Methanol is probably one of the most versatile solvents in the natural gas processing industry. Historically, methanol was the first commercial organic physical solvent and has been used for hydrate inhibition, dehydration, gas sweetening, and liquids recovery (Kohl and Nielsen, 1997; Font-Freide, 2004). Most of these applications involve low temperature where the physical properties of methanol are advantageous compared with other solvents that exhibit high viscosity problems or even solids formation. Operation at low temperatures tends to suppress the most significant disadvantage of methanol, high solvent loss. Furthermore, methanol is relatively inexpensive and easy to produce, making the solvent a very attractive alternate for gas processing applications.

Methanol has favorable physical properties relative to other solvents except for vapor pressure. The benefits of the low viscosity of methanol at at low temperature are manifested in the pressure drop improvement in the cold box of injection facilities and improved heat transfer. Methanol has a much lower surface tension relative to the other solvents. High surface tension tends to promote foaming problems in contactors. Methanol processes are probably not susceptible to foaming. However, the primary drawback of methanol is the high vapor pressure, which is several times greater than that of glycols or amines. To minimize methanol losses and enhance water and acid gas absorption, the absorber or separator temperatures are usually less than −20°F.

The high vapor pressure of methanol may initially appears to be a significant drawback because of high solvent losses. However, the high vapor pressure also has significant advantages. Although often not considered, lack of thorough mixing of the gas and solvent can pose significant problems. Because of the high vapor pressure, methanol is completely mixed in the gas stream before the cold box. Glycols, because they do not completely vaporize, may require special nozzles and nozzle placement in the cold box to prevent freeze-up. Solvent carryover to other downstream processes may also represent a significant problem. Because methanol is more volatile than glycols, amines, and other physical solvents, including lean oil, methanol is usually rejected in the regeneration step of these downstream processes. The stripper concentrates the methanol in the overhead condenser where it can be removed and further purified. Unfortunately, if glycols are carried over to amine units, the glycol becomes concentrated in the solution and potentially starts to degrade and possibly dilute the amine solution .

The use of methanol has been exploited further in the development of the Rectisol process either alone or as toluene–methanol mixtures are used to more selectively remove H_2S and slip CO_2 to the overhead product (Ranke and Mohr, 1985). Toluene has an additional advantage insofar as carbonyl sulfide is more soluble in toluene than in methanol. The Rectisol process was developed primarily to remove both carbon dioxide and hydrogen sulfide (along with other sulfur-containing species) from gas streams resulting from the partial oxidation of coal, oil, and petroleum residua. The ability of methanol to absorb these unwanted components made it the natural solvent of choice. Unfortunately, at cold temperatures, methanol also has a high affinity for hydrocarbon constituents of the gas streams. For example, propane is more soluble in methanol than carbon dioxide. There are two versions of the Rectisol process (Hoochgesand, 1970): the two stage and the once through. The first step of the two-stage process is desulfurization before shift conversion; the concentrations of hydrogen sulfide and carbon dioxide are about 1 and 5 vol%, respectively. Regeneration of the methanol following the desulfurization of the feed gas produces high sulfur feed for sulfur recovery. The once-through process is only applicable for high-pressure partial oxidation products. The once-through process is also applicable when the hydrogen sulfide to carbon dioxide content is unfavorable, in the neighborhood of 1:50 (Esteban *et al*., 2000).

The Institute of the Francais Petrole (IFP) has developed a process using methanol that has the simultaneous capability to dehydrate, to remove acid gas, and to control the hydrocarbon dew point (Rojey and Larue, 1988; Rojey *et al.,* 1990). The IFPEXOL-1 is used for water removal and hydrocarbon dew point control; the IFPEXOL-2 process is used for acid gas removal. The novel concept behind the IFPEXOL-l process is to use a portion of the water-saturated inlet feed to recover the methanol from the aqueous portion of the low temperature separator. That approach has solved a major problem with methanol injection in large facilities, methanol recovery via distillation. Beyond that very simple discovery, the cold section of the process is remarkably similar to a basic methanol injection process. Modifications to the process include water washing the hydrocarbon liquid from the low temperature separator to enhance methanol recovery. The IFPEXOL-2 process for acid gas removal is very similar to an amine-type process except for the operating temperatures. The absorber operates below −20°F to minimize methanol losses, and the regenerator operates at about 90 psi. Cooling is required on the regenerator condenser to recover the methanol. This process usually follows the IFPEXOL-1 process so excessive hydrocarbon absorption is not as great a problem (Minkkinen and Jonchere, 1997).

7.2.5 Other Processes

The process using *potassium phosphate* is known as phosphate desulfhurization and is used in the same way as the Girbotol process to remove acid gases from liquid hydrocarbons as well as from gas streams. The treatment solution is a water solution of tripotassium phosphate (K_3PO_4), which is circulated through an absorber tower and a reactivator tower in much the same way as the ethanolamine is circulated in the Girbotol process; the solution is regenerated thermally.

Other processes include the *Alkazid process* (Figure 7-3), which removes hydrogen sulfide and carbon dioxide using concentrated aqueous solutions of amino acids.

The hot potassium carbonate process (Figure 7-4) decreases the acid content of natural and refinery gas from as much as 50% to as low as 0.5% and operates in a unit similar to that used for amine treating.

The *Giammarco-Vetrocoke* process is used for hydrogen sulfide and/or carbon dioxide removal (Figure 7-5). In the hydrogen sulfide removal section, the reagent consists of sodium or potassium carbonates containing a mixture of arsenites and arsenates; the carbon dioxide removal section utilizes hot aqueous alkali carbonate solution activated by arsenic trioxide or selenous acid or tellurous acid.

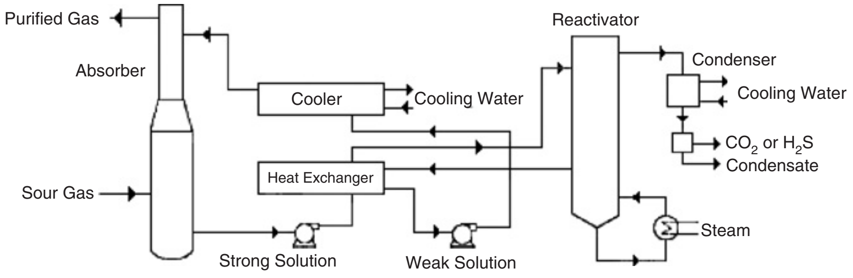

Figure 7-3. The Alkazid process flow diagram (Speight, 1993).

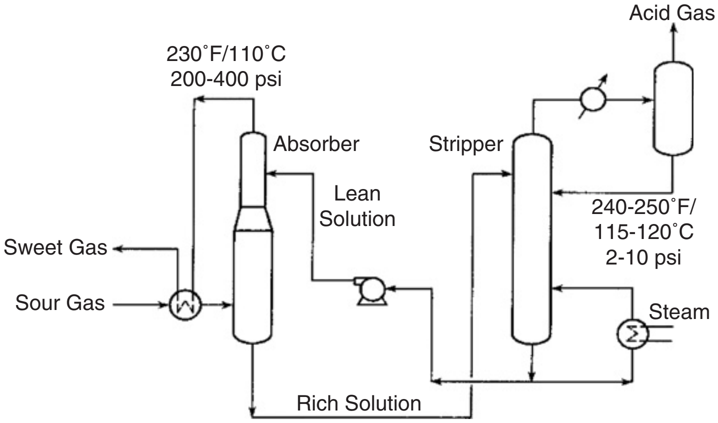

Figure 7-4. The hot potassium carbonate process flow diagram (Speight, 1993).

Molecular sieves are highly selective for the removal of hydrogen sulfide (as well as other sulfur compounds) from gas streams and over a continuously high absorption efficiency. They are also an effective means of water removal and thus offer a process for the simultaneous sdehydration and desulfhurization of gas. Gas that has an excessively high water content may require upstream dehydration, however. The molecular sieve process (Figure 7-6) is similar to the iron oxide process. Regeneration of the bed is achieved by passing heated clean gas over the bed.

As the temperature of the bed increases, it releases the adsorbed hydrogen sulfide into the regeneration gas stream. The sour effluent regeneration gas is sent to a flare stack, and up to 2% of the gas seated can be lost in the regeneration process. A portion of the natural gas may also be lost by the

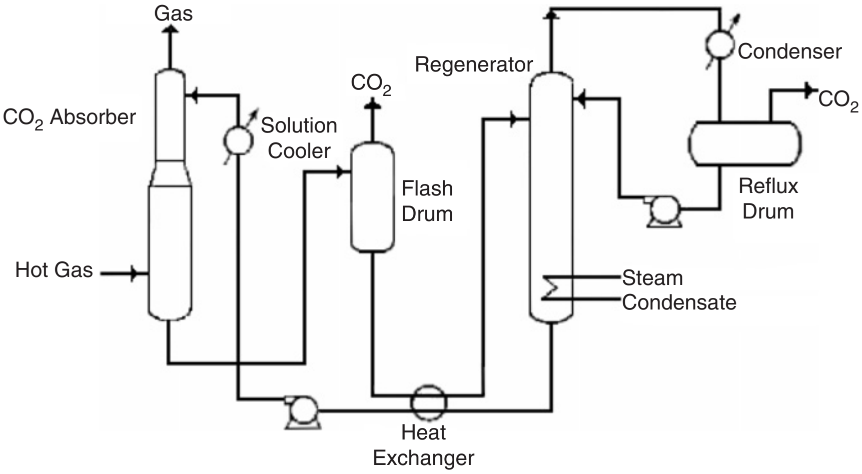

Figure 7-5. The Giammarco-Vetrocke process flow diagram (Speight, 1993).

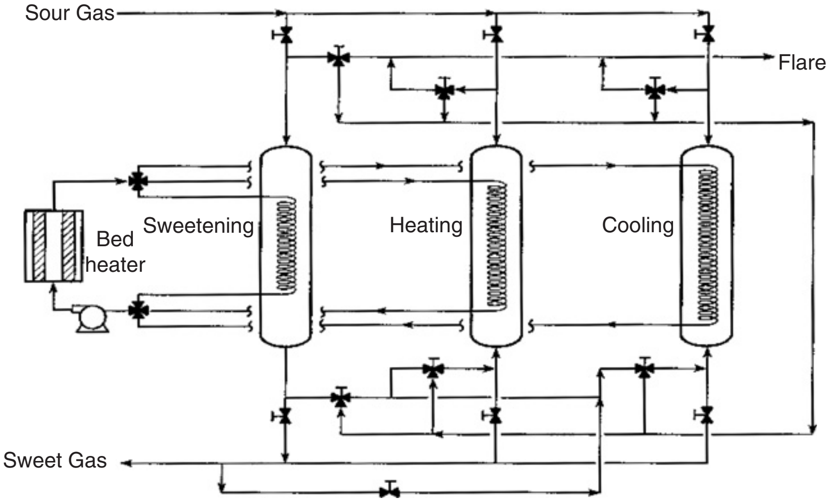

Figure 7-6. The molecular sieve process flow diagram (Speight, 1993).

adsorption of hydrocarbon components by the sieve. In this process, unsaturated hydrocarbon components, such as olefins and aromatics, tend to be strongly adsorbed by the molecular sieves. The molecular sieves are susceptible to poisoning by such chemicals as glycols and require thorough gas cleaning methods before the adsorption step. Alternatively, the sieve can be offered some degree of protection by the use of *guard beds* in which a less expensive catalyst is placed in the gas stream before contact of the gas with the sieve, thereby protecting the catalyst from poisoning. This concept is analogous to the use of guard beds or attrition catalysts in the petroleum industry (Speight, 1993).

Until recently, the use of *membranes* for gas separation has been limited to carbon dioxide removal (Alderton, 1993). Improvements in membrane technology have now made membranes competitive in other applications in the natural gas area. New membrane materials and configurations exhibit superior performance and offer improved stability against contaminants found in natural gas. The new membranes are targeted at three separations: nitrogen, carbon dioxide/hydrogen sulfide, and natural gas liquids (U.S. Patent 6,425,267, 2002). The process uses a two-step membrane system design; the methane-selective membranes do not need to be operated at low temperatures, and capital and operating costs are within economically acceptable limits.

New membranes have been developed (Lokhandwala and Jacobs, 2000) for the gas industry. For example, the membranes allow permeation of condensable vapors, such as C_{3+} hydrocarbons, aromatics, and water vapor, while rejecting the noncondensable gases, such as methane, ethane, nitrogen, and hydrogen. Since the early 1990s, more than 50 systems have been installed in the chemical process industry worldwide. The main applications are nitrogen removal, recovery of natural gas liquids, and dew point control for associated natural gas and fuel gas conditioning for gas turbines and engines (Hall and Lokhandwala, 2004).

In another process (U.S. Patent 6,053,965, 2000), a membrane-based process for upgrading natural gas that contains C_{3+} hydrocarbons and/or acid gas is described. The conditioned natural gas can be used as fuel for gas-powered equipment, including compressors, in the gas field or the processing plant. Optionally, the process can be used to produce natural gas liquids.

7.2.6 Process Selection

Each of the previous treating processes has advantages relative to the others for certain applications; therefore, in selection of the appropriate process, the following facts should be considered (Morgan, 1994; GPSA, 1998).

- Air pollution regulations regarding sulfur compound disposal and/or Tail Gas Clean Up (TGCU) requirements

- Type and concentration of impurities in the sour gas
- Specifications for the residue gas
- Specifications for the acid gas
- Temperature and pressure at which the sour gas is available and at which the sweet gas must be delivered
- Volume of gas to be precessed
- Hydrocarbon composition of the gas
- Selectivity required for acid gas removal
- Capital cost and operating cost
- Royalty cost for process
- Liquid product specifications
- Disposal of by-products considered hazardous chemicals

Decisions in selecting a gas treating process can often be simplified by gas composition and operating conditions. High partial pressures (50 psia) of acid gases enhance the probability of using a physical solvent. The presence of significant quantities of heavy hydrocarbons in the feed discourages using physical solvents. Low partial pressures of acid gases and low outlet specifications generally require the use of amines for adequate treating. Process selection is not easy and a number of variables must be weighed prior to making a process selection. After preliminary assessment, a study of relevant alternatives is usually required.

In general, batch and amine processes are used for over 90% of all onshore wellhead applications. Amine processes are preferred, because of the lower operating cost, when the chemical cost for this process is prohibitive, and justifies the higher equipment cost. The key determinant is the sulfur content of the feed gas. When the sulfur content is below 20-pound sulfur per day, batch processes are more economical, and when the sulfur content is over 100-pound sulfur per day amine solutions are preferred (Manning and Thompson, 1991).

7.3 SULFUR RECOVERY PROCESSES

The side stream from acid gas treating units consists mainly of hydrogen sulfide/or carbon dioxide. Carbon dioxide is usually vented to the atmosphere but sometimes is recovered for CO_2 floods. Hydrogen sulfide could be routed to an incinerator or flare, which would convert the H_2S to SO_2. The release of H_2S to the atmosphere may be limited by environmental regulations. There are many specific restrictions on these limits, and the allowable limits are revised periodically. In any case, environmental regulations severely restrict the amount of H_2S that can be vented or flared in the regeneration cycle.

Most sulfur recovery processes use chemical reactions to oxidize H_2S and produce elemental sulfur. These processes are generally based either on the reaction of H_2S and O_2 or H_2S and SO_2. Both reactions yield water and elemental sulfur. These processes are licensed and involve specialized catalysts and/or solvents. These processes can be used directly on the produced gas stream. Where large flow rates are encountered, it is more common to contact the produced gas stream with a chemical or physical solvent and use a direct conversion process on the acid gas liberated in the regeneration step.

There are two common methods of sulfur recovery: liquid redox and Claus sulfur recovery processes.

Liquid redox sulfur recovery processes are liquid-phase oxidation processes that use a dilute aqueous solution of iron or vanadium to remove H_2S selectively by chemical absorption from sour gas streams. These processes can be used on relatively small or dilute H_2S stream to recover sulfur from the acid gas stream or, in some cases, can be used in place of an acid gas removal process. The mildly alkaline lean liquid scrubs the H_2S from the inlet gas stream, and the catalyst oxidizes the H_2S to elemental sulfur. The reduced catalyst is regenerated by contact with air in the oxidizer(s). Sulfur is removed from the solution by flotation or settling, depending on the process.

The Claus sulfur recovery process (Figure 7-7) is the most widely used technology for recovering elemental sulfur from sour gas. The Claus process is used to recover sulfur from the amine regenerator vent gas stream in plants where large quantities of sulfur are present. However, this process is used to treat gas streams with a maximum H_2S content of 15%. The chemistry of the units involves partial oxidation of hydrogen sulfide to sulfur dioxide and the catalytically promoted reaction of H_2S and SO_2 to produce elemental sulfur. The reactions are staged and are as follows.

$$H_2S + 3/2O_2 \rightarrow SO_2 + H_2O; \quad \text{thermal stage} \tag{7-26}$$

$$SO_2 + 2H_2S \rightarrow 3S + 2H_2O; \quad \text{thermal and catalytic stage} \tag{7-27}$$

The first stage of the process converts hydrogen sulfide to sulfur dioxide and to sulfur by burning the acid gas stream with air in the measurement across the tower. The tower is more susceptible to flooding at lower pressures. If an overhead compressor is employed to boost the tower overhead up to fuel system pressure, then the pressure can be lowered subject to the maximum speed of the compressor driver, rod load limitations for a reciprocating compressor, or surge considerations for a centrifugal compressor.

6.2.2.2 Design Considerations of Stabilization Column

In most cases of lease operation, the stabilization column will operate as a nonrefluxed tower. This type of operation is simpler but less efficient than the refluxed tower operation. Because the nonrefluxed tower requires no external cooling source, it is particularly applicable to remote locations. A condensate stabilization column with reflux will recover more intermediate components from the gas than a cold-feed stabilizer. However, it requires more equipment to purchase, install, and operate. This additional cost must be justified by the net benefit of the incremental liquid recovery, less the cost of natural gas shrinkage and loss of heating value, over that obtained from a cold-feed stabilizer. When a condenser is used in a stabilization column, it will always be a partial condenser because of the quantities of methane and ethane that must be removed from the tower feed.

The stabilization tower pressure depends on the amount of liquid to be stabilized and whether it is sweet or sour. For sweet stabilization, the pressure should be as high as possible to minimize overhead vapor recompression, as this gas is remixed with the separator vapor. This also tends to decrease the cost of reflux cooling, if it is used. However, relative volatility of the components also decreases with pressure and, as stated previously, driving H_2S overhead requires a relatively low pressure.

Figure 6-3 shows the maximum recommended feed temperature to a stabilizer as a function of operating pressure of the stabilizer. An exception to this may be the case where either sour or small quantities of liquid are being handled and where first cost therefore is very critical. In these cases, many times a 40- to 70- psia working pressure for a nonrefluxed tower may show an economic advantage (Campbell, 1992).

In some cases the reboiler for the stabilizer will be an indirect salt bath heater or a steam-fired heat exchanger. Figure 6-4 shows suggested bottom (reboiler) temperatures for producing a specified Reid vapor pressure product. In fact, the temperature used on the bottom is limited by the TGCU process is therefore of some importance to operators whereas the choice is made complicated by the different performance levels and life cycle costs. A selection of the most popular TGCU processes based on their sulfur recovery efficiency and, most importantly, in terms of the recovery cost per tonne of SO_2 has been compared by Gall and Gadelle (2003).

REFERENCES

Alderton, P.D., "Natural Gas Treating Using Membranes," paper presented at the 2nd GPA Technical Meeting, GCC Chapter, Bahrain (Oct.27, 1993).

Anerousis, J.P., and Whitman, S.K., "An Updated Examination of Gas Sweetening by the Iron Sponge Process," SPE 13280, paper presented at the SPE Annual Technical Conference & Exhibition, Houston, TX (Sept. 1984).

Arnold, K., and Stewart, M., "Surface Production Operations: Vol. 2: Design of Gas-Handling systems and Facilities," 2nd Ed. Gulf Professional Publishing, Houston, TX (1999).

Bhatia, K., and Allford, K.T., One-step process takes H_2S from gas stream. *Oil Gas J.* **84**(42), 44 (1986).

Blanc, C., Elgue, J., and Lallemand, F., "MDEA Process Selects H2S," Hydrocarbon Processing (August 1981).

Bullin, J.A., "Why Not Optimize Your Amine Sweetening Unit." Paper presented at GPA Europe Annual Conference, Heidelberg, Germany (Sept. 25–27, 2003).

Bullin, J.A., Polasek, J.C., and Donnelly, S.T., "The Use of MDEA and Mixtures of Amines for Bulk CO_2 Removal." Paper presented at the 69th GPA Annual Convention, Tulsa, OK (1990).

Burnes, E.E., and Bhatia, K., "Process for Removing Hydrogen Sulfide from Gas Mixtures," U.S. Patent, 4, 515, 759 (May 1985).

Dobbs, J.B., "One Step Process." Paper presented at the Laurance Reid Gas Conditioning Conference, Norman, OK (March 1986).

Duckworth, G.L., and Geddes, J.H., Natural gas desulfurization by the iron sponge process. *Oil Gas J.* **63**(37), 94–96 (1965).

Esteban, A., Hernandez, V., and Lunsford, K., "Exploit the Benefits of Methanol." Paper presented at the 79th GPA Annual Convention. Atlanta, GA (March 2000).

Font Freide, JHM, I.J., "Methanol: A Gas Processing Option." Paper presented at the GPA Europe Meeting, Provence, France (Sept. 22–24, 2004).

Fox, I., "Process for Scavenging Hydrogen Sulfide from Hydrocarbon Gases." U.S. Patent, 4, 246, 274 (Jan. 1981).

Gall, A.L., and Gadelle, D., "Technical and Commercial Evaluation of Processes for Claus Tail Gas Treatment." Paper presented at the GPA Europe Technical Meeting, Paris, France (Feb. 2003).

Gollmar, H.A., Removal of sulfur compounds from coal gas, 947–1007 *In* "Chemistry of Coal Utilization", (H.H. Lowry, (ed). Wiley, New York (1945).

GPSA Engineering Data Book, 11th Ed. Gas Processors Suppliers Association, Tulsa, OK (1998).

Hall, P., and Lokhandwala, K.A., "Advances in Membrane Materials Provide New Gas Processing Solutions." Paper presented at the GPA Annual Convention, New Orleans, LA (March 2004).

Hohlfeld, R.W., New H_2S scrubber cleans small or large gas streams. *Oil Gas J.* 129–130 (Oct. 15, 1979).

Hoochgesand, G., Rectisol and purisol. *Ind. Eng. Chem.* **62**(7), 37–43 (1970).

Jenkins, J.L., and Haws, R., Understanding gas treating fundamentals. *Petr. Technol. Q*. 61–71 (Jan. 2002).

Jones, V.W., and Perry, C.R., "Fundamentals of Gas Treating." Paper presented at the Laurance Reid Gas Conditioning Conference, Norman, OK (1973).

Kattner, J.E., Samuels, A., and Wendt, R.P., Iron oxide slurry process for removing hydrogen sulfide. *J. Petr. Technol.* **40**(9), 1237 (1988).

Khan, M., and Manning, W.P., "Practical Designs for Amine Plants," Petroenergy Workshop, Houston, TX (Sept. 1985).

Kohl, A.L., and Nielsen, R.B., "Gas Purification." Gulf Publishing Company, Houston, TX (1997).

Kohl, A. L., and Riesenfeld, F.C., "Gas Purification," 4th Ed. Gulf Publishing Company, Houston, TX (1985).

Kutsher, G.S., and Smith, G.A., "CO_2 Removal and Recovery Pegasol: A Solvent Process." Paper presented at the Laurence Reid Gas Conditioning Conference, Norman, OK (1966).

Lallemand, F., and Minkkinen, A., "High Sour Gas Processing in an Ever-Greener World." Paper presented at the 9th GPA-GCC Chapter Technical Conference, Abu Dhabi (May 23, 2001).

Lokhandwala, K.A., and Jaqcobs, M.L., "New Membrane Application in Gas Processing." Paper presented at the GPA Annual Convention, Atlanta, GA (March, 2000).

Maddox, R.N., "Gas and Liquid Sweetening," 2nd Ed. Campbell Petroleum Series, Norman, OK (1974).

Manning, F.S., and Thompson, R.E., "Oil Field Processing of Petroleum. Vol. 1: Natural Gas." Pennwell Publishing Company, Tulsa, OK (1991).

Manning, W.P., Chemsweet, a new process for sweetening low-value sour gas. *Oil Gas J*. **77**(42), 122–124 (1979).

Minkkinen, A., and Jonchere, J.P., "Methanol Simplifies Gas Processing." Paper presented at the 5th GPA-GCC Chapter Technical Conference, Bahrain (May 1997).

Morgan, D.J. "Selection Criteria for Gas Sweetening." Paper presented at the GPA Technical Meeting, GCC Chapter, Bahrain (Nov.30, 1994).

Newman, S.A. (ed.), "Acid and Sour Gas Treating Processes." Gulf Publishing Company, Houston, TX (1985).

Polasek, J., and Bullin, J.A., "Selecting Amines for Sweetening Units." Paper presented at the GPA Regional Meeting, Tulsa, OK (Sept. 1994).

Polasek, J.C., Iglesias-Silva, G.A., and Bullin, J.A., "Using Mixed Amine Solutions for Gas Sweetening." Paper presented at the 71st GPA Annual Convention, Tulsa, OK (1992).

Ranke, G., and Mohr, V.H., The rectisol wash new developments in acid gas removal from synthesis gas." *In* "Acid and Sour Gas Treating Processes" (S. A. Newman, ed.). Gulf Publishing Company, Houston, TX (1985).

Rojey, A., and Larue, J., "Integrated Process for the Treatment of a Methane-Containing Wet Gas in Order to Remove Water Therefom." Assigned to the Institut Français du Petrole (IFP), U.S. Patent 4,775,395 (Oct.4, 1988).

Rojey, A., Procci, A., and Larue, J., "Process and Apparatus for Dehydration, De-acidification, and Separation of Condensate from a Natural Gas", Assigned to the Institut Français du Petrole (IFP), U.S. Patent 4,979,966 (1990).

Samuels, A., "Gas Sweetener Associates", Technical Manual, 3-88, Metairie, LA (1988).

Schaack, J.P., and Chan, F., Caustic-based process remains attractive", *Oil Gas J*. **5**, 81–82 (1989).

Souders, M., and Brown, G.G., Fundamental design of absorbing and stripping columns for complex vapors.” *Ind. Eng. Chem.* **24**, 519 (1932).

Speight, J.G., “Gas Processing: Environmental Aspects and Methods.” Butterworth Heinemann, Oxford, England (1993).

U.S. Patent 6,053,965, “Fuel Gas Conditioning Process” (April 25, 2000).

U.S. Patent 6,425,267, “Two-Step Process for Nitrogen Removal from Natural Gas” (July 30, 2002).

Zapffe, F., Iron sponge process removes mercaptans. *Oil Gas J.* **61** (33), 103–104 (1963).

CHAPTER 8

NATURAL GAS COMPRESSION

8.1 INTRODUCTION

“Compression” is used in all aspects of the natural gas industry, including gas lift, reinjection of gas for pressure maintenance, gas gathering, gas processing operations (circulation of gas through the process or system), transmission and distribution systems, and reducing the gas volume for shipment by tankers or for storage. In recent years, there has been a trend toward increasing pipeline-operating pressures. The benefits of operating at higher pressures include the ability to transmit larger volumes of gas through a given size of pipeline, lower transmission losses due to friction, and the capability to transmit gas over long distances without additional boosting stations. In gas transmission, two basic types of compressors are used: reciprocating and centrifugal compressors. Reciprocating compressors are usually driven by either electric motors or gas engines, whereas centrifugal compressors use gas turbines or electric motors as drivers. The key variables for equipment selections are life cycle cost, capital cost, maintenance costs, including overhaul and spare parts, fuel, or energy costs. The units level of utilization, as well as demand fluctuations, plays an important role. While both gas engines and gas turbines can use pipeline gas as a fuel, an electric motor has to rely on the availability of electric power. Due to the number of variables involved, the task of choosing the optimum driver can be quite involved, and a comparison between the different types of drivers should be done before a final selection is made (Kurz *et al*., 2003). An economic feasibility study is of fundamental importance to determine the best selection for the economic life of a project. Furthermore, it must be decided whether the compression task should be divided into multiple compressor trains, operating in series or in parallel.

This chapter presents a brief overview of the two major types of compressors and a procedure for calculation of the required compression power, as well as additional and useful considerations for the design of compressor stations. All performance calculations should be based on inlet flange to discharge flange conditions. This means that for centrifugal compressors, the conditions at the inlet flange into the compressor and the discharge flange of the compressor are used. For reciprocating compressors, this means that pressure losses at the cylinder valves, as well as pressure losses in pulsation dampeners, have to be included in the calculation. Additional losses for process equipment such as suction scrubbers or aftercoolers have to be accounted for separately.

8.2 RECIPROCATING COMPRESSORS

A reciprocating compressor is a positive displacement machine in which the compressing and displacing element is a piston moving linearly within a cylinder. The reciprocating compressor uses automatic spring-loaded valves that open when the proper differential pressure exists across the valve. Figure 8-1 describes the action of a reciprocating compressor using a theoretical pressure–volume (PV) diagram. In position A, the suction valve is open and gas will flow into the cylinder (from point 1 to point 2 on the PV diagram) until the end of the reverse stroke at point 2, which is the start of compression. At position B, the piston has traveled the full stroke within the cylinder and the cylinder is full of gas at suction pressure. Valves remain closed. The piston begins to move to the left, closing the suction valve. In moving from position B to position C, the piston moves toward the cylinder head, reducing the volume of gas with an accompanying rise in pressure. The PV diagram shows compression from point 2 to point 3. The piston continues to move to the end of the stroke (near the cylinder head) until the cylinder pressure is equal to the discharge pressure and the discharge valve opens (just beyond point 3). After the piston reaches point 4, the discharge valve will close, leaving the clearance space filled with gas at discharge pressure (moving from position C to position D). As the piston reverses its travel, the gas remaining within the cylinder expands (from point 4 to point 1) until it equals suction pressure and the piston is again in position A.

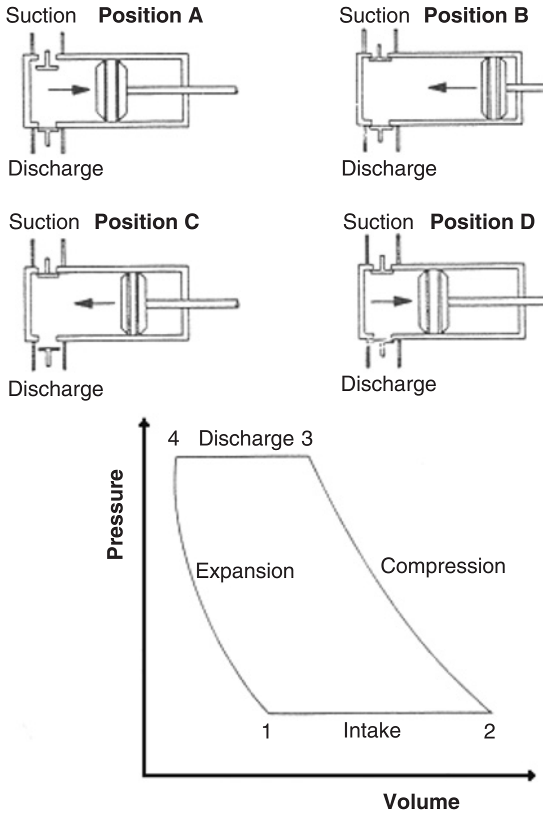

Figure 8-1. Reciprocating compressor compression cycle.

The flow to and from reciprocating compressors is subject to significant pressure fluctuations due to the reciprocating compression process. Therefore, pulsation dampeners have to be installed upstream and downstream of the compressor to avoid damages to other equipment. The pressure losses (several percent of the static flow pressure) of these dampeners have to be accounted for in the station design.

Reciprocating compressors are widely utilized in the gas processing industries because they are flexible in throughput and discharge pressure range. Reciprocating compressors are classified as either “high speed” or “slow speed.” Typically, high-speed compressors operate at speeds of 900 to 1200 rpm and slow-speed units at speeds of 200 to 600 rpm. High-speed units are normally “separable,” i.e., the compressor frame and driver are separated by a coupling or gearbox. For an "integral" unit, power cylinders are mounted on the same frame as the compressor cylinders, and power pistons are attached to the same drive shaft as the compressor cylinders. Low-speed units are typically integral in design.

8.3 CENTRIFUGAL COMPRESSORS

We want to introduce now the essential components of a centrifugal compressor that accomplish the tasks specified earlier (Figure 8-2). The gas entering the inlet nozzle of the compressor is guided to the inlet of the impeller. An impeller consists of a number of rotating vanes that impart mechanical energy to the gas. The gas will leave the impeller with an increased velocity and increased static pressure. In the diffuser, part of the velocity is converted into static pressure. Diffusers can be vaned, vaneless, or volute type. If the compressor has more than one impeller, the gas will be again brought in front of the next impeller through the return channel and the return vanes. If the compressor has only one impeller, or after the diffuser of the last impeller in a multistage compressor, the gas enters the discharge system. The discharge system can either make use

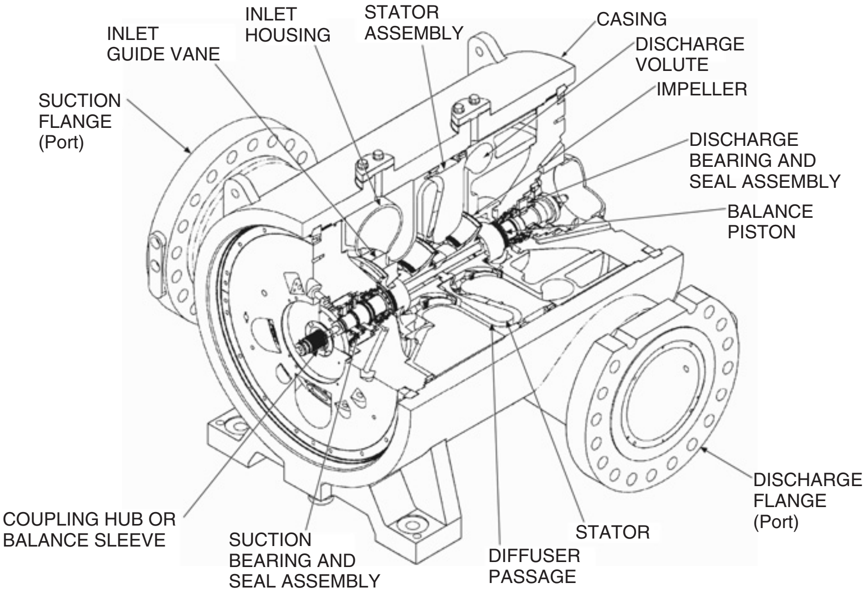

Figure 8-2. Typical centrifugal compressor cutaway (GPSA, 2004).

of a volute, which can further convert velocity into static pressure, or a simple cavity that collects the gas before it exits the compressor through the discharge flange.

The rotating part of the compressor consists of all the impellers. It runs on two radial bearings (on all modern compressors, these are hydrodynamic tilt pad bearings), while the axial thrust generated by the impellers is balanced by a balance piston, and the resulting force is balanced by a hydrodynamic tilt pad thrust bearing. To keep the gas from escaping at the shaft ends, dry gas seals are used. The entire assembly is contained in a casing (usually barrel type).

A compressor stage is defined as one impeller, with the subsequent diffuser and (if applicable) return channel. A compressor body may hold one or several (up to 8 or 10) stages. A compressor train may consist of one or multiple compressor bodies. It sometimes also includes a gearbox. Pipeline compressors are typically single body trains, with one or two stages.

The different working principles cause differences in the operating characteristics of the centrifugal compressors compared to those of the reciprocating unit. Centrifugal compressors are used in a wide variety of applications in chemical plants, refineries, onshore and offshore gas lift and gas injection applications, gas gathering, and in the transmission of natural gas. Centrifugal compressors can be used for outlet pressures as high as 10,000 psia, thus overlapping with reciprocating compressors over a portion of the flow rate/pressure domain. Centrifugal compressors are usually either turbine or electric motor driven. Typical operating speeds for centrifugal compressors in gas transmission applications are about 14,000 rpm for 5000-hp units and 8000 rpm for 20,000-hp units.

8.4 COMPARISON BETWEEN COMPRESSORS

Differences between reciprocating and centrifugal compressors are summarized as follow (GPSA, 2004).

Advantages of a reciprocating compressor over a centrifugal machine include

- Ideal for low volume flow and high-pressure ratios
- High efficiency at high-pressure ratios
- Relatively low capital cost in small units (less than 3000 hp)
- Less sensitive to changes in composition and density

Advantages of a centrifugal compressor over a reciprocating machine include

- Ideal for high volume flow and low head
- Simple construction with only one moving part
- High efficiency over normal operating range
- Low maintenance cost and high availability
- Greater volume capacity per unit of plot area
- No vibrations and pulsations generated

8.5 COMPRESSOR SELECTION

The design philosophy for choosing a compressor should include the following considerations.

- Good efficiency over a wide range of operating conditions
- Maximum flexibility of configuration
- Low maintenance cost
- Low life cycle cost
- Acceptable capital cost
- High availability

However, additional requirements and features will depend on each project and on specific experiences of the pipeline operator. In fact, compressor selection consists of the purchaser defining the operating parameters for which the machine will be designed. The “process design parameters” that specify a selection are as follows (Akhtar, 2002).

1. Flow rate
2. Gas composition
3. Inlet pressure and temperature
4. Outlet pressure
5. Train arrangement

    a. For centrifugal compressors: series, parallel, multiple bodies, multiple sections, intercooling, etc.
    b. For reciprocating compressors: number of cylinders, cooling, and flow control strategy

6. Number of units

In many cases, the decision whether to use a reciprocating compressor or a centrifugal compressor, as well as the type of driver, will already have been made based on operator strategy, emissions requirements, general life cycle cost assumptions, and so on. However, a hydraulic analysis should be made for each compressor selection to ensure the best choice. In fact, compressor selection can be made for an operating point that will be the most likely or most frequent operating point of the machine. Selections based on a single operating point have to be evaluated carefully to provide sufficient speed margin (typically 3 to 10%) and surge margin to cover other, potentially important situations. A compressor performance map (for centrifugal compressors, this would be preferably a head vs flow map) can be generated based on the selection and is used to evaluate the compressor for other operating conditions by determining the head and flow required for these other operating conditions. In many applications, multiple operating points are available, e.g., based on hydraulic pipeline studies or reservoir studies. Some of these points may be frequent operating points, while some may just occur during upset conditions. With this knowledge, the selection can be optimized for a desired target, such as lowest fuel consumption.

Selections can also be made based on a “rated” point, which defines the most onerous operating conditions (highest volumetric flow rate; lowest molecular weight; highest head or pressure ratio; highest inlet temperature). In this situation, however, the result may be an oversized machine that does not perform well at the usual operating conditions.

Once a selection is made, the manufacturer is able to provide parameters such as efficiency, speed, and power requirements and, based on this and the knowledge of the ambient conditions (prevailing temperatures, elevation), can size the drivers. At this point, the casing arrangement and the number of units necessary or desirable (flexibility requirements, growth scenarios and sparing considerations will play an important role in this decision) can be discussed.

8.6 THERMODYNAMICS OF GAS COMPRESSION

The task of gas compression is to bring gas from a certain suction pressure to a higher discharge pressure by means of mechanical work. The actual compression process is often compared to one of three ideal processes: isothermal, isentropic, and polytropic compression.

Isothermal compression occurs when the temperature is kept constant during the compression process. It is not adiabatic because the heat generated in the compression process has to be removed from the system.

The compression process is isentropic or adiabatic reversible if no heat is added to or removed from the gas during compression and the process is frictionless. With these assumptions, the entropy of the gas does not change during the compression process.

The polytropic compression process is like the isentropic cycle reversible, but it is not adiabatic. It can be described as an infinite number of isentropic steps, each interrupted by isobaric heat transfer. This heat addition guarantees that the process will yield the same discharge temperature as the real process.

It is important to clarify certain properties at this time and, in particular, find their connection to the first and second law of thermodynamics written for steady-state fluid flows. The first law (defining the conservation of energy) becomes

$$\left(h_2 + \frac{u_2^2}{2} + gz_2\right) - \left(h_1 + \frac{u_1^2}{2} + gz_1\right) = q_{12} + W_{t,12} \qquad \text{(8-1)}$$

where h is enthalpy, u is velocity, g is gravitational acceleration, z is elevation coordinate, q is heat, and W_t is work done by the compressor on the gas.

Neglecting the changes in potential energy (because the contribution due to changes in elevation is not significant for gas compressors), the energy balance equation for adiabatic processes ($q_{12} = 0$) can be written as

$$\left(h_2 + \frac{u_2^2}{2}\right) - \left(h_1 + \frac{u_1^2}{2}\right) = W_{t,12} \qquad \text{(8-2)}$$

$W_{t,12}$ is the amount of work[1] we have to apply to affect the change in enthalpy in the gas. The work $W_{t,12}$ is related to the required power, P,

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[1] Physically, there is no difference among work, head, and change in enthalpy. In systems with consistent units (such as the SI system), work, head, and enthalpy difference, have the same unit (e.g., kJ/kg in SI units). Only in inconsistent systems (such as US customary units) do we need to consider that the enthalpy difference (e.g., in Btu/lb_m) is related to head and work (e.g., in $\text{ft.lb}_\text{f}/\text{lb}_\text{m}$) by the mechanical equivalent of heat (e.g., in $\text{ft.lb}_\text{f}/\text{Btu}$).

by multiplying it with the mass flow.

$$P = \dot{m} W_{t,12} \tag{8-3}$$

Combining enthalpy and velocity into a total enthalpy ($h_{\mathrm{t}} = h + \frac{u^2}{2}$), power and total enthalpy difference are thus related by

$$P = \dot{m}(h_{t,2} - h_{t,1}) \tag{8-4}$$

If we can find a relationship that combines enthalpy with the pressure and temperature of a gas, we have found the necessary tools to describe the gas compression process. For a perfect gas, with constant heat capacity, the relationship among enthalpy, pressures, and temperatures is

$$\Delta h = C_{\mathrm{p}}(T_2 - T_1) \tag{8-5}$$

where T_1 is suction temperature, T_2 is discharge temperature, and C_{p} is heat capacity at constant pressure.

For an isentropic compression, the discharge temperature ($T_{2\mathrm{s}}$) is determined by the pressure ratio as

$$T_{2\mathrm{s}} = T_1 \left[\left(\frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] + T_1 \tag{8-6}$$

where $k = C_{\mathrm{p}}/C_{\mathrm{v}}$, p_1 is suction pressure, and p_2 is discharge pressure.

Note that the specific heat at constant pressure (C_{p}) and the specific heat at constant volume (C_{v}) are functions of temperature only for ideal gases and can be related together with $C_{\mathrm{p}} - C_{\mathrm{v}} = R$, where R is the universal gas constant. The isentropic exponent (k) for ideal gas mixtures can also be determined as

$$k = \frac{\sum y_{\mathrm{i}} C_{\mathrm{pi}}}{\left[\sum y_{\mathrm{i}} C_{\mathrm{pi}}\right] - R} \tag{8-7}$$

where C_{Pi} is the molar heat capacity of the individual component and y_{i} is the molar concentration of the component.

The heat capacities of real gases are a function of the pressure and temperature and thus may differ from the ideal gas case. For hand calculations, the ideal gas k is sufficiently accurate.

If the gas composition is not known, and the gas is made up of alkanes (such as methane and ethane) with no substantial quantities of contaminants and whose specific gravity (SG) does not exceed unity, the following empirical correlation can be used (Campbell, 1992).

$$k = 1.3 - 0.31(\mathrm{SG} - 0.55) \tag{8-8}$$

Combining Equations (8-5) and (8-6), the isentropic head (Δh_s) for the isentropic compression of a perfect gas can thus be determined as

$$\Delta h_s = C_p T_1 \left[\left(\frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] \tag{8-9}$$

For real gases [where k and C_p in Equation (8-9) become functions of temperature and pressure], the enthalpy of a gas, h, is calculated in a more complicated way using equations of state. These represent relationships that allow one to calculate the enthalpy of gas of known composition if any two of its pressure, its temperature, or its entropy are known.

We therefore can calculate the actual head for the compression by

$$\Delta h = h(p_2, T_2) - h(p_1, T_1) \tag{8-10}$$

and the isentropic head by

$$\Delta h_s = h(p_2, s_1) - h(p_1, T_1) \tag{8-11}$$

where entropy of the gas at suction condition (s_1) is

$$s_1 = s(p_1, T_1) \tag{8-12}$$

The relationships described can be seen easily in a Mollier diagram (Figure 8-3).

The performance quality of a compressor can be assessed by comparing the actual head (which relates directly to the amount of power we need to spend for the compression) with the head that the ideal, isentropic compression would require. This defines the isentropic efficiency (η_s) as

$$\eta_s = \frac{\Delta h_s}{\Delta h} \tag{8-13}$$

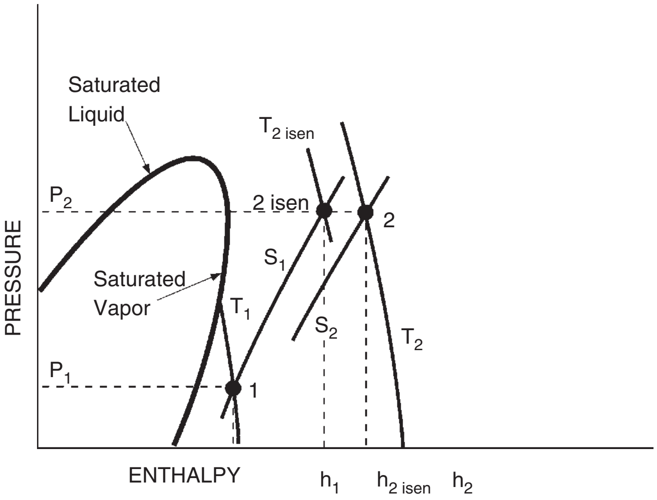

Figure 8-3. Compression process in a Mollier (pressure–enthalpy) diagram with lines of constant temperature and constant entropy. Courtesy of Solar Turbines Inc.

For ideal gases, the actual head can be calculated from

$$\Delta h = \frac{1}{\eta_s} C_p T_1 \left[\left(\frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] \tag{8-14}$$

and further, the actual discharge temperature (T_2) becomes

$$T_2 = \frac{T_1}{\eta_s} \left[\left(\frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] + T_1 \tag{8-15}$$

The second law of thermodynamics tells us

$$\dot{m}(s_2 - s_1) = \int_1^2 \frac{dq}{T} + s_{irr} \tag{8-16}$$

For adiabatic flows, where no heat q enters or leaves, the change in entropy simply describes the losses generated in the compression process. These losses come from the friction of gas with solid surfaces and the mixing of gas of different energy levels. An adiabatic, reversible compression process therefore does not change the entropy of the system, it is isentropic. Our equation for the actual head implicitly includes the entropy rise Δs, because

$$\Delta h = h(p_2,T_2) - h(p_1,T_1) = h(p_2,s_1 + \Delta s) - h(p_1,s_1) \tag{8-17}$$

If cooling is applied during the compression process (e.g., with intercoolers between two compressors in series), then the increase in entropy is smaller than for an uncooled process. Therefore, the power requirement will be reduced.

Using the polytropic process for comparison reasons works fundamentally the same way as using the isentropic process for comparison reasons (Beinecke and Luedtke, 1983). The difference lies in the fact that the polytropic process uses the same discharge temperature as the actual process, while the isentropic process has a different (lower) discharge temperature than the actual process for the same compression task. In particular, both the isentropic and the polytropic processes are reversible processes. The isentropic process is also adiabatic, whereas the polytropic processes assumes a specific amount of heat transfer. In order to fully define the isentropic compression process for a given gas, suction pressure, suction temperature, and discharge pressure have to be known. To define the polytropic process, in addition either the polytropic compression efficiency or the discharge temperature has to be known. The polytropic efficiency (η_p) is constant for any infinitesimally small compression step, which then allows us to write

$$\Delta h = \frac{1}{\eta_\mathrm{p}} \int_{p_1}^{p_2} v\,dp = \frac{\Delta h_\mathrm{p}}{\eta_\mathrm{p}} \tag{8-18}$$

where v is specific volume, and the polytropic head (Δh_p) can be calculated from

$$\Delta h_\mathrm{p} = Z_1 R T_1 \frac{n}{n-1}\left[\left(\frac{p_2}{p_1}\right)^{\frac{n-1}{n}} - 1\right] \tag{8-19}$$

This determines the polytropic efficiency as

$$\eta_\mathrm{p} = \frac{\Delta h_\mathrm{p}}{\Delta h} \tag{8-20}$$

For compressor designers, the polytropic efficiency has an important advantage. If a compressor has five stages, and each stage has the same isentropic efficiency η_s , then the overall isentropic compressor efficiency will be lower than η_s. If, for the same example, we assume that each stage has the same polytropic efficiency η_p, then the polytropic efficiency of the entire machine is also η_p. As far as performance calculations are concerned, the approach either using a polytropic head and efficiency or using isentropic head and efficiency will lead to the same result:

$$P = \dot{m} \cdot \Delta h = \dot{m}\left(\frac{\Delta h_\mathrm{s}}{\eta_\mathrm{s}}\right) = \dot{m}\left(\frac{\Delta h_\mathrm{p}}{\eta_\mathrm{p}}\right) \tag{8-21}$$

We also encounter energy conservation on a different level in turbomachines. The aerodynamic function of a turbomachine relies on the capability to trade two forms of energy: kinetic energy (velocity energy) and potential energy (pressure energy), as was outlined earlier.

8.7 REAL GAS BEHAVIOR AND EQUATIONS OF STATE

Understanding gas compression requires an understanding of the relationship among pressure, temperature, and density of a gas. An ideal gas exhibits the following behavior:

$$\frac{p}{\rho} = RT \tag{8-22}$$

where ρ is the density of gas and R is the gas constant.

Any gas at very low pressures ($p \to 0$) can be described by this equation.

For the elevated pressures found in natural gas compression, this equation becomes inaccurate, and an additional variable, the compressibility factor (Z), has to be added:

$$\frac{p}{\rho} = ZRT \tag{8-23}$$

Unfortunately, the compressibility factor itself is a function of pressure, temperature, and gas composition.

A similar situation arises when the enthalpy has to be calculated. For an ideal gas, we find

$$\Delta h = \int_{T_1}^{T_2} C_{\mathrm{p}}(T)\,dT \tag{8-24}$$

where C_{p} is only a function of temperature. This is a better approximation of the reality than the assumption of a perfect gas used in Equation (8-5).

In a real gas, we get additional terms for the deviation between real gas behavior and ideal gas behavior (Poling *et al.*, 2001):

$$\Delta h = [h^0 - h(p_1)]_{T_1} + \int_{T_1}^{T_2} C_{\mathrm{p}}\,dT - [h^0 - h(p_2)]_{T_2} \tag{8-25}$$

The terms $[h^0\text{-}h(p_1)]_{T_1}$ and $[h^0\text{-}h(p_2)]_{T_2}$ are called departure functions because they describe the deviation of real gas behavior from ideal gas behavior. They relate the enthalpy at some pressure and temperature to a reference state at low pressure, but at the same temperature. The departure functions can be calculated solely from an equation of state, while the term $\int C_{\mathrm{p}}dT$ is evaluated in the ideal gas state.

Equations of state are semiempirical relationships that allow one to calculate the compressibility factor as well as the departure function. For gas compression applications, the most frequently used equations of state are Redlich–Kwong, Soave–Redlich–Kwong, Benedict–Webb–Rubin, Benedict–Webb–Rubin–Starling and Lee–Kessler–Ploecker (Poling *et al.*, 2001).

Kumar *et al.* (1999) and Beinecke and Luedtke (1983) have compared these equations of state regarding their accuracy for compression applications. In general, all of these equations provide accurate results for typical applications in pipelines, i.e., for gases with a high methane content, and at pressures below about 3500 psia. It should be noted that the Redlich–Kwong equation of state is the most effective equation from a computational point of view (because the solution is found directly rather than through an iteration).

8.8 COMPRESSION RATIO

The compression ratio (CR) is the ratio of absolute discharge pressure to the absolute suction pressure. Mathematically

$$\mathrm{CR} = \frac{p_2}{p_1} \tag{8-26}$$

By definition, the compression ratio is always greater than one. If there are "*n*" stages of compression and the compression ratio is equal on each stage, then the compression ratio per stage is given by

$$\mathrm{CR}_{\mathrm{stage}} = \left(\frac{p_2}{p_1}\right)^{1/n} \tag{8-27}$$

If the compression ratio is not equal on each stage, then Equation (8-26) should be applied to each stage.

The term compression ratio can be applied to a single stage of compression and multistage compression. When applied to a single compressor or a single stage of compression, it is defined as the stage or unit compression ratio; when applied to a multistage compressor it is defined as the overall compression ratio. The compression ratio for typical gas pipeline compressors is rather low (usually below 2), except for stations that feed into pipelines. These low-pressure ratios can be covered in a single compression stage for reciprocating compressor and in a single body (with one or two impellers) in a centrifugal compressor.

While the pressure ratio is a valuable indicator for reciprocating compressors, the pressure ratio that a given centrifugal compressor can achieve depends primarily on gas composition and gas temperature. The centrifugal compressor is better characterized by its capability to achieve a certain amount of head (and a certain amount of head per stage). From Equation (8-9) it follows that the compressor head translates into a pressure ratio depending on gas composition and suction temperature. For natural gas (SG = 0.58–0.65), a single centrifugal stage can provide a pressure ratio of 1.4. The same stage would yield a pressure ratio of about 1.6 if it would compress air (SG = 1.0). The pressure ratio per stage is usually lower than the values given earlier for multistage machines.

For reciprocating compressors, the pressure ratio per compressor is usually limited by mechanical considerations (rod load) and temperature limitations. Reciprocating compressors can achieve cylinder pressure ratios of 3 to 6. The actual flange-to-flange ratio will be (due to the losses in valves and bottles) lower. For lighter gases (such as natural gas), the temperature limit will often limit the pressure ratio before the mechanical limits do. Centrifugal compressors are also limited by mechanical considerations (rotordynamics, maximum speed) and temperature limits. Whenever any limitation is involved, it becomes necessary to use multiple compression stages in series and intercooling. Furthermore, multistage compression may be required from a purely optimization standpoint. For example, with an increasing compression ratio, compression efficiency decreases and mechanical stress and temperature problems become more severe.

For pressure ratios higher than 3, it may be advantageous to install intercoolers between the compressors. Intercoolers are generally used between the stages to reduce the power requirements as well as to lower the gas temperature that may become undesirably high.[2] Theoretically, a minimum power requirement is obtained with perfect intercooling and no pressure loss between stages by making the ratio of compression the same in all stages. However, intercoolers invariably cause pressure losses (typically between 5 and 15 psi), which is a function of the cooler design. In the preliminary design, the pressure should be on the order of 10 psi for coolers (especially gas-to-air coolers, where the economics may be out of balance for lower pressure drop).

Note that an actual compressor with an infinite number of compression stages and intercoolers would approach isothermal conditions (where the power requirement of compression cycle is the absolutely minimum power necessary to compress the gas) if the gas were cooled to the initial temperature in the intercoolers.

Interstage cooling is usually achieved using gas-to-air coolers. The gas outlet temperature depends on the ambient air temperature. The intercooler exit temperature is determined by the cooling media. If ambient air is used, the cooler exit temperature, and thus the suction temperature to the second stage, will be about 20 to 30°F above the ambient dry bulb temperature. Water coolers can achieve exit temperatures about 20°F above the water supply temperature, but require a constant supply of

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[2] After the cooling, liquids may form. These liquids are removed in interstage scrubbers or knockout drums.

cooling water. Cooling towers can provide water supply temperatures of about wet bulb temperature plus 25°F.

For applications where the compressor discharge temperature is above some temperature limit of downstream equipment (a typical example is pipe coatings that limit gas temperatures to about 125 to 140°F) or has to be limited for other reasons (e.g., to not disturb the permafrost), an aftercooler has to be installed.

8.9 COMPRESSION DESIGN

Compressor design involves several steps. These include selection of the correct type of compressor, as well as the number of stages required. In addition, depending on the capacity, there is also a need to determine the horsepower requirement for the compression.

8.9.1 Determining Number of Stages

For reciprocating compressors, the number of stages is determined from the overall compression ratio as follows (Arnold and Stewart, 1999; Arnold, 2002).

1. Calculate the overall compression ratio. If the compression ratio is under 4, consider using one stage. If it is not, select an initial number of stages so that $\text{CR} < 4$. For initial calculations it can be assumed that the compression ratio per stage is equal for each stage. Compression ratios of 6 can be achieved for low-pressure applications, however, at the cost of higher mechanical stress levels and lower volumetric efficiency.
2. Calculate the discharge gas temperature for the first stage. If the discharge temperature is too high (more than 300°F), either increase the number of stages or reduce the suction temperature through precooling. It is recommended that the compressors be sized so that the discharge temperatures for all stages of compression be below 300°F. It is also suggested that the aerial gas coolers be designed to have a maximum of 20°F approach to ambient, provided the design reduces the suction temperature for the second stage, conserving horsepower and reducing power demand. If the suction gas temperature to each stage cannot be decreased, increase the number of stages by one and recalculate the discharge temperature.

Note that the 300°F temperature limit is used for reciprocating compressors because the packing life gets shortened above about 250°F and the lube oil, being involved directly in the compression process, will degrade faster at higher temperatures. The 350°F temperature limit pertains to centrifugals and is really a limit for the seals (although special seals can go to 400 to 450°F) or the pressure rating of casings and flanges. Because the lube oil in a centrifugal compressor does not come into direct contact with the process gas, lube oil degradation is not a factor.

If oxygen is present in the process gas in the amount that it can support combustion (i.e., the gas-to-oxygen ratio is above the lower explosive limit), much lower gas temperatures than mentioned earlier are required. In reciprocating machines, oil-free compression may be required (no lube oil can come into contact with the process gas). This requires special piston designs that can run dry. Also, a special precaution has to be taken to avoid hot spots generated by local friction.

8.9.2 Inlet Flow Rate

The compressor capacity is a critical component in determining the suitability of a particular compressor. We can calculate the actual gas flow rate[3] at suction conditions using

$$Q_{\mathrm{G}} = 0.0283 \frac{Z_1 T_1}{p_1} Q_{\mathrm{G,SC}} \tag{8-28}$$

where Q_{G} represents an actual cubic feet per minute flow rate of gas, T_1 represents the suction temperature in °R, p_1 represents suction pressure in psia, and $Q_{\mathrm{G,SC}}$ represents the standard volumetric flow rate of gas in MMSCFD.

Note that using the value of actual gas volumetric flow rate and discharge pressure, we can roughly determine the type of compressor appropriate for a particular application. Although there is a significant

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[3] While the sizing of the compressor is driven by the actual volumetric flow rate (Q_{G}), the flow in many applications is often defined as standard flow. Standard flow is volumetric flow at certain, defined conditions of temperature and pressure (60°F or 519.7°R and 14.696 psia) that are usually not the pressures and temperatures of the gas as it enters the compressor.

overlap, however, some of the secondary considerations, such as reliability, availability of maintenance, reputation of vendor, and price, will allow one to choose one of the acceptable compressors.

8.9.3 Compression Power Calculation

Once we have an idea about the type of compressor we will select, we also need to know the power requirements so that an appropriate prime mover can be designed for the job. After the gas horsepower (GHP) has been determined by either method, horsepower losses due to friction in bearings, seals, and speed increasing gears must be added. Bearings and seal losses can be estimated from Scheel's equation (GPSA, 2004). For reciprocating compressors, the mechanical and internal friction losses can range from about 3 to 8% of the design gas horsepower. For centrifugal compressors, a good estimate is to use 1 to 2% of the design GHP as mechanical loss.

To calculate brake horsepower (BHP), the following equation can be used.

$$\mathrm{BHP} = \mathrm{GHP} + \text{mechanical losses} \tag{8-29}$$

The detailed calculation of brake horsepower depends on the choice of type of compressor and number of stages. The brake horsepower per stage can be determined from Equation (8-30) (GPSA, 2004):

$$\mathrm{BHP} = 0.0854 \cdot Z_{\mathrm{ave}} \left[\frac{(Q_{\mathrm{G,SC}})(T_1)}{E \cdot \eta}\right]\left[\frac{k}{k-1}\right]\left[\left(\frac{p_2}{p_1}\right)^{\frac{k-1}{k}} - 1\right] \tag{8-30}$$

where BHP is brake horsepower per stage; Z_{ave} is average compressibility factor; $Q_{\mathrm{G,SC}}$ is standard volumetric flow rate of gas, MMSCFD; T_1 is suction temperature, °R; p_1,p_2 are pressure at suction and discharge flanges, respectively, psia; E is parasitic efficiency (for high-speed reciprocating units, use 0.72 to 0.82; for low-speed reciprocating units, use 0.72 to 0.85; and for centrifugal units, use 0.99); η is compression efficiency (1.0 for reciprocating and 0.80 to 0.87 for centrifugal units).

In Equation (8-30), parasitic efficiency (E) accounts for mechanical losses, and the pressure losses incurred in the valves and pulsation dampeners of reciprocating compressors (the lower efficiencies are usually associated with low-pressure ratio applications typical for pipeline compression).[4] Hence, suction and discharge pressures may have to be adjusted for the pressure losses incurred in the pulsation dampeners for reciprocation compressors. The compression efficiency accounts for the actual compression process. For centrifugal compressors, the lower efficiency is usually associated with pressure ratios of 3 and higher. Very low flow compressors (below 1000 acfm) may have lower efficiencies.

The total horsepower for the compressor is the sum of the horsepower required for each stage. Reciprocating compressors require an allowance for interstage pressure losses. It can be assumed that there is a 3% loss of pressure in going through the cooler, scrubbers, piping, and so on between the actual discharge of the cylinder and the actual suction of the next cylinder. For a centrifugal compressor, any losses incurred between the stages are already included in the stage efficiency. However, the exit temperature from the previous stage becomes the inlet temperature in the next stage. If multiple bodies are used, the losses for coolers and piping have to be included as described previously.

Example

Given the following information for a centrifugal compressor, answer the following questions.

Operating Conditions

$$P_\mathrm{s} = 750\,\mathrm{psia},\ P_\mathrm{d} = 1046.4\,\mathrm{psia},\ T_\mathrm{s} = 529.7^\circ\mathrm{R},\ T_\mathrm{d} = 582.6^\circ\mathrm{R},$$
$$Q_\mathrm{G,SC} = 349\,\mathrm{MMSCFD}$$

Gas Properties

$$\mathrm{SG} = 0.6,\ k = 1.3,\ Z_\mathrm{ave} = 0.95$$

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[4] Many calculation procedures for reciprocating compressors use numbers for E that are higher than the ones referenced here. These calculations require, however, that the flange-to-flange pressure ratio [which is used in Equation (8-30)] is increased by the pressure losses in the compressor suction and discharge valves, and pulsation dampeners. These pressure losses are significant, especially for low head, high flow applications.

Questions

1. What is the isentropic efficiency?
2. What is the actual volumetric flow rate?
3. What is the isentropic head?
4. What is the power requirement (assume a 98% mechanical efficiency)?

Solution

1. With rearranging Equation (8-15), we find

$$\eta_s = \frac{T_1}{T_2 - T_1}\left[\left(\frac{p_2}{p_1}\right)^{\frac{k-1}{k}} - 1\right]$$

$$\eta_s = \frac{529.7}{582.6 - 529.7}\left[\left(\frac{1046.4}{750}\right)^{\frac{0.3}{1.3}} - 1\right] = 0.80$$

2. Mass flow is

$$\dot{m} = 0.884 \times Q_{G,SC} \times SG = 0.884\frac{\text{lbm}}{\text{MMSCFD}} \times 349\text{MMSCFD} \times 0.6 = 185.1\,\text{lb}_\text{m}/\text{s}$$

and thus the volumetric flow becomes

$$Q_1 = \frac{53.35}{144} \cdot \frac{\dot{m} \cdot Z_1 \cdot T_1}{p_1 \cdot SG} = \frac{53.35}{144} \cdot \frac{185.1 \cdot 0.95 \cdot 529.7}{750 \cdot 0.6} = 76.68\,\text{cfs} = 4601\,\text{cfm}$$

3. The isentropic head follows from Equation (8-9) with $C_p = (53.35/SG)\,Z\,k/(k-1)$

$$\Delta h_s = \frac{53.35}{0.6} \times 0.95 \times \frac{1.3}{0.3} \times 529.7\left[\left(\frac{1046.4}{750}\right)^{\frac{0.3}{1.3}} - 1\right] = 14482\,\text{ft.lb}_\text{f}/\text{lb}_\text{m}$$

4. The power can be calculated from Equation (8-30)

$$\begin{aligned} \text{BHP} &= 0.0854 \times 0.95 \left[\frac{349 \cdot 529.7}{0.98 \cdot 0.8}\right]\left[\frac{1.3}{0.3}\right]\left[\left(\frac{1046.4}{750}\right)^{\frac{0.3}{1.3}} - 1\right] \\ &= 6622\,\text{hp} \end{aligned}$$

8.10 COMPRESSOR CONTROL

To a large extent, the compressor operating point will be the result of the pressure conditions imposed by the system. However, the pressures imposed by the system may in turn be dependent on the flow. Only if the conditions fall outside the operating limits of the compressor (e.g., frame loads, discharge temperature, available driver power, surge, choke, speed), do control mechanisms have to be in place. However, the compressor output may have to be controlled to match the system demand. The type of application often determines the system behavior. In a pipeline application, suction and discharge pressure are connected with the flow by the fact that the more flow is pushed through a pipeline, the more pressure ratio is required at the compressor station to compensate for the pipeline pressure losses. In process-related applications, the suction pressure may be fixed by a back pressure-controlled production separator. In boost applications, the discharge pressure is determined by the pressure level of the pipeline the compressor feeds into, whereas the suction pressure is fixed by the process. In oil and gas field applications, the suction pressure may depend on the flow because the more gas is moved out of the gas reservoir, the lower the suction pressure has to be. The operation may require constant flow despite changes in suction or discharge pressure. Compressor flow, pressure, or speed may have to be controlled. The type of control also depends on the compressor driver. Both reciprocating compressors and centrifugal compressors can be controlled by suction throttling or recirculating of gas. However, either method is very inefficient for process control (but may be used to protect the compressor) because the reduction in flow or head is not accompanied by a significant reduction in the power requirement.

8.10.1 Reciprocating Compressors

The following mechanisms may be used to control the capacity of reciprocating compressors: suction pressure, variation of clearance, speed, valve unloading, and recycle. Reciprocating compressors tend to have a rather steep head versus flow characteristic. This means that changes in pressure ratio have a very small effect on the actual flow through the machine.

Controlling the flow through the compressor can be accomplished by varying the operating speed of the compressor. This method can be used if the compressor is driven by an internal combustion engine or a variable speed electric motor. Internal combustion engines, along with variable speed electric motors, produce less power if they operate at a speed different from their optimum speed. Internal combustion engines allow for speed control in the range of 70 to 100% of maximum speed.

If the driver is a constant speed electric motor, the capacity control consists of either inlet valve unloaders or clearance unloaders. Inlet valve unloaders can hold the inlet valve into the compressor open, thereby preventing compression. Clearance unloaders consist of pockets that are opened when unloading is desired. The gas is compressed into them at the compression stroke and expands back into the cylinder on the return stroke, thus reducing the intake of additional gas and subsequently the compressor capacity. Additional flexibility is achieved using several steps of clearance control and combinations of clearance control and inlet valve control. Figure 8-4 shows the control characteristic of such a compressor.

Reciprocating compressors generate flow pulsations in the suction and discharge lines that have to be controlled to prevent over- and underloading of the compressors, to avoid vibration problems in the piping or other machinery at the station, and to provide a smooth flow of gas. Flow pulsations can be reduced greatly by properly sized pulsation bottles or pulsation dampeners in the suction and discharge lines.

8.10.2 Centrifugal Compressors

As with reciprocating compressors, the compressor output must be controlled to match the system demand. The operation may require constant flow despite changes in suction or discharge pressure. Compressor flow, pressure, or speed may have to be controlled. The type of control also depends on the compressor driver. Centrifugal compressors tend to have

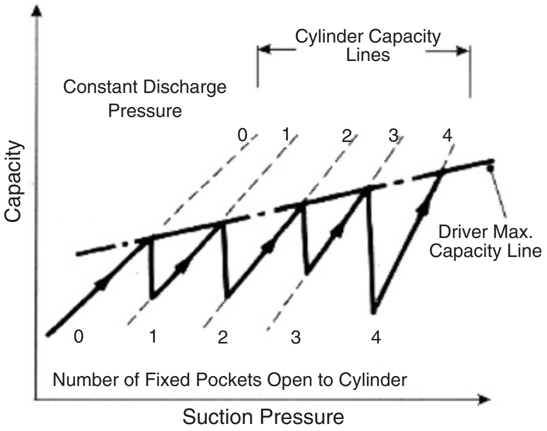


Figure 8-4. Control characteristic of a reciprocating compressor with constant speed driver and pockets (GPSA, 2004).

a rather flat head versus flow characteristic. This means that changes in pressure ratio have a significant effect on the actual flow through the machine.

Compressor control is usually accomplished by speed control, variable guide vanes, suction throttling, and recycling of gas. Only in rare cases are adjustable diffuser vanes used. To protect the compressor from surge, recycling is used. Controlling the flow through the compressor can be accomplished by varying the operating speed of the compressor. This is the preferred method of controlling centrifugal compressors. Two shaft gas turbines and variable speed electric motors allow for speed variations over a wide range (usually from 50 to 100% of maximum speed or more). Virtually any centrifugal compressor installed since the early 1990s in pipeline service is driven by variable speed drivers, usually a two-shaft gas turbine. Older installations and installations in other than pipeline services sometimes use single shaft gas turbines (which allow a speed variation from about 90 to 100% speed) and constant speed electric motors. In these installation, suction throttling or variable inlet guide vanes are used to provide means of control.

The operating envelope of a centrifugal compressor is limited by the maximum allowable speed, the minimum allowable speed, the minimum flow (surge flow), and the maximum flow (choke or stonewall) (Figure 8-5). Another limiting factor may be the available driver power. Only the minimum flow requires special attention because it is defined by

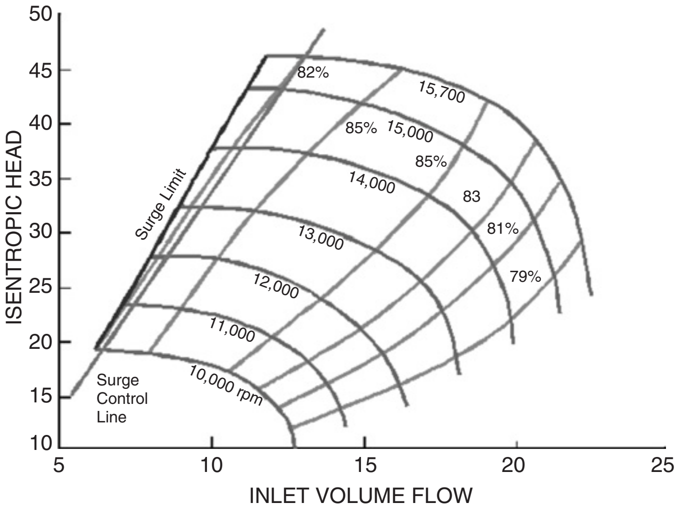


Figure 8-5. Typical performance map for a centrifugal compressor (Kurz and Brun, 2001).

an aerodynamic stability limit of the compressor. Crossing this limit to lower flows will cause pulsating intermittent flow reversals in the compressor (surge), which eventually can damage the compressor. Modern control systems can detect this situation and shut the machine down or prevent it entirely by automatically opening a recycle valve. For this reason, virtually all modern compressor installations use a recycle line (Figure 8-6) with a control valve that allows the flow to increase through the compressor if it comes near the stability limit. Modern control systems constantly monitor the operating point of the compressor in relation to its surge line and automatically open or close the recycle valve if necessary. The control system is designed to compare the measured operating point of the compressor with the position of the surge line (Figure 8-5). To that end, flow, suction pressure, discharge pressure, and suction temperature, as well as compressor speed, have to be measured.

8.11 COMPRESSOR PERFORMANCE MAPS

8.11.1 Reciprocating Compressors

Figure 8-7 shows some typical performance maps for reciprocating compressors. Because the operating limitations of a reciprocating compressor

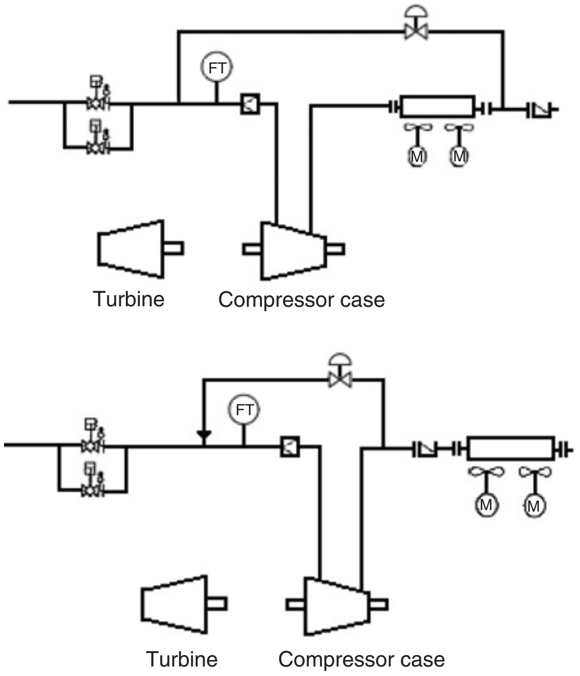


Figure 8-6. Recycle lines for centrifugal compressors: cooled (top) and hot (bottom) recycle lines. Courtesy of Solar Turbines Inc.

are often defined by mechanical limits (especially maximum rod load) and because the pressure ratio of the machine is very insensitive to changes in suction conditions and gas composition, we usually find maps depicting suction and discharge pressures and actual flow. Maps account for the effect of opening or closing pockets and for variations in speed.

8.11.2 Centrifugal Compressors

For a centrifugal compressor, the head (rather than the pressure ratio) is rather invariant with the change in suction conditions and gas composition. As with the reciprocating compressor, the flow that determines the operating point is the actual flow as opposed to mass flow or standard flow.

Head versus actual flow maps (Figure 8-5) are therefore the most usual way to describe the operating range of a centrifugal compressor.

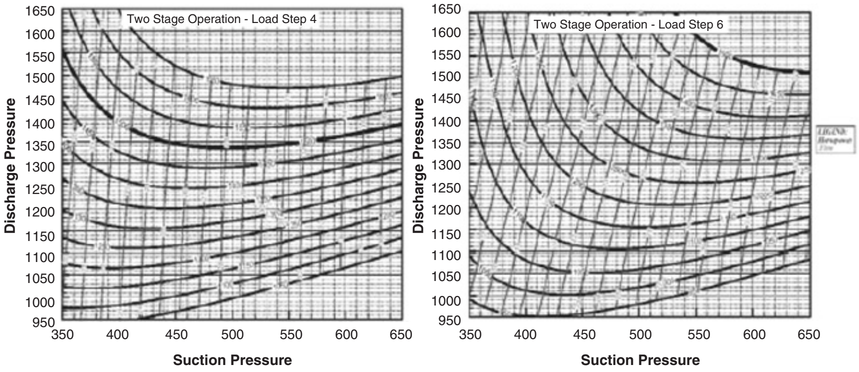


Figure 8-7. Typical maps for a speed-controlled reciprocating compressor for two different load steps. Courtesy of Solar Turbines Inc.

These maps change very little even if the inlet conditions or the gas composition changes. They depict the effect of changing the operating speed and define the operating limits of the compressor, such as surge limit, maximum and minimum speed, and maximum flow at choke conditions. Every set of operating conditions, given as suction pressure, discharge pressure, suction temperature, flow, and gas composition, can be converted into isentropic head and actual flow using the relationships described previously. Once the operating point is located on a head flow map, the efficiency of the compressor, and the required operating speed, as well as the surge margin, can be determined.

REFERENCES

Akhtar, M.S., “Selection and Optimization of Centrifugal Compressors for Oil and Gas Applications.” Paper presented at the GPA Europe Spring Meeting, Bergen, Norway (May 2002).

Arnold, J.L., “Compressors in Energy-Related Applications.” Compression, Chapter 5, pp. 17–19 (2002).

Arnold, K., and Stewart, M., “Surface Production Operations, Vol. 2: Design of Gas-Handling Systems and Facilities,” 2nd Ed. Gulf Professional Publishing, Houston, TX (1999).

Beinecke, D., and Luedtke, K., “Die Auslegung von Turboverdichtern unter Beruecksichtigung des realen Gasverhaltens,” VDI Berichte 487 (1983).

Campbell, J.M., “Gas Conditioning and Processing,” 3rd Ed. Campbell Petroleum Series, Norman, OK (1992).

“GPSA Engineering Data Book,” 12th Ed. Gas Processors Suppliers Association, Tulsa, OK (2004).

Kumar, S., Kurz, R., and O’Connell, J.P., “Equations of State for Compressor Design and Testing.” ASME Paper 99-GT-12 (1999).

Kurz, R., and Brun, K., Degradation in gas turbine systems. *ASME J. Eng. Gas Turbines Power* **123**(1), 70–77 (2001).

Kurz, R., Ohanian, S., and Lubomirsky, M., “On Compressor Station Layout.” ASME Paper GT2003-38019 (2003).

Mohitpour, M., Golshan, H., and Murray, A., “Pipeline Design and Construction: A Practical Approach.” ASME Press, American Society of Mechanical Engineers, New York (2000).

Poling, B.E., Prausnitz, J.M, and O’Connell, J.P., “The Properties of Gases and Liquids.” McGraw-Hill, New York (2001).

C H A P T E R   9

NATURAL GAS DEHYDRATION

9.1 INTRODUCTION

Natural, associated, or tail gas usually contains water, in liquid and/or vapor form, at source and/or as a result of sweetening with an aqueous solution. Operating experience and thorough engineering have proved that it is necessary to reduce and control the water content of gas to ensure safe processing and transmission. The major reasons for removing the water from natural gas are as follow.

1. Natural gas in the right conditions can combine with liquid or free water to form solid hydrates that can plug valves fittings or even pipelines.
2. Water can condense in the pipeline, causing slug flow and possible erosion and corrosion.
3. Water vapor increases the volume and decreases the heating value of the gas.
4. Sales gas contracts and/or pipeline specifications often have to meet the maximum water content of 7 lb H_2O per MMscf.

Pipeline drips installed near wellheads and at strategic locations along gathering and trunk lines will eliminate most of the free water lifted from the wells in the gas stream. Multistage separators can also be deployed to ensure the reduction of free water that may be present. However, removal of the water vapor that exists in solution in natural gas requires a more complex treatment. This treatment consists of "dehydrating" the natural gas, which is accomplished by lowering the dew point temperature of the gas at which water vapor will condense from the gas.

There are several methods of dehydrating natural gas. The most common of these are liquid desiccant (glycol) dehydration, solid desiccant dehydration, and refrigeration (i.e., cooling the gas). The first two methods utilize mass transfer of the water molecule into a liquid solvent (glycol solution) or a crystalline structure (dry desiccant). The third method employs cooling to condense the water molecule to the liquid phase with the subsequent injection of inhibitor to prevent hydrate formation. However, the choice of dehydration method is usually between glycol and solid desiccants. Each of these methods is discussed in this chapter. Refrigeration technology is also discussed in Chapter 10.

Several other dehydration technologies (i.e., membranes, vortex tube, and supersonic processes) are used less commonly and are not discussed here. There are also many commercially available processes for customized dehydration systems. These types of processes (i.e., IFPEXOL) or solvents (i.e., Selexol) are often designed to enhance conventional equipment performance. However, the suitability of these processes should be evaluated on a case-specific basis.

9.2 WATER CONTENT DETERMINATION

The first step in evaluating and/or designing a gas dehydration system is to determine the water content of the gas. This datum is most important when one designs sour gas dehydration facilities and estimates water production with sour gas in the plant inlet separator. For most gas systems the McKetta and Wehe (1958) chart, generated from empirical data, provides the standard for water content determination. This chart can be used to predict the saturated water content of sweet, pipeline quality natural gas. There are also several methods available for determining the water content of sour natural gas at saturation (Carroll, 2002). In general, for acid gas concentrations less than about 30%, existing methods are satisfactory. For higher acid gas concentrations (above 50%), particularly at higher pressures, existing methods can lead to serious errors in estimating water contents (Hubbard, 1993). An appropriate method has been introduced by Wichert and Wichert (2003). It is chart based and provides good estimates of the equilibrium water vapor content of sour natural gas for a range of conditions, including H_2S contents of 3–38 mole % with CO_2 contents of 3–43 mole %, pressures from 290 to 10153 psia, and temperatures from 50 to 347°F. The overall average error of this method is less than 1%. However, a few points showed a discrepancy of more than ±10%. To estimate the water content of sour gas:

1. Determine the equilibrium water vapor content of sweet gas at the operating temperature and pressure conditions using Figure 9-1.
2. Determine the mole % H_2S equivalent concentration of the sour gas as

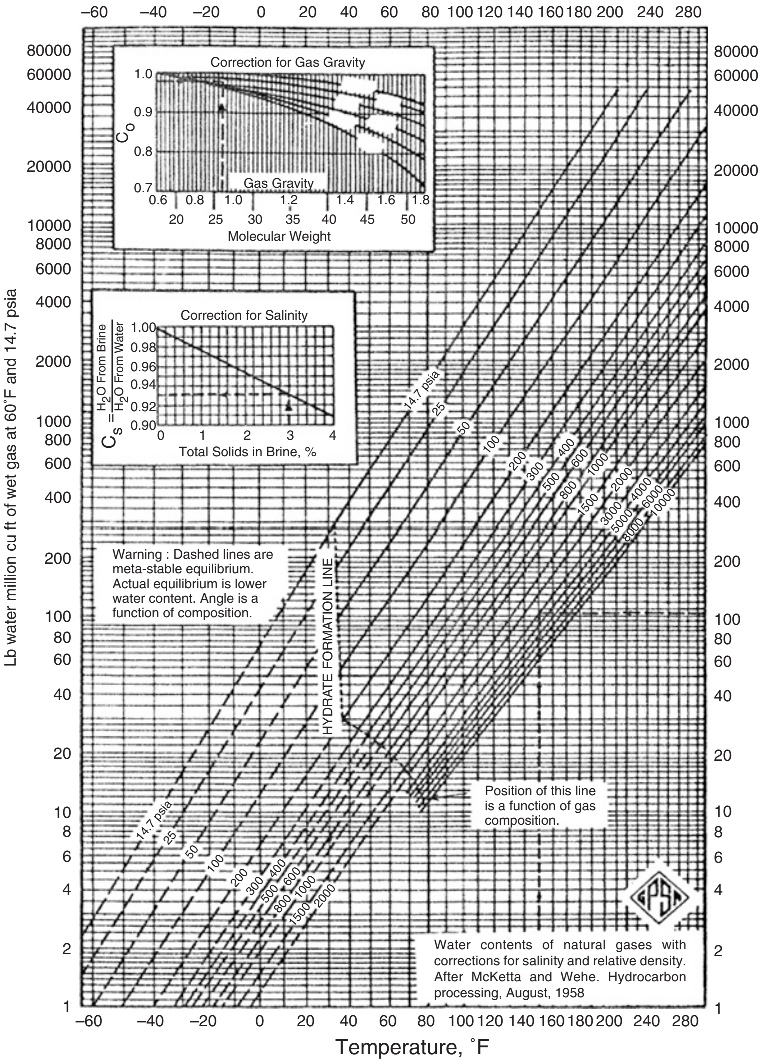
$$\text{mole \% } H_2S \text{ equivalent} = \text{mole \% } H_2S + 0.7 \times \text{mole \% } CO_2 \tag{9-1}$$

3. From Figure 9-2 at the bottom left-hand temperature scale, move to the right to the mole % H_2S equivalent (interpolate between the lines if necessary).
4. From this point, move to the upper chart, to the pressure of interest. From the pressure point, move to the left, to the ratio scale.
5. Multiply the value from step 4 by the water content determined for sweet gas in step 1. The result is the estimate of the saturated water content of the sour gas at the pressure and temperature of interest.

Water content calculations based on proprietary/literature equations of state (EOS) are also common. Equation of state methods typically overestimate the water content for higher acid gas concentrations, particularly at higher pressures, when compared to experimental data (Hubbard, 1993).

9.3 GLYCOL DEHYDRATION

Among the different gas drying processes, absorption is the most common technique, where the water vapor in the gas stream becomes absorbed in a liquid solvent stream. Glycols are the most widely used absorption liquids as they approximate the properties that meet commercial application criteria. Several glycols have been found suitable for commercial application.



Warning : Dashed lines are meta-stable equilibrium. Actual equilibrium is lower water content. Angle is a function of composition.

Position of this line is a function of gas composition.

Water contents of natural gases with corrections for salinity and relative density. After McKetta and Wehe. Hydrocarbon processing, August, 1958

Figure 9-1. McKetta and Wehe (1958) pressure–temperature correlation (GPSA, 1998).

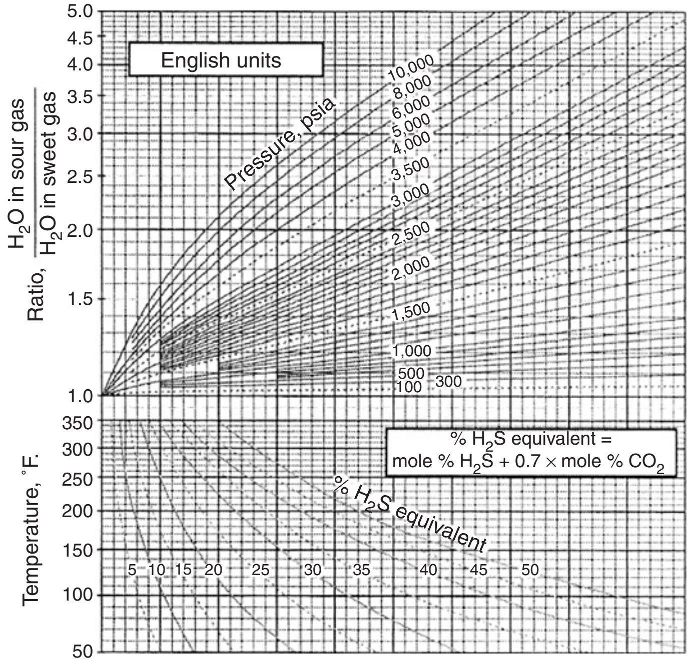


Figure 9-2. Water content ratio chart (Wichert and Wichert, 2003).

The commonly available glycols[1] and their uses are described as follows (Katz *et al*., 1959).

1. Monoethylene glycol (MEG); high vapor equilibrium with gas so tend to lose to gas phase in contactor. Use as hydrate inhibitor where it can be recovered from gas by separation at temperatures below 50°F.
2. Diethylene glycol (DEG); high vapor pressure leads to high losses in contactor. Low decomposition temperature requires low reconcentrator temperature (315 to 340°F) and thus cannot get pure enough for most applications.
3. Triethylene glycol (TEG); most common. Reconcentrate at 340–400°F, for high purity. At contactor temperatures in excess of 120°F, there is a tendency to high vapor losses. Dewpoint depressions up to 150°F are possible with stripping gas.

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[1] Physical properties of the most common glycols are given in Appendix 3.

4. Tetraethylene glycol (TREG); more expensive than TEG but less loss at high gas contact temperatures. Reconcentrate at 400 to 430°F.

TEG is by far the most common liquid desiccant used in natural gas dehydration. It exhibits most of the desirable criteria of commercial suitability listed here (Manning and Thompson, 1991; Hubbard, 1993).

1. TEG is regenerated more easily to a concentration of 98–99% in an atmospheric stripper because of its high boiling point and decomposition temperature.
2. TEG has an initial theoretical decomposition temperature of 404°F, whereas that of diethylene glycol is only 328°F (Ballard, 1966).
3. Vaporization losses are lower than monoethylene glycol or diethylene glycol. Therefore, the TEG can be regenerated easily to the high concentrations needed to meet pipeline water dew point specifications.
4. Capital and operating costs are lower.

Natural gas dehydration with TEG is discussed under the following topics: process description and design considerations.

9.3.1 Process Description

In this section, dehydration of natural gas by TEG is first outlined by summarizing the flow paths of natural gas and glycol. Then the individual components of a typical TEG unit are described in detail. As shown in Figure 9-3, wet natural gas first typically enters an inlet separator to remove all liquid hydrocarbons from the gas stream. Then the gas flows to an absorber (contactor) where it is contacted countercurrently and dried by the lean TEG. TEG also absorbs volatile organic compounds (VOCs[2]) that vaporize with the water in the reboiler. Dry natural gas exiting the absorber passes through a gas/glycol heat exchanger and then into the sales line. The wet or "rich" glycol exiting the absorber flows through a coil in the accumulator where it is preheated by hot lean glycol. After the glycol–glycol heat exchanger, the rich glycol enters the stripping column and flows down the packed bed section into the reboiler. Steam generated in the reboiler strips absorbed water and VOCs out of the glycol as it rises

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[2] VOC emissions are an environmental challenge for the natural gas industry, hence glycol dehydration systems require monitoring and control of VOC emissions.

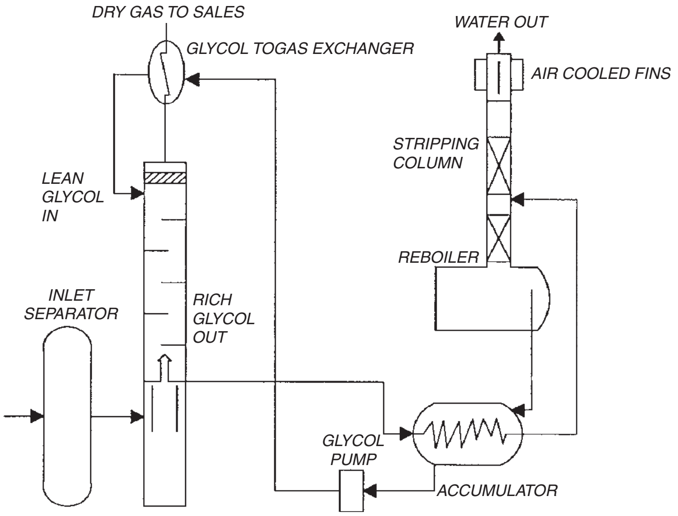

Figure 9-3. Simplified flow diagram for TEG dehydration (Manning and Thompson, 1991).

up the packed bed. The water vapor and desorbed natural gas are vented from the top of the stripper. The hot regenerated lean glycol flows out of the reboiler into the accumulator (surge tank) where it is cooled via cross exchange with returning rich glycol; it is pumped to a glycol/gas heat exchanger and back to the top of the absorber.

The simple flow diagram shown in Figure 9-3 is typical of small gas dehydration units where unattended operation is the prime concern. Larger units are monitored daily and the efficiency and operational cost are improved by the additional equipment, shown in Figure 9-4, where:

1. Rich glycol leaves the absorber and enters a cooling coil that controls the water reflux rate at the top of the stripper. This temperature control ensures that the water vapor leaving the still does not carry over excess glycol.
2. Heat exchange between the cool rich glycol and the hot lean glycol is improved by using two or more shell and tube heat exchangers in series. The increased heat recovery reduces fuel consumption in the reboiler and protects the glycol circulation pump from being overheated, it also allows the flash tank and filter to operate at approximately 150°F. Higher flash temperatures will ensure maximum entrained gas to be removed from the rich TEG.

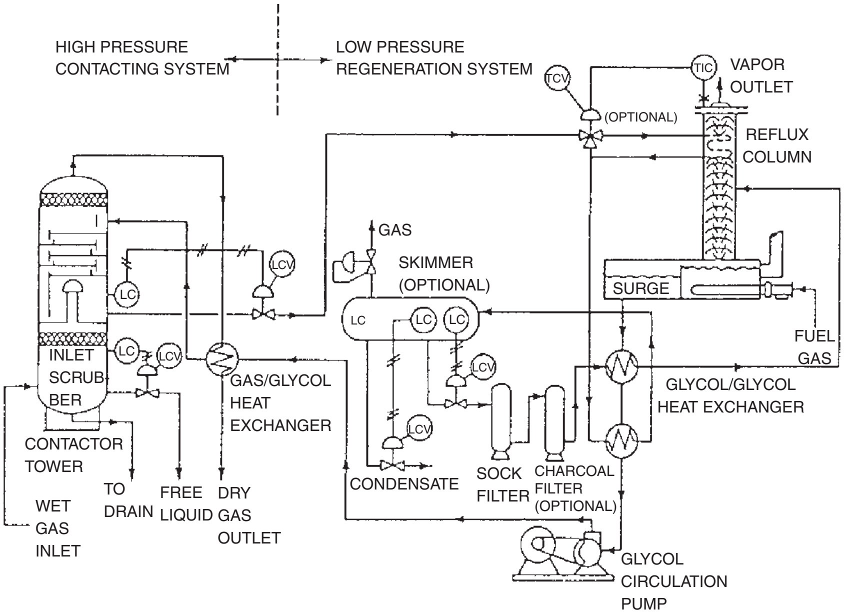

Figure 9-4. Typical flow diagram for a glycol dehydration unit (NATCO, 1984).

3. Rich glycol is flashed to remove dissolved hydrocarbons. The latter can be used for fuel and/or stripping gas.
4. The rich glycol is filtered before being heated in the reconcentrator. This prevents impurities such as solids and heavy hydrocarbons from plugging the packed column and fouling the reboiler fire tube.

9.3.2 Design Considerations

More detailed information about each equipment operation and design used in a TEG dehydration unit is described as follow.

9.3.2.1 Absorber (Contactor)

The incoming wet gas and the lean TEG are contacted countercurrently in the absorber to reduce the water content of the gas to the required specifications. (In an ethylene glycol system, glycol is injected directly into the natural gas stream; therefore, an absorber is not used.) The key design parameters for the absorber are:

- Gas flow rate and specific gravity
- Gas temperature
- Operating pressure (gas pressure)
- Outlet dew point or water content required

The amount of water to be removed in a TEG system is calculated from the gas flow rate, the water content of incoming gas, and the desired water content of outgoing gas. The water removal rate, assuming the inlet gas is water saturated, can be determined as

$$W_{\mathrm{r}} = \frac{Q_{\mathrm{G}}\,(W_{\mathrm{i}} - W_{\mathrm{o}})}{24} \tag{9-2}$$

where W_{r} is water removed, lb/hr; W_{i} is water content of inlet gas, lb/MMscf; W_{o} is water content of outlet gas, lb/MMscf; and Q_{G} gas flow rate, MMscfd.

The glycol circulation rate is determined on the basis of the amount of water to be removed and is usually between 2 and 6 gallons of TEG per pound of water removed, with 3 gallons TEG/lb water being typical. Higher circulation rates provide little additional dehydration while increasing reboiler fuel and pumping requirements. Problems can arise if the TEG circulation rate is too low; therefore, a certain amount of overcirculation is desired. An excessive circulation rate may overload the reboiler and prevent good glycol regeneration. The heat required by the reboiler is directly proportional to the circulation rate. Thus, an increase in circulation rate may decrease reboiler temperature, decreasing lean glycol concentration, and actually decrease the amount of water that is removed by the glycol from the gas. Only if the reboiler temperature remains constant will an increase in circulation rate lower the dew point of the gas. An overly restricted circulation rate can also cause problems with tray hydraulics, contactor performance, and fouling of glycol-to-glycol heat exchangers. Therefore, operators should include a margin of safety, or "comfort zone," when calculating reductions in circulation rates. An optimal circulation rate for each dehydration unit typically ranges from 10 to 30% above the minimum circulation rate (EPA430-B-03-013).

The minimum glycol circulation rate can then be calculated as

$$Q_{\mathrm{TEG,min}} = G \times W_{\mathrm{r}} \tag{9-3}$$

where $Q_{\mathrm{TEG,min}}$ is the minimum TEG circulation rate (gal TEG/hr) and G is the glycol-to-water ratio (gal TEG/lb water removed). The industry accepted rule of thumb is 3 gallons of TEG per pound of water removed.

Figure 9-5 shows the effect of TEG concentration and circulation rate on dew point depression for a fixed amount of absorber contact. Note that the curves become relatively flat at high circulation rates.

The diameter of the absorber and the number of absorber stages are selected on the basis of the gas and glycol flow rates and gas-specific gravity. The diameter of the contactor (absorber) can be estimated from the Souders and Brown (1932) correlation as follows.

$$V_{\max} = K_{\mathrm{SB}}\left[\frac{\rho_{\mathrm{L}} - \rho_{\mathrm{G}}}{\rho_{\mathrm{G}}}\right] = \left[\frac{4Q_{\mathrm{G}}}{\pi D^2}\right] \tag{9-4}$$

where D is internal diameter of glycol contactor, ft; Q_{G} is gas volumetric flow rate, ft^3/hr; $V_{\max}$ is maximum superficial gas velocity, ft/hr; K_{SB} is

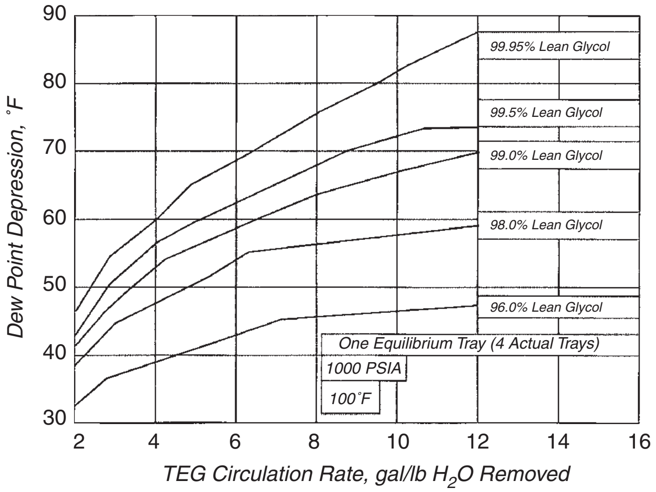

Figure 9-5. Effect of TEG concentration and circulation rate on dew point depression (GPSA, 1998).

Souders and Brown coefficient, ft/hr; ρ_L is glycol density, lb/ft^3; and ρ_G is gas density at column condition, lb/ft^3.

Traditionally, the glycol absorber contains 6–12 trays that provide an adequate contact area between the gas and the glycol (Kean *et al*., 1991). The more trays, the greater the dew point depression for a constant glycol circulation rate and lean glycol concentration. Conversely, specifying more trays with the same TEG concentration, a lower circulation rate is required. By specifying more trays, fuel savings can be realized because the heat duty of the reboiler is directly related to the glycol circulation rate.

Although either bubble cap trays or valve trays may be used, some operators prefer bubble cap trays because they are suitable for viscous liquids, handle high turndown ratios and low liquid/gas ratios well, and are not subject to weeping. Calculated tray efficiency values are dependent on the TEG/water equilibrium data used. To achieve an accurate design method, column efficiencies consistent with accurate equilibrium data must be therefore recommended. There are still uncertainties in equilibrium data for the TEG/water/natural gas system. However, when using accurate equilibrium data, an overall bubble cap tray efficiency of 40–50% and a Murphree[3] efficiency of 55–70% can be expected at normal absorption conditions, 86°F and 99–99.5% wt TEG (Oi, 1999). Earlier, overall tray efficiencies between 25 and 40% have been recommended for design. This is regarded as too conservative. It has been suggested that using 50% Murphree efficiency based on accurate phase equilibrium data should give a conservative design at normal high-pressure dehydration conditions (Oi, 2003).

The standard tray spacing in glycol contactors is 24 inches; closer spacing increases glycol losses if foaming occurs because of greater entrainment. The total height of the contactor column will be based on the number of trays required plus an additional 6–10 ft to allow space for a vapor disengagement area above the top tray and an inlet gas area at the bottom of the column.

One option to the trayed TEG contactor is the use of structured packing. Structured packing was developed as an alternative to random packing to improve mass transfer control by use of a fixed orientation of the transfer

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[3] The Murphree tray efficiency (for tray number *n*) is defined as (the mole fraction in the gas from tray *n* – the mole fraction from the tray below)/(the mole fraction in equilibrium with the liquid at tray *n* – the mole fraction from the tray below).

surface. The combination of high gas capacity and reduced height of an equilibrium stage, compared with trayed contactors, makes the application of structured packing desirable for both new contactor designs and existing trayed-contactor capacity upgrades. Hence, the structured packing may offer potential cost savings over trays (Kean *et al.*, 1991). A detailed discussion on calculating column diameter and the required packing height is given by Ghoshal and Mukhopadhyay (1993).

The absorber is usually vertical to allow proper glycol flow with sufficient gas/liquid contact and operates at the pressure of the incoming gas. A demister pad or mist eliminator at the top of the absorber or a separator vessel following the absorber can reduce glycol losses by preventing glycol from being carried out with the dry gas.

9.3.2.2 Still (Stripper)

The still or stripper column is used in conjunction with the reboiler to regenerate the glycol. On many dehydrators, the still is placed vertically on top of the reboiler so that vapor from the reboiler directly enters the bottom of the distillation column. A given lean TEG concentration is produced in the reboiler and still column (regenerator) section by the control of reboiler temperature, pressure, and the possible use of a stripping gas. As shown in Figure 9-6, the reboiler temperature controls

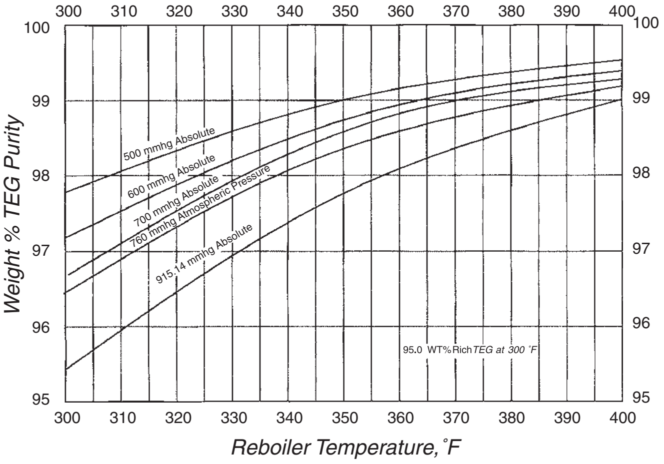

Figure 9-6. Glycol purity vs reboiler temperature at different levels of vacuum (GPSA, 1998).

the concentration of the water in the lean glycol. Reboiler temperatures for TEG are limited to 400°F, which limits the maximum lean glycol concentration without stripping gas. Some operators limit the reboiler temperature to between 370 and 390°F to minimize degradation of the glycol. This effectively limits the lean glycol concentration to between 98.5 and 98.9%.

There are improved regeneration techniques that have higher glycol concentration and, therefore, lower treated gas dew points. By injecting dry (stripping) gas into the base of the glycol reboiler in order to (1) strip off water vapor from the glycol by reducing the vapor partial pressure and (2) agitate the glycol to accelerate the release of water vapor, TEG concentration increases from 99.1 to 99.6% by weight. A dry gas injection process can be enhanced using a packed column of countercurrent gas stripping, which increases the capability of TEG for gas dehydration by reconcentrating the glycol to as high as 99.6% (Coker, 1994). The Drizo process, developed by Dow Chemical Co., also uses hydrocarbon solvents as azeotropic stripping agents for improving trace quantities of water in the TEG. The process performs a tertiary azeotropic distillation, resulting in glycol concentrations to 99.9% (Smith and Tsai, 1987). This process avoids flaring stripping gas and reduces operating costs. Other patented processes in use to enhance the glycol purity and thereby achieve a more stringent water dew point depression (CLEANOL+, COLDFINGER, PROGLY, and ECOTEG) are described in GPSA (1998).

The diameter of the still is based on the liquid load (rich glycol and reflux) and the vapor load (water vapour and stripping gas). Manufacturers'chart or standard sizes based on the required reboiler heat load may be used to determine the column diameter (Caldwell, 1976; Sivalls, 1976). Alternatively, the following approximate equation is used:

$$D = 9(Q_{\mathrm{TEG}})^{0.5} \tag{9-5}$$

where Q_{TEG} is TEG circulation rate, gal/min; and D is inside diameter of stripping column, inch.

Smaller diameter towers (less than 2 ft in diameter) are often packed with ceramic Intalox saddles or stainless steel Pall rings instead of trays. Larger-diameter towers may be designed with 10 to 20 bubble cap trays or structured packing.

To prevent excessive glycol losses from vaporization at the top of the still column, reflux is controlled by a condenser maintained at about 215°F or lower if stripping gas is used. A few larger gas dehydration units may use a tubular water-cooled condenser with temperature control. Temperature control can also be obtained by circulating cool rich glycol through a reflux coil inside the top of the stripping column. This system normally includes a bypass valve that allows the operator to better control the temperature at the top of the column. Many smaller field dehydrators employ a finned atmospheric condenser at the outlet at the top of the still.

9.3.2.3 Reboiler

The reboiler and still are typically a single piece of equipment. The reboiler supplies heat to regenerate the rich glycol in the still by simple distillation. The separation is relatively easy because of the wide difference in boiling points of water and glycol. Most remote field locations use a direct-fired firebox to provide the heat for vaporization. A horizontal U-shaped firetube fires a portion of the natural gas or draws from the fuel gas system, which may include flash gas from the phase separator. Some sites have also burned noncondensable vent gas from their condenser system in the reboiler, although there are safety concerns associated with this practice. Larger dehydrator systems (such as at gas plants) may use indirect heat sources such as dowtherm heat transfer fluid (hot oil), electricity, or medium pressure steams. Reboiler duty can be estimated by the following equation (Sivalls, 1976):

$$Q_{\mathrm{R}} = 900 + 966\ (\mathrm{G}) \qquad (9\text{-}6)$$

where Q_{R} is regenerator duty, Btu/lb H_2O removed; and G is glycol-to-water ratio, gal TEG/lb H_2O removed. This estimate does not include stripping gas and makes no allowance for combustion efficiency.

The reboiler normally operates at a temperature of 350 to 400°F for a TEG system; this temperature controls the lean glycol water concentration. The purity of the lean glycol can be increased by raising the reboiler temperature, but TEG starts to decompose at 404°F. The reboiler should not be operated above 400°F to allow a safety margin to prevent decomposition, and the burner should have a high temperature shutdown for safety. A continuous spark ignition system, or a spark igniter to relight the pilot if it goes out, is also useful. Normally, a conventional reboiler operating at slightly above atmospheric pressure can provide TEG with a purity of 98.7% at 400°F, which is sufficient for an 85°F dew point depression. The heat flux in the reboiler must be high enough for vaporization, but not so high as to cause glycol decomposition. For a typical TEG reboiler with a bulk temperature of 400°F, a design heat flux of 8000 Btu/(fr^{2}.hr) is recommended.

9.3.2.4 Surge Tank (Accumulator)

Lean glycol from the reboiler is routed through an overflow pipe or weir to a surge tank or accumulator. Because this vessel is not insulated in many cases, the lean glycol is cooled to some extent via heat loss from the shell. The surge tank may also contain a glycol/glycol heat exchanger. If the surge tank contains the glycol/glycol heat exchanger, the tank should be sized to allow a 30-minute retention time for the lean glycol. Some designs also provide for a separate lean glycol storage tank. The storage tank, when used as a surge tank, may be vented to allow accumulated gases to escape or is sometimes fitted with an inert gas blanket to prevent the oxygen from contacting the glycol and causing oxidation. Venting from the surge tank is only a minor emission source because the lean glycol has already been stripped of most VOCs in the still.

9.3.2.5 Heat Exchanger

Two types of heat exchangers are found in glycol plants: glycol/glycol and gas/glycol. The design and operation of the two types of exchangers are discussed next.

9.3.2.5.1 Glycol/Glycol Exchanger A glycol/glycol exchanger cools the lean glycol while preheating the rich glycol. It may be an external exchanger or may be located within the surge tank (an integral exchanger). For small standard designs, the integral exchanger is economical to fabricate but may not heat the rich glycol above 200°F. Types of external glycol/glycol exchangers include the following.

- Insulated double pipe with finned inner pipe
- Shell and tube
- Plate and frame type

All three types of external heat exchangers can preheat the rich glycol to about 300°F. The 100°F preheat improvement possible with an external exchanger reduces the reboiler duty by approximately 600 Btu/gallon.

9.3.2.5.2 Dry Gas/Lean Glycol Exchanger This type of exchanger uses the exiting dry natural gas to control the lean glycol temperature to the absorber. High glycol temperatures relative to the gas temperature reduce the moisture-absorption capacity of TE6. Conversely, temperatures that are too low promote glycol loss due to foaming and increase the glycol's hydrocarbon uptake and potential still-vent emissions (Kirchgessner *et al.*, 2004).

Heat exchangers can be sized using conventional design procedures. Typical guidelines are as follows (Manning and Thompson, 1991).

1. Glycol–Glycol Exchanger

    a. Design duty is calculated as design requirement plus 5% for fouling and flow variations.
    b. The entering temperatures of the lean and rich glycol streams are known; a hot end (lean glycol in–rich glycol out) temperature approach of 60°F maximizes the preheat of the rich glycol.
    c. Two or more heat exchangers should be placed in series to avoid any temperature cross.

In smaller units, the exchanger may be replaced by a surge tank and heat-transfer coil. The shell volume is based on a 30-minutes retention time, an L/D ratio of 4, and a minimum size of D = 1.5 ft, L = 3.5 ft.

2. Lean Glycol/Dry Gas Exchanger

    a. The lean glycol outlet temperature should be 5–10°F hotter than the inlet gas temperature to the absorber. Therefore, the lean glycol is cooled from 180–200°F to 110–120°F. This may be accomplished in (1) a double pipe exchanger for smaller units (less than 25 MMscfd) and (2) an aerial, fin-fan exchanger or a water-cooled, shell and tube exchanger for larger units (greater than 25 MMscfd).
    b. The design heat duty must provide for a 5 to 10% allowance for fouling and flow variations.

9.3.2.6 Phase Separator (Flash Tank)

Many glycol dehydration units contain an emissions separator and a three-phase vacuum separator. An emissions separator removes dissolved gases from the warm rich glycol (about 90% of the methane and 10 to 40% of the VOCs entrained in the glycol) and reduces VOC emissions from the still. The wet or rich glycol is flashed at 50–100 psia and 100–150°F. The flash gas from the emissions separator can be used as supplemental fuel gas or as stripping gas on the reboiler. The wet glycol, largely depleted of methane and light hydrocarbons, flows to the glycol regenerator where it is heated to boil off the absorbed water, remaining methane, and VOCs. These gases are normally vented to the atmosphere and the lean glycol is circulated back to the gas contactor (EPA430-B-03-013, 2003).

A three-phase vacuum separator is desirable if liquid hydrocarbons are present. This allows the liquid hydrocarbon to be removed before it enters the still, where it could result in emissions or cause excess glycol losses from the still vent. Because the hydrocarbons are collected under a vacuum, they are stable and no vapor losses or weathering occurs (Kirchgessner *et al.*, 2004).

The design of the three-phase separator is similar to that of the two-phase separator except that it has a second control valve and liquid level controller to drain the accumulated hydrocarbon phase. Recommended liquid retention times are 5 to 10 minutes for two-phase (gas-glycol) and 20–30 minutes for three-phase (gas–liquid hydrocarbon–glycol) separation.

On glycol units that do not have a phase separator, the rich glycol is likely to have separate gas–glycol phases and be in plug or slug flow. This will likely bias any rich glycol samples collected and should be accounted for with an appropriate correction factor.

9.3.2.7 Glycol Circulation Pumps

A circulation pump is used to move the glycol through the unit. A wide variety of pump and driver types are used in glycol systems, including gas/glycol-powered positive displacement or glycol balance pumps (e.g., Kimray pumps) and electric motor-driven reciprocating or centrifugal pumps. The gas/glycol pump is common in field TEG dehydrators where electricity is typically not available. This type of pump uses the high-pressure glycol leaving the absorber to provide part of its required driving energy. Gas, taken under pressure from the absorber, is used to supply the remaining driving energy. This additional pump gas accounts for up to 8 scf/gal (depending on the absorber pressure) and should be recovered in the phase separator and used to the extent possible as fuel or stripping gas in the reboiler. Larger dehydrators in plants may use motor-driven pumps for higher efficiency. Where feasible, using electric pumps as alternatives to glycol balance pumps can yield significant economic and environmental benefits, including a financial return through reduced gas losses, increased operational efficiency, and reduced maintenance costs (EPA430-B-03-014, 2004).

A reciprocating pump is sized using manufacturers' catalogs or by the standard mechanical energy balance and an assumed pump efficiency of 70–80%. The temperature rise through the pump may be estimated by increasing the glycol enthalpy by the pump work. The BS&B Company (1960) recommends the following quick estimates based on 80% pump efficiency and 90% motor efficiency:

$$\text{Pump BHP} = 2 \times 10^{-7}(Q_{\mathrm{TEG}})(P) \tag{9-7}$$

$$\text{Electrical kW} = 1.833 \times 10^{-7}(Q_{\mathrm{TEG}})(P) \tag{9-8}$$

where Q_{TEG} is TEG circulation rate, gal/min; and P is system pressure, psig.

9.3.2.8 Filters

Two types of filters are commonly used in glycol systems. Fabric filters (e.g., sock) are used to remove particulate matter, and carbon filters are used to adsorb dissolved organic impurities from the glycol.

9.3.2.8.1 Fabric (Particulate) Filters The suspended solids content of glycol should be kept below 0.01 wt % to minimize pump wear, plugging of exchangers, fouling of absorber trays and stripper packing, solids deposition on the firetube in the reboiler, and glycol foaming. Solids filters are selected to remove particles with a diameter of 5 μm and larger. A common design is a 3-inch diameter by 36-inch-long cylindrical element in housing, sized for a flow rate of 1 to 2 gallon per minutes per element. The filters are sized for a 12- to 15-psi pressure drop when plugged.

The preferred location for solids filtration is on the high-pressure side at the bottom of the absorber. This filter will remove any foreign solid particles picked up in the absorber before entering the glycol pump. A low-pressure glycol filter can also be installed between the glycol/glycol exchanger and the reboiler for added filtration. An advantage of this placement is that the viscosity of the glycol is lower after it has been warmed in the glycol/glycol exchanger.

9.3.2.8.2 Carbon Filters Activated carbon filters are used to remove dissolved impurities in the glycol, such as high-boiling hydrocarbons, surfactants, well-treating chemicals, compressor lubricants, and TEG degradation products. The preferred type of carbon is the hard, dense, coal-based carbon. The carbon filter should be located downstream of the sock filter to prevent the carbon filter from becoming plugged with particles. The most common arrangement for a carbon filter is as a slipstream filter, with a minimum flow rate of 20% of the total flow. A preferred arrangement is to use a full-flow carbon filter vessel. Such a vessel is sized for a residence time of 15 to 20 minutes, with a superficial velocity of 2 to 3 gal/min/ft^2.

9.3.3 Operational Problems

The operating problems associated with each equipment in the TEG dehydration unit are described individually in the following sections (Ghoshal and Mukhopadhyay, 1993; Hubbard, 1993; Gas Technology Institute, 2001; Chakraborty and Bagde, 2004).

9.3.3.1 Absorber

The main operating problems associated with the absorber are insufficient dehydration, foaming, and hydrocarbon solubility in glycol, which are discussed next.

9.3.3.1.1 Insufficient Dehydration Causes of insufficient dehydration (i.e., wet sales gas) include excessive water content in the lean glycol, inadequate absorber design, high inlet gas temperature, low lean glycol temperature, and overcirculation/undercirculation of glycol. Lean glycol purity (i.e., glycol concentration) plays a vital role in the rate of absorption of moisture. A minimum lean glycol concentration is therefore needed to achieve a specified dew point depression. Higher water concentrations in the lean glycol result in poor dehydration.

The outlet gas moisture dew point indicates the performance of the absorber. Inadequate absorber design most often occurs when a glycol unit is moved from its original field. In fact, gases with the same flow rate can contain very different quantities of water, depending on the field temperatures, pressures, and gas composition. Therefore, when a glycol unit is moved to a different facility, the water load on the unit should be checked.

Temperature of the inlet gas dictates the amount of water fed into the unit, where a lower inlet gas temperature will require less water to be removed by the glycol. Hence, performance of the sweet gas cooler, as well as lean amine cooler of the gas sweetening unit, needs to be monitored.

Lean glycol temperature at the top of the absorber will affect the water partial pressure at the top stage, where high TEG temperature may cause high moisture content of the outlet gas. Reboiler temperature can therefore be increased up to 400°F above which glycol degradation starts.

9.3.3.1.2 Foaming Foaming causes glycol to be carried out of the absorber top with the gas stream, resulting in large glycol losses and decreased glycol unit efficiency. Foaming can normally be traced to mechanical or chemical causes. High gas velocity is usually the source of mechanical entrainment. At excessive velocities, glycol can be lifted off the trays and out of the vessel with the gas. High velocity can be caused by poor design, operating at gas flow rates above design levels, or damaged/plugged trays/packing. Although an efficient demister pad is normally fitted at the top of the absorber, extremely high gas velocities may carry glycol through the demister.

Chemical foaming is caused by contaminants in the glycol, liquid hydrocarbons, well-treating chemicals, salts, and solids. Adequate inlet separation and filtration systems (cartridge filter and activated carbon bed) are therefore needed to prevent foaming due to chemical contamination. The filters are generally effective until they become plugged by particulate matter (indicated by a high pressure drop across the filter) or saturated with hydrocarbons; thus, the only operational issue is the filter replacement frequency.

9.3.3.1.3 Hydrocarbon Solubility in TEG Solution Aromatic hydrocarbon solubility in glycols is a significant issue in gas dehydration technology due to the potential release of aromatics to the atmosphere at the regenerator. In fact, in the absorber, TEG can absorb significant amounts of aromatic components in the gas (benzene, toluene, ethyl benzene, and xylene), which are often released to the atmosphere at the regenerator. While these emissions are generally small on a mass basis, they have received a great deal of attention from environmental and safety regulatory agencies. Several BTEX emission mitigation methods have been proposed. By far the most common method implemented to date is condensation of the BTEX components in the regenerator overhead and subsequent separation from the condensed water. This scheme is simple and relatively low cost, although it does complicate water disposal due to the high solubility of BTEX in water.

9.3.3.2 Still (Stripper)

The major operational problem with the still is excessive glycol losses due to vaporization. The TEG concentration in the vapor (and thus glycol vaporization losses) increases significantly above 250°F. The appearance of the plume leaving the glycol still can identify excess glycol vaporization. Because glycol is heavier than air, a plume sinking to the ground instead of rising indicates that glycol is being vaporized. Excessive glycol vaporization is more of a problem for finned atmospheric condensers than for water-cooled or glycol-cooled condensers. Although finned atmospheric condensers are simple and inexpensive, they are sensitive to extremes in ambient temperature. For example, during cold winter periods, a low temperature at the top of the still column causes excessive condensation and floods the reboiler. This prevents adequate regeneration of the glycol and reduces the potential dew point depression of the glycol, causing insufficient dehydration in the absorber. Also, excessive glycol losses may occur as the reboiler pressure increases and blows the liquids out the top of the column. During the summer months, inadequate cooling may allow excessive glycol vaporization losses.

9.3.3.3 Reboiler

Operational problems associated with the reboiler include salt contamination, glycol degradation, and acid gas-related problems.

9.3.3.3.1 Salt Contamination Carry over of brine solutions from the field can lead to salt contamination in the glycol system. Sodium salts (typically sodium chloride, NaCl) are a source of problems in the reboiler, as NaCl is less soluble in hot TEG than in cool TEG; NaCl will precipitate from the solution at typical reboiler temperatures of 350–400°F. The salt can deposit on the fire tube, restricting heat transfer. If this occurs, the surface temperature of the fire tube will increase, causing hot spots and increased thermal degradation of the glycol. The deposition of salt may also result in corrosion of the fire tube. Dissolved salts cannot be removed by filtration. As a general rule, when the salt content reaches 1%, the glycol should be drained and reclaimed. If the level of salts is allowed to increase beyond 1%, both severe corrosion and thermal degradation threaten the system.

9.3.3.3.2 Glycol Degradation Glycol degradation is caused primarily by either oxidation or thermal degradation. Glycol readily oxidizes to form corrosive acids. Oxygen can enter the system with incoming gas, from unblanketed storage tanks or sumps, and through packing glands. Although oxidation inhibitors [such as a 50–50 blend of monoethanolamine (MEA) and 33% hydrazine solution] can be used to minimize corrosion, a better approach to controlling oxidation is blanketing the glycol with natural gas, which can be applied to the headspace in storage tanks and any other area where glycol may contact oxygen. Thermal degradation of the glycol results from the following conditions: high reboiler temperature, high heat flux rate, and localized overheating. The reboiler temperature should always be kept below 402°F to prevent degradation, and a good fire tube design should inherently prevent a high heat flux rate. Localized overheating can be caused by deposits of salts or hydrocarbons. In addition, thermal degradation of the glycol produces acidic degradation products that lower the pH and increase the rate of degradation, creating a destructive cycle.

9.3.3.3.3 Acid Gas Some natural gas contains H_2S and/or CO_2, and these acid gases may be absorbed in the glycol. Acid gases can be stripped in the reboiler and still. Mono-, di-, or triethanolamine may be added to the glycol to provide corrosion protection from the acid gases.

9.3.3.4 Surge Tank

When surge tanks also serve as glycol/glycol heat exchangers, the level must be monitored to ensure that the lean glycol covers the rich glycol coil.

Otherwise, inadequate heat exchange will occur, and the lean glycol will enter the absorber at an excessively high temperature.

9.3.3.5 Heat Exchanger

The primary operational problem with heat exchangers is poor heat transfer, which results in lean glycol that is too warm. When this occurs, poor dehydration and insufficient dew point depression can result. Also, glycol vaporization losses to the product gas may be higher with increased lean glycol temperature. Poor heat transfer and the resulting high lean glycol temperature are usually caused by fouled heat exchangers, undersized heat exchangers, or overcirculation. Exchangers may be fouled by deposits such as salt, solids, coke, or gum. In the case of undersized exchangers, additional heat exchangers may be required. Corrosion of the coil in surge tank heat exchangers can also present operating problems, as it can lead to cross-contamination of rich and lean glycol.

9.3.3.6 Phase Separator (Flash Tank)

Inadequate residence time in the phase separator may result in a large quantity of glycol being included in the hydrocarbon stream and vice versa. This is most likely a result of overcirculation of the glycol. The effects of hydrocarbons and overcirculation are discussed next.

9.3.3.7 Glycol Circulation Pump

Major problems associated with the circulation pump and rates are related to reliability, pump wear, and overcirculation or undercirculation.

9.3.3.7.1 Reliability Pump reliability is important because pumps are the only moving parts in the entire dehydrator system. It is good design practice to include a strainer or sock filter in the pump suction line to prevent damage by foreign objects. Pump reliability is also enhanced by limiting the lean glycol temperature to 180 to 200°F and ensuring good filtration. Pump wear, leakage, and failures increase if the glycol becomes dirty or hot. In severe cases, glycol losses of as much as 35 gal/day from seal leakage have been observed.

9.3.3.7.2 Pump Wear As the O-rings and seals on the glycol balance pumps wear out, there is the potential for contamination of the lean glycol by the rich glycol. This increases the water content of the lean glycol and may cause (1) the gas to no longer be dried to pipeline specifications and/or (2) the operator to increase the glycol circulation rate (and therefore the emissions) in an effort to compensate for the wetter lean glycol. Because of the leakage, it may also be difficult to determine an accurate glycol circulation rate.

9.3.3.7.3 Over Circulation/Under Circulation Excessively high glycol circulation rates can lead to many problems. If the unit is overcirculating the glycol, the lean glycol may have insufficient heat exchangers to be cooled properly, and the resulting hot lean glycol may not achieve the desired water removal rate. A high circulation rate may not allow adequate residence time in the phase separator for the hydrocarbons to be removed, which may lead to hydrocarbon deposits, glycol losses, foaming, and emissions. Excessive glycol circulation rates can also result in increased sensible heat requirements in the reboiler. Also, because emissions are proportional to the circulation rate, overcirculation results in greater VOCs emissions.

Undercirculating the glycol provides an insufficient quantity of glycol in the absorber for the quantity of water to be removed and results in wet sales gas.

Considering the aforementioned matters, the glycol flow rate should be optimized, where it can be done by checking the treated gas moisture dewpoint.

9.4 SOLID DESICCANT DEHYDRATION

Solid desiccant dehydration systems work on the principle of adsorption. Adsorption involves a form of adhesion between the surface of the solid desiccant and the water vapor in the gas. The water forms an extremely thin film that is held to the desiccant surface by forces of attraction, but there is no chemical reaction.

Solid desiccant dehydrators are typically more effective than glycol dehydrators, as they can dry a gas to less than 0.1 ppmV (0.05 lb/MMcf). However, in order to reduce the size of the solid desiccant dehydrator, a glycol dehydration unit is often used for bulk water removal. The glycol unit would reduce the water content to around 60 ppmV, which would help reduce the mass of solid desiccant necessary for final drying.

Using desiccant dehydrators as alternatives to glycol dehydrators can yield significant economic and environmental benefits, including reduced capital cost, reduced operation and maintenance cost, and minimal VOC and hazardous air pollutants (BTEX). A detailed discussion on determining their economics and environmental benefits can be found in EPA430-B-03-016 (2003).

9.4.1 Desiccant Capacity

The capacity of a desiccant for water is expressed normally in mass of water adsorbed per mass of desiccant. The dynamic moisture sorption capacity of a desiccant will depend on a number of factors, such as the relative humidity of the inlet gas, the gas flow rate, the temperature of the adsorption zone, the mesh size of the granule, and the length of service and degree of contamination of the desiccant (Huntington, 1950) and not the least on the desiccant itself. The moisture sorption capacity is not affected by variations in pressure, except where pressure may affect the other variables listed previously. There are three capacity terms used (Campbell, 1992).

- *Static equilibrium capacity*: The water capacity of new, virgin desiccant as determined in an equilibrium cell with no fluid flow (corresponding to the adsorption isotherm).
- *Dynamic equilibrium capacity*: The water capacity of desiccant where the fluid is flowing through the desiccant at a commercial rate.
- *Useful capacity*: The design capacity that recognizes loss of desiccant capacity with time as determined by experience and economic consideration and the fact that all of the desiccant bed can never be fully utilized.

9.4.2 Desiccant Selection

A variety of solid desiccants are available in the market for specific applications. Some are good only for dehydrating the gas, whereas others are capable of performing both dehydration and removal of heavy hydrocarbon components. The selection of proper desiccant for a given application is a complex problem. For solid desiccants used in gas dehydration, the following properties are desirable (Campbell, 1992; Daiminger and Lind, 2004).

(1) High adsorption capacity at equilibrium. This lowers the required adsorbent volume, allowing for the use of smaller vessels with reduced capital expenditures and reduced heat input for regeneration.
(2) High selectivity. This minimizes the undesirable removal of valuable components and reduces overall operating expenses.
(3) Easy regeneration. The relatively low regeneration temperature minimizes overall energy requirements and operating expenses.
(4) Low pressure drop.
(5) Good mechanical properties (such as high crush strength, low attrition, low dust formation, and high stability against aging). These factors lower overall maintenance requirements by reducing the frequency of adsorbent change out and minimizing downtime-related losses in production.
(6) Inexpensive, noncorrosive, nontoxic, chemically inert, high bulk density and no significant volume changes upon adsorption and desorption of water.

The most common commercial desiccants used in dry bed dehydrators are silica gel (i.e., Sorbead), molecular sieves, and activated alumina.

Silica gel (a generic name for a gel manufactured from sulfuric acid and sodium silicate) is a widely used desiccant, which can be used for gas and liquid dehydration and hydrocarbon recovery from natural gas. It is characterized by the following.

- Best suited for normal dehydration of natural gas.
- Easily regenerated than molecular sieves.
- Has high water capacity, where it can adsorb up to 45% of its own weight in water.
- Costs less than molecular sieve.
- Capable of dew points to −140°F.

Silica gel used for natural gas drying should be of the Sorbead type because this is the water-stable silica gel type. Most other silica gel types will produce fines in contact with water. Engelhard Sorbead is a high-performance, extremely robust silica gel adsorbent used primarily for the control of hydrocarbon dew point in natural gas. However, it can also be used for dehydration only; – its main benefit then is a longer lifetime. High adsorption capacity, drying performance, and low dew points (−158°F) are characteristic of Sorbead. Sorbead adsorbents come in a range of sizes and physical characteristics to fit any manufacturing environment. Their longer life reduces operating costs while their high performance enhances the operating safety of natural gas treatment plants, among others.

Molecular sieves are crystalline alkali metal alumino silicates comprising a three-dimensional interconnecting network of silica and alumina tetrahedral. The structure is an array of cavities connected by uniform pores with diameters ranging from about 3 to 10°A, depending on the sieve type. A detailed discussion on different types of molecular sieves and their applications is given by Bruijn *et al*. (2002) and Meyer (2005). A molecular sieve is the most versatile adsorbent because it can be manufactured for a specific pore size, depending on the application. It is:

- Capable of dehydration to less than 0.1 ppm water content.
- The overwhelming choice for dehydration prior to cryogenic processes (especially true for LNG).
- Excellent for H_2S removal, CO_2, dehydration, high temperature dehydration, heavy hydrocarbon liquids, and highly selective removal.
- More expensive than silica gel, but offers greater dehydration.
- Requires higher temperatures for regeneration, thus has a higher operating cost.

A molecular sieve dehydration system is also an alternative to the Drizo process. However, because of multiple high pressure and high temperature vessels, the installed cost of a molecular sieve system is two or three times more than an equivalent Drizo system (Dow Chemical Co., 1985).

There are several types of alumina available for use as a solid desiccant. Activated alumina is a manufactured or natural occurring form of aluminum oxide that is activated by heating. It is widely used for gas and liquid dehydration and will produce a dew point below −158°F if applied properly. Less heat is required to regenerate alumina than for molecular sieve, and the regeneration temperature is lower. However, molecular sieves give lower outlet water dew points (GPSA, 1998).

It should be noted that no desiccant is perfect or best for all applications. In some applications the desiccant choice is determined primarily by economics. Sometimes the process conditions control the desiccant choice. If a unit is designed properly it is quite rare that desiccants can be interchangeable. What is often possible is to replace within one class of adsorbents, i.e., molecular sieve of one supplier with molecular sieve from another.

9.4.3 Process Description

The flow scheme of a typical solid adsorbent-based dehydration process is shown in Figure 9-7. The process is conducted alternately and periodically, with each bed going through successive steps of adsorption and desorption. During the adsorption step, the gas to be processed is sent on the adsorbent bed, which selectively retains the water. When the bed is saturated, hot gas is sent to regenerate the adsorbent. After regeneration and before the adsorption step, the bed must be cooled. This is achieved by passing through cold gas. After heating, the same gas can be used for regeneration. In these conditions, four beds are needed in practice in cyclic operation to dry the gas on a continuous[4] basis: two beds operating simultaneously in adsorption or gas drying cycle, one bed in the cooling cycle, and one bed in the regeneration cycle. In the simplest case

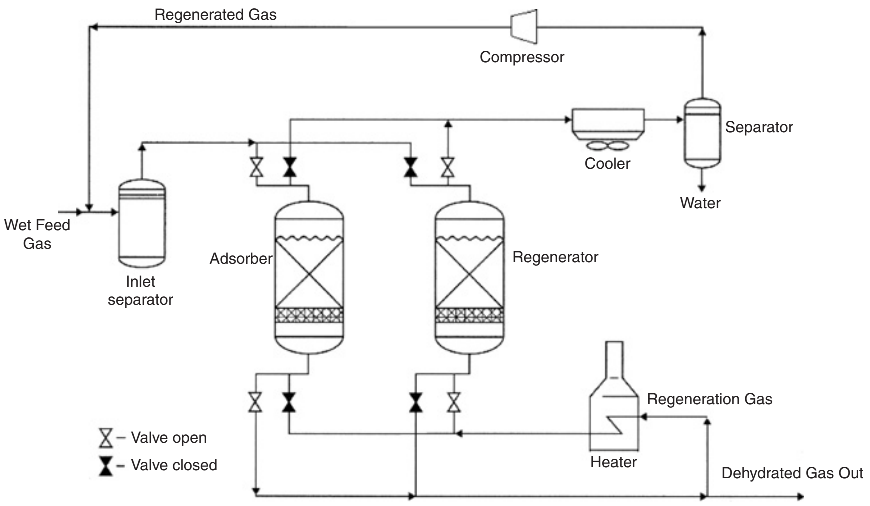

Figure 9-7. Typical flow scheme for a solid desiccant dehydration plant (GPSA, 1998).

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[4] Adsorbents are normally unsuitable for continuous circulation due to mechanical problems and also due to the risks of attrition.

(as shown in Figure 9-7), one bed operates in adsorption, while the second operates in desorption, and both beds are switched periodically (Rojey *et al*., 1997).

In the gas drying cycle, the wet inlet gas first passes through an inlet separator where free liquids, entrained mist, and solid particles are removed. This is a very important part of the system because free liquids can cause the adsorbent materials to break down. This leads to higher pressure drop and channeling, reducing the overall performance of the unit. If the adsorption unit is downstream from an amine unit, glycol unit, or compressors, a filter separator is required.

In addition to the use of an inlet separator to maximize water droplet removal, a guard layer (equal to about 10–20% of bed volume) of specialized, water-stable adsorbent can be added on top of the main adsorbent bed. This water stability can be imparted using tempering (i.e., heating to high temperatures over a long time), however, this process reduces water adsorption capacity greatly (Daiminger and Lind, 2004).

In the adsorption cycle, the wet inlet gas flows usually downward through the tower. The adsorbable components are adsorbed at rates dependent on their chemical nature, the size of their molecules, and the size of the pores in the solid material. The water molecules are adsorbed first in the top layers of the desiccant bed. Dry hydrocarbon gases are adsorbed throughout the bed. As the upper layers of desiccant become saturated with water, the water in the wet gas stream begins displacing the previously adsorbed hydrocarbons in the lower desiccant layers. Liquid hydrocarbons will also be absorbed and will fill pore spaces that would otherwise be available for water molecules. For each component in the inlet gas stream, there will be a section of bed depth, from top to bottom, where the desiccant is saturated with that component and where the desiccant below is just starting to adsorb that component. The depth of bed from saturation to initial adsorption is known as the mass transfer zone (MTZ). This is simply a zone or section of the bed where a component is transferring its mass from the gas stream to the surface of the desiccant. In the mass transfer zone, the water content of the gas is reduced from saturation to less than 1 ppm (GPSA, 1998). As the flow of gas continues, the mass transfer zone moves downward through the bed and water displaces the previously adsorbed gases until finally the entire bed is saturated with water vapor. When the leading edge of the MTZ reaches the end of the bed, breakthrough occurs. If the entire bed becomes completely saturated with water vapor, the outlet gas is just as wet as the inlet gas. Obviously, the towers must be switched from the adsorption cycle to the regeneration cycle (heating and cooling) before the desiccant bed is completely saturated with water.

At any given time, at least one of the towers will be adsorbing while the other towers will be in the process of being heated or cooled to regenerate the desiccant. When a tower is switched to the regeneration cycle, some wet gas (i.e., the inlet gas downstream of the inlet gas separator) is heated to temperatures of 450 to 600°F in the high-temperature heater and routed to the tower to remove the previously adsorbed water. As the temperature within the tower is increased, the water captured within the pores of the desiccant turns to steam and is absorbed by the natural gas. This gas leaves the top of the tower and is cooled by the regeneration gas cooler. When the gas is cooled the saturation level of water vapor is lowered significantly and water is condensed. The water is separated in the regeneration gas separator and the cool, saturated regeneration gas is recycled to be dehydrated. This can be done by operating the dehydration tower at a lower pressure than the tower being regenerated or by recompressing the regeneration gas. Once the bed has been “dried” in this manner, it is necessary to flow cool gas through the tower to return it to normal operating temperatures (about 100 to 120°F) before placing it back in service to dehydrate gas. The cooling gas could either be wet gas or gas that has already been dehydrated. If wet gas is used, it must be dehydrated after being used as cooling gas, where a hot tower will not sufficiently dehydrate the gas.

The switching of beds is controlled by a time controller that performs switching operations at specified times in the cycle. The length of the different phases can vary considerably. Longer cycle times will require larger beds and additional capital investment, but will increase the bed life. A typical two-bed cycle might have an 8-hour adsorption period with 6 hours of heating and 2 hours of cooling for regeneration. The 16-hour adsorption time for an adsorption unit with three beds, two beds in adsorption and one bed in regeneration, makes also a full cycle time of 24 hours, which gives a good 3-year guarantee.

Internal or external insulation for the adsorbers may be used. The main purpose of internal insulation is to reduce the total regeneration gas requirements and costs (invest cost, however, is higher). Internal insulation eliminates the need to heat and cool the steel walls of the adsorber vessel. Normally, a castable refactory lining is used for internal insulation. The refractory must be applied and cured properly to prevent liner cracks. Liner cracks will permit some of the wet gas to bypass the desiccant bed.

Only a small amount of wet, bypassed gas is needed to cause freeze ups in cryogenic plants. Ledges installed every few feet along the vessel wall can help eliminate this problem.

9.4.4 Design Considerations

The following considerations are a good approximation for estimation of the solid desiccant dehydration behavior. This information serves only as a basis for performing preliminary design calculations based on a given cycle length, number of vessels and their configuration, and a given desiccant. Therefore, it is highly recommended to refer to desiccant vendors (e.g., CECA, 1984; Grace, 1988; Union Carbide, 1988; Zeochem, 1989) for designing a solid desiccant dehydration unit, as the useful capacity of a desiccant is highly dependent on its aging behavior. To take aging into account, experience is very important and no published correlations exist as every desiccant vendor regards that as their own intellectual property and know how. To use only literature data will result in either uneconomical units or potentially nonworking units.

9.4.4.1 Allowable Gas Velocity

Generally, as the gas velocity during the drying cycle decreases, the ability of the desiccant to dehydrate the gas increases (there is certainly an upper limit on space velocity due to adsorption kinetics, but more often other considerations such as fluidization of the bed restrict upper velocities). At lower actual velocities, drier effluent gases will be obtained. Consequently, it would seem desirable to operate at minimum velocities to fully use the desiccant. However, low velocities require towers with large cross-sectional areas to handle a given gas flow and allow the wet gas to channel through the desiccant bed with incomplete dehydration. In selecting the design velocity, therefore, a compromise must be made between the tower diameter and the maximum use of the desiccant. Figure 9-8 shows a maximum allowable superficial gas velocity for mole sieve materials. Smaller velocities may be required due to pressure drop considerations.

An alternative method for determining superficial gas velocity in a molecular sieve bed is using the Ergun (1952) equation, which relates ΔP to V_{SG}, μ, ρ, and desiccant size, as follows:

$$\frac{\Delta P}{L} = B\mu V_{SG} + C\rho_G V_{SG}^2 \tag{9-9}$$

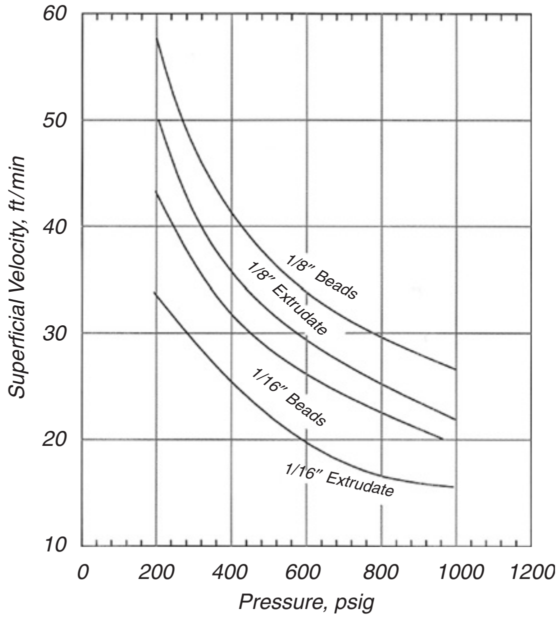


Figure 9-8. Maximum allowable gas superficial velocity for mole sieve dehydration (GPSA, 1998).

where $\Delta P/L$ is pressure drop per length of bed, psi/ft; μ is gas viscosity, cP; ρ_G is gas density, lb/ft^3; and V_{SG} is superficial gas velocity, ft/min. Constants for Equation (9-9) are given in Table 9-1 for mole sieve materials.

An important observation from Equation (9-9) is that the higher the gas superficial velocity, the greater the pressure drop across the bed. The design gas velocity is therefore a trade-off between the maximum gas velocity and the acceptable pressure drop. The design pressure drop across

Table 9-1
Parameters Used in Equation (9-9)

Particle type	Coefficients	
	B	C
1/8-in. beads	0.0560	0.0000889
1/8-in. extrudate	0.0722	0.000124
1/16-in. beads	0.152	0.000136
1/16-in. extrudate	0.238	0.000210

the entire bed should be about 5 psi; values higher than approximately 8 psi are not recommended (GPSA, 1998). Most designs are based on a $\Delta P/L$ of about 0.31–0.44 psi/ft and typical superficial gas velocities of 30–60 ft/min (Grace, 1988; Coker, 1994).

9.4.4.2 Bed Length to Diameter Ratio

Once the superficial gas velocity is determined, then the diameter and length of the bed can be calculated from geometry of the adsorber. In its simplest form, an adsorber is normally a cylindrical tower filled with a solid desiccant. The depth of the desiccant may vary from a few feet to 30 ft or more. The minimum bed internal diameter for a specified superficial gas velocity is given by the following equation (Ergun, 1952):

$$D^2 = \frac{25(Q_{\mathrm{G}})(T)(Z)}{(P)(V_{\mathrm{SG}})} \tag{9-10}$$

where D is bed diameter, ft; Q_{G} is gas flow rate, MMscfd; T is inlet gas temperature, °R; P is inlet gas pressure, psia; Z is compressibility factor; and V_{SG} is superficial gas velocity, ft/min.

Also, the bed length, L_{B}, can be determined by the following equation (Collins, 1967):

$$L_{\mathrm{B}} = \frac{127.3(W)}{(\rho_{\mathrm{b}})(D^2)(X)} \tag{9-11}$$

where L_{B} is bed length, ft; W is weight of water adsorbed, lb per cycle; ρ_{b} is bulk density of desiccant, lb/ft^3; and X is maximum desiccant useful capacity, lb water/100 lb desiccant.

A bed length to diameter ratio of higher than 2.5 is desirable. A ratio as low as 1:1 is sometimes used; however, poor gas dehydration, caused by nonuniform flow, channeling and an inadequate contact time between the wet gas and the desiccant sometimes result.

9.4.4.3 Desiccant Capacity

The maximum desiccant useful capacity can be calculated as Equation (9-12). However, this is an empirical equation that does not accurately encompass all the factors affecting the mass transfer zone. It is reasonable only within a limited range of pressure, temperature, aging mechanisms, compositions, and so on (Campbell, 1992).

$$(X)(L_B) = (X_S)(L_B) - (0.45)(L_Z)(X_S) \tag{9-12}$$

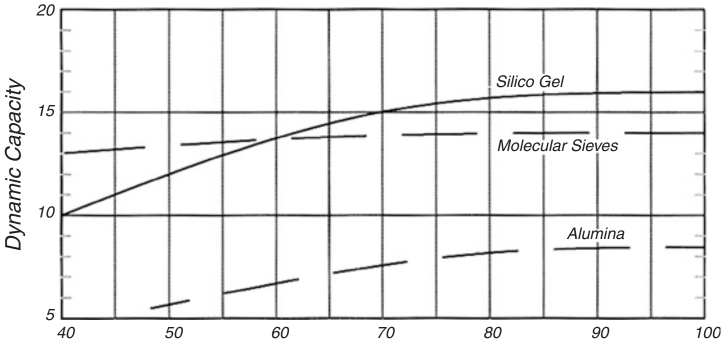
where X is desiccant useful capacity, lb water per 100-lb desiccant; X_S is dynamic capacity at saturation, lb water per 100-lb desiccant; L_Z is MTZ length, ft; and L_B is bed length, ft.

The numerical value of (0.45) in Equation (9-12) is an average number based on test. It is a function of MTZ length but only varies from 0.40 to 0.52 in a wide range of applications. The value used is the mode of the distribution curve for most services. The dynamic capacity "X_S" must reflect desiccant aging and other such factors. It is the effective capacity of the desiccant for water behind the adsorption zone. Because desiccant degrades in service, the value used must reflect a capacity at some future time to optimize desiccant replacement cost. Figure 9-9a may be used to predict "X_S" in Equation (9-12) as a function of relative saturation. Remember that relative gas saturation is fixed by the last pressure and temperature at which liquid water was present — the last separator producing any water or the reservoir. If any water is drained off the inlet scrubber, the inlet gas is saturated. If not, the saturated water content at the previous contact divided by that at the dehydrator gives the relative saturation (RS) value as a fraction.

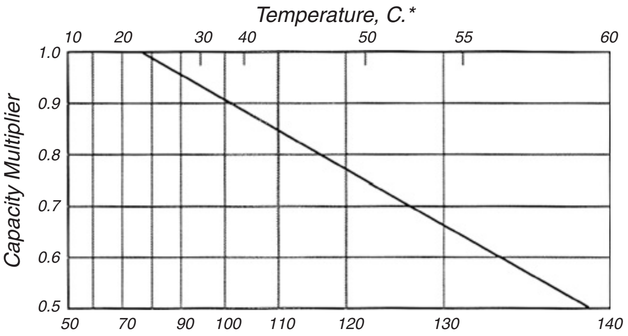
The values in Figure 9-9a are lower than the theoretical numbers published for air. They are for natural gas and reflect the competition of the hydrocarbons for adsorption surface and expected dynamic saturation after a short-time use in an actual plant with normal degradation only. For gels and aluminas, the values in Figure 9-9a must be corrected for temperature. No temperature correction is needed for molecular sieves in the temperature range shown. The value from Figure 9-9b is multiplied by that in Figure 9-9a to obtain the "X_S" for use in Equation (9-12). The "X" obtained from Equation (9-12) will be the useful capacity of virgin, activated desiccant. This is greater than the useful capacity after degradation. In effect, Equation (9-12) simply corrects for that amount of bed unused because of zone length.

9.4.4.4 MTZ Length

The MTZ length, L_{MTZ}, depends on gas composition, flow rate, RS of the water in the gas, and the loading capability of the desiccant. Pressure



(a) Relative Water Saturation (R.S) – %



(b) Temperature– °F,

Figure 9-9. Effect of relative saturation and temperature on the dynamic capacity of desiccants (NATCO, 1987).

has only a small effect, particularly above 300 psia. For silica gel, the MTZ length may be estimated from the following equation (Simpson and Cummings, 1964):

$$L_{\mathrm{MTZ}} = 375\left[\frac{m_{\mathrm{w}}^{0.7895}}{V_{\mathrm{SG}}^{0.5506}(\mathrm{RS})^{0.2646}}\right] \tag{9-13}$$

where L_{MTZ} is MTZ length, inch; m_{w} is water loading, lb/(hr.ft^2); V_{SG} is gas superficial velocity, ft/min; RS is percentage relative saturation of inlet gas.

The values of "L_{MTZ}" from Equation (9-13) are those quoted from Simpson and Cummings (1964) for air drying. The values, however, are useful for natural gas drying. For alumina and molecular sieves, the zone length determined using Equation (9-13) is multiplied by 0.8 and 0.6, respectively. Shorter zones are obtained with these materials because they have less capacity for hydrocarbon.

The relevant equation for calculation of the water loading (m_w) can be written as follows (Ledoux, 1948):

$$m_\mathrm{w} = 0.053\left[\frac{Q_\mathrm{G}(W)}{D^2}\right] \tag{9-14}$$

where Q_G is gas flow rate, MMscfd; D is bed diameter, ft; and W is water content of gas, lb/MMscf. This is the water loading on a mass basis. Equation (9-14) is merely a conversion from water content per standard volume divided by bed cross-sectional area.

9.4.4.5 Breakthrough Time

The breakthrough time for the water zone formed, t_b in hours, can be estimated as follows (McCabe *et al*., 1985):

$$t_\mathrm{b} = \frac{(0.01)(X)(\rho_\mathrm{b})(L_\mathrm{B})}{m_\mathrm{w}} \tag{9-15}$$

For a given set of gas flow conditions, water loading, cycle length, and tower configuration, one can size the desiccant bed. However, as stated earlier, desiccant vendors typically perform these kinds of calculations and will even guarantee. This is important today as design is becoming more effective and incorrect assumptions might lead to misguided process choice.

9.4.5 Operational Problems

Operational problems that may occur because of poor design, operation, and maintenance in a solid desiccant unit are described in this section (Manning and Thompson, 1991).

9.4.5.1 Bed Contamination

The most frequent cause is incomplete removal of contaminants in the inlet gas separator. Also, if the regeneration gas leaving the separator is commingled with the feed gas to the dehydrators, then a separator malfunction can dump liquid hydrocarbons and water onto the desiccant. Regeneration separators should usually be equipped with filtration levels similar to the inlet gas to prevent recontamination.

9.4.5.2 High Dew Point

High dew point is one of the two common problems that can cause operating trouble. Possible causes include the following.

1. "Wet" inlet gas bypasses the dehydrator through cracks in the internal insulation. Cracks in a liner or in sprayed-on insulation can be detected by "hot spots" and peeling paint on the outer shell. Other symptoms are fast water breakthrough and an unusually rapid rise in the effluent gas temperature during regeneration.
2. Leaking valves also permit wet gas to bypass the dehydrators. Even a slight leak of hot gas usually produces a detectable temperature rise in what should be the cold side of the valve. Ultrasonic translators are also useful.
3. Incomplete desiccant regeneration will lead to a sudden loss in adsorption capacity and a significantly premature breakthrough. To be sure to well regenerate the adsorbents the inlet and outlet temperatures of the adsorber in regeneration should be analyzed. At the end of the heating step the outlet temperature should be almost constant during a certain time (30 minutes to 2 hours) depending on the design of the adsorber, and the temperature difference between inlet and outlet should not be more than 59–68°F depending on the quality of the heat insulation.
4. Excessive water content in the wet feed gas due to increased flow rate, higher temperatures, and lower pressure. It is very important to respect the inlet temperature (feed temperature) of the adsorbers in case of saturated gas. Small variations in temperature will lead to significant increases in the water content.

9.4.5.3 Premature Breakthrough

Satisfactory dew points are observed at the beginning but not for the entire duration of the drying cycle. Desiccant capacity should decrease with use but should stabilize at 55–70% of the initial capacity (Ballard, 1983). However, premature symptoms of "old age" are caused by an unrecognized increase in inlet water loading, an increase in heavy hydrocarbons

(C_4^+) in feed gas, methanol vapor in feed, desiccant contamination, or incomplete regeneration.

9.4.5.4 Hydrothermal Damaging

Heating up the adsorber without using a heating ramp or an intermediate heating step leads to a strong temperature difference in the vessel. At the bottom, the molecular sieve will be very hot and will desorb rapidly the adsorbed water while the layers at the top of the adsorber will be still at adsorption temperature. The water desorbed in the bottom layer will condense in the top layer. This phenomenon is called refluxing. The heating going on will heat up the liquid water and boil the molecular sieves in liquid water. Hydrothermal damaging will appear in consequence, which is different depending on the type of molecular sieve. In order to prevent hydrothermal damaging of molecular sieves it is not only important to choose the right formulation of the molecular sieve (binder and zeolite) but the operating conditions; the regeneration conditions should be determined carefully (Meyer, 2003). In fact, the higher the regeneration temperature and the higher the amount of liquid water present on the sieves, the heavier the damaging of the molecular sieves. In an industrial unit it is also important to limit the quantity of water appearing in liquid phase (condensing water due to oversaturation of the gas phase), as this decreases the temperature where hydrothermal destruction may occur with the water acting as a stabilizer for intermediates formed by dissolution of the zeolites (Suckow *et al*., 1992).

9.4.5.5 Liquid Carryover

Liquid (particularly amines) carryover in the molecular sieve bed has a negative impact on the drying process (i.e., poor gas flow distribution due to cake or fines formation as a consequence of chemical attack causing an increase of the pressure drop and a decrease of adsorption time). In order to reduce the liquid carryover in adsorbers, separators must be modified to improve their efficiency. The regeneration procedure should also be changed so as to have a moderate temperature increase to avoid water recondensation. Although on-site mechanical changes of the drying unit can improve the performance, however, the right corrective action can be found by using a more resistant molecular sieve, i.e., SRA. The SRA adsorbent offers a better mechanical resistance than the regular one to severe operating conditions simulating the thermal regeneration step of a natural gas purification unit (Le Bec *et al.*, 1998).

9.4.5.6 Bottom Support

Sometimes operators have problems with the support grid and leakage of molecular sieves through the support grid. As a result, they have to replace the whole bed. Important point here is the good mechanical design of the support bed, putting three wire mesh on the support grid (4, 10, 20 mesh) and installing the correct quantity and size of ceramic balls.

REFERENCES

Ballard, D., How to operate a glycol plant. *Hydrocarb. Proc.* **45**(6), 171–188 (1966).

Ballard, D., “The Fundamentals of Gas Dehydration.” Gas Conditioning Conference, University of Oklahoma, Norman, OK (1979).

Ballard, D., “How to Improve Cryogenic Dehydration.” Petroenergy, 83, Houston, TX (Sept. 12–16, 1983).

Bruijn, J.N.H., *et al.*, “Maximizing Molecular Sieve Performance in Natural Gas Processing.” Paper presented at the 81st Annual GPA Convention, Dallas, TX (March 11–13, 2002).

BS&B Engineering Company, Inc., Dehydrate (conc, glycol). *Hydrocarb. Proc.* **69**(4), 76 (1960).

BS&B, “Product Section 34:00 Dehydrators.” Black, Sivalls & Bryson, Inc., Oklahoma City, OK (1960).

Caldwell, R.E., “Glycol Dehydration Manual.” NATCO Group, Tulsa, OK (Jan 30, 1976).

Campbell, J.M., “Gas Conditioning and Processing,” 3rd Ed. Campbell Petroleum Series, Norman, OK (1992).

Carroll, J.J., “Natural Gas Hydrates, A Guide for Engineers.” Gulf Professional Publishing, Amsterdam, The Netherlands (2002).

CECA, SILIPORITE Molecular Sieves, Product Brochure, Paris La Defense, France (1984).

Chakraborty, A., and Bagde, A., “Operational Problems in Gas Sweetening and Dehydration: A Reckoner.” Paper presented at the GPA Europe February Meeting, Hammersmith, London, UK (2004).

Coker, A.K., Program sizes solid desiccant dryer for natural gas. *Oil Gas J.* **92**(7), 74–78 (1994).

Collins, J.J., AIChE Symp. Ser., No. 74, 63: 31 (1967).

Daiminger, U., and Lind, W, Adsorption-based processes for purifying natural gas. *World Refining* **14**(7), 32–37 (2004).

Dow Chemical Co., "Economic Comparison of WT-2, TEG with Stripping Gas and Molecular Sieves." Sales presentation (1985).

EPA430-B-03-016, "Replacing Glycol Dehydrators with Desiccant Dehydrators." Lessons Learned, Natural Gas Star Program, U.S. Environmental Protection Agency (EPA), Washington, DC (2003).

EPA430-B-03-013, "Optimize Glycol Circulation and Install Flash Tank Separators in Glycol Dehydrators." Lessons Learned, Natural Gas Star Program, U.S. Environmental Protection Agency (EPA), Washington, DC (2003).

EPA430-B-03-014, "Replace Gas-Assisted Glycol Pumps with Electric Pumps." Lessons Learned, Natural Gas Star Program, U.S. Environmental Protection Agency (EPA), Washington, DC (2004).

Ergun, S., Fluid flow through packed columns. *Chem. Eng. Prog*. **48**, 2 (1952).

Ghoshal, P.C., and Mukhopadhyay, S., "Improving Glycol Dehydration Unit Efficiency." Paper presented at the 2nd Technical Meeting of Gas Processors Association GCC Chapter, Bahrain (Oct. 27, 1993).

GPSA Engineering Data Book, 11th Ed. Gas Processors Suppliers Association, Tulsa, OK (1998).

GTI, "Glycol Dehydrator Operational Problems." Gas Technology Institute Publications, Des Plaines, IL (2001).

Grace, "Molecular Sieves." Davison Chemical Division, W.R. Grace & Co., Baltimore, MD (1988).

Hubbard, R., "Recent Developments in Gas Dehydration and Hydrate Inhibition." Paper presented at the 2nd Technical Meeting of Gas Processors Association GCC Chapter, Bahrain (Oct. 27, 1993).

Huntington, R.L., "Natural Gas and Natural Gasoline." McGraw-Hill, New York (1950).

Katz, D.L., *et al*., "Handbook of Natural Gas Engineering." McGraw-Hill, New York (1959).

Kean, J.A., Turner, H.M., and Price, B.C., Structured packing proven superior for TEG gas drying. *Oil Gas J*. **89**(38), 41–46 (1991).

Kirchgessner, D.A., Richards, R.G., Heath, F., and Smith, R.D., Advanced dehydrator design recovers gas, reduces emissions. *Oil Gas J*. **102**(28), 52–60 (2004).

Kohl, A.L., and Riesenfeld, F.C., "Gas Purification," 4th Ed. Gulf Professional Publishing, Houston, TX (1997).

Le Bec, R. *et al*., “New Developments in the Molecular Sieves Field and Their Impact on the Natural Gas Industry.” LNG 12, Perth, Australia (1998).

Ledoux, E., *Chem. Eng*. (March 1948).

Manning, F.S., and Thompson, R.E., “Oil Field Processing of Petroleum,” Vol. 1. Pennwell Publishing Company, Tulsa, OK (1991).

McCabe, W.L., Smith, C.J., and Harriott, P., “Unit Operations of Chemical Engineering,” 4th Ed. McGraw-Hill, New York (1985).

Mcketta, J.J., and Wehe, A.H., “Use This Chart for Water Content of Natural Gases.” Petroleum Refiner (Hydrocarbon Processing), 37, 8, 153 (Aug. 1958).

Meyer, P.B. Chr., “Hydrothermal Damaging of Molecular Sieve and How to Prevent It.” Paper presented at the GPA Europe Conference, Paris, France (Feb. 2003).

Meyer, P.B. Chr., “Overview of Molecular Sieves Technology for Gas Treatment.” Presented at the GPA Europe Gas Processing Overview Session, London, UK (May 19, 2005).

NATCO, “Adsorption Systems-Gas Processing.” Technical Development Program, Tulsa, OK (1987).

NATCO, “Glycol Dehydration Systems.” TSL Catalog, 511-A2, Tulsa, OK (1984).

Oi, L.E., “Calculation of Dehydration Absorbers Based on Improved Phase Equilibrium Data.” Proceeding of the 78th GPA Annual Convention, Nashville, Tennessee (March 2–3, 1999).

Oi, L.E., Estimation of tray efficiency in dehydration absorbers.” *Chem. Eng. Proc*. **42**, 867–878 (2003).

Rojey, A., Jaffret, C., Cornot-Gandolphe, S., Durand, B., Jullian, S., and Valais, M., “Natural Gas Production, Processing, Transport.” Editions Technip, IFP Publications, Paris, France (1997).

Simpson, E.A., and Cummings, W.P., A practical way to predict silica gel performance.” *Chem. Eng. Prog*. **60**(4), 57–60 (1964).

Sivalls, C.R., “Glycol Dehydration Design Manual.” Sivalls, Inc., Odessa, TX (June 1976).

Smith, R.S., and Tsai, M.J., “Application of Drizo Dehydration in North Sea Gas Processing.” Unit Operations in Offshore and Onshore Gas Processing, Int. Chem. Eng. Symposium, Aberdeen (June 1987).

Souders, M., and Brown, G.G., Fundamental design of absorbing and stripping columns for complex vapours. *Ind. Eng. Chem*. **24**, 519 (1932).

Suckow, M., Lutz, W., Kornatowski, J., Rozwadowksi, M., and Wark, M., Calculation of the hydrothermal long term stability of zeolites in gas-desulphurization and gas-drying processes. *Gas Separ. Purif. J.* **6**(2), 101–108 (1992).

Union Carbide, Technical Data, Danbury, CT (1988).

Wichert, G.C., and Wichert, E., New charts provide accurate estimations for water content of sour natural gas. *Oil Gas J.* **101**(41), 64–66 (2003).

Zeochem, "Molecular Sieve Type 4A." Technical Data Z12-1M-82, Zeochem, Louisville (1989).

CHAPTER 10

NATURAL GAS LIQUIDS RECOVERY

10.1 INTRODUCTION

Most natural gas is processed to remove the heavier hydrocarbon liquids from the natural gas stream. These heavier hydrocarbon liquids, commonly referred to as natural gas liquids (NGLs), include ethane, propane, butanes, and natural gasoline (condensate). Recovery of NGL components in gas not only may be required for hydrocarbon dew point control in a natural gas stream (to avoid the unsafe formation of a liquid phase during transport), but also yields a source of revenue, as NGLs normally have significantly greater value as separate marketable products than as part of the natural gas stream. Lighter NGL fractions, such as ethane, propane, and butanes, can be sold as fuel or feedstock to refineries and petrochemical plants, while the heavier portion can be used as gasoline-blending stock. The price difference between selling NGL as a liquid and as fuel, commonly referred to as the "shrinkage value," often dictates the recovery level desired by the gas processors. Regardless of the economic incentive, however, gas usually must be processed to meet the specification for safe delivery and combustion. Hence, NGL recovery profitability is not the only factor in determining the degree of NGL extraction. The removal of natural gas liquids usually takes place in a relatively centralized processing plant, where the recovered NGLs are then treated to meet commercial specifications before moving into the NGL transportation infrastructure.

This chapter briefly presents the basic processes used to separate natural gas liquids from the gas, fractionating them into their various components, and describes different methods for the natural gasoline and liquefied petroleum gas (LPG) treatment.

10.2 NGL RECOVERY PROCESSES

Figure 10-1 shows the phase behavior of a natural gas as a function of pressure and temperature. Obviously any cooling outside the retrograde condensation zone will induce condensation and yield NGL. Retrograde condensation phenomenon has an important application in NGL production. Some plants operate at inlet pressures above the critical point and thus revaporize NGLs when the temperature drops below the retrograde temperature. It is therefore important to know where we are on the phase envelope.

The basic NGL recovery processes are related to Figure 10-1, where possible. Process description and additional details for several NGLs processes are discussed later.

10.2.1 Refrigeration Processes

Refrigeration processes are used at many different temperature levels to condense or cool gases, vapor, or liquids. This section discusses several refrigeration processes for NGL recovery.

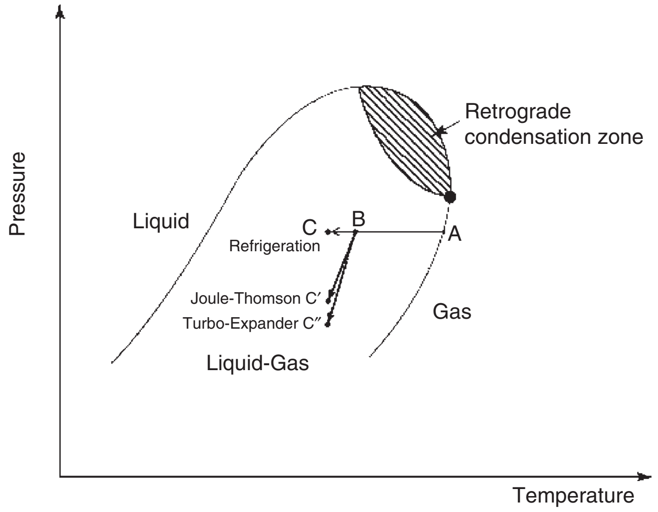


Figure 10-1. Thermodynamic pathways of different NGL recovery technologies.

10.2.1.1 Mechanical Refrigeration

Mechanical refrigeration is the simplest and most direct process for NGL recovery. Mechanical or external refrigeration, as shown in Figure 10-2, is supplied by a vapor-compression refrigeration cycle that usually uses propane as the refrigerant and reciprocating or centrifugal types of compressors to move the refrigerants from low to high pressure operating conditions. As shown in Figure 10-2, the gas-to-gas heat exchanger recovers additional refrigeration by passing the gas, leaving the cold separator countercurrent to the warm inlet gas. The temperature of the cold gas stream leaving this exchanger "approaches" that of the warm inlet gas. Economically, this approach can be as close as 5°F. The chiller in Figure 10-2 is typically a shell and tube, kettle-type unit. The process gas flows inside the tubes and gives up its energy to the liquid refrigerant surrounding the tubes. The refrigerant (often propane) boils off and leaves the chiller vapor space essentially as a saturated vapor.

The thermodynamic path followed by the gas in an external refrigeration process is shown as line ABC in Figure 10-1. From A to B indicates gas-to-gas heat exchange; from B to C, chilling.

When water is present in a natural gas processed by refrigeration, hydrate formation is prevented either by dehydration of the gas or by injection of a hydrate inhibitor. If the processing temperature is relatively low, the gas is usually dehydrated before the refrigeration step.

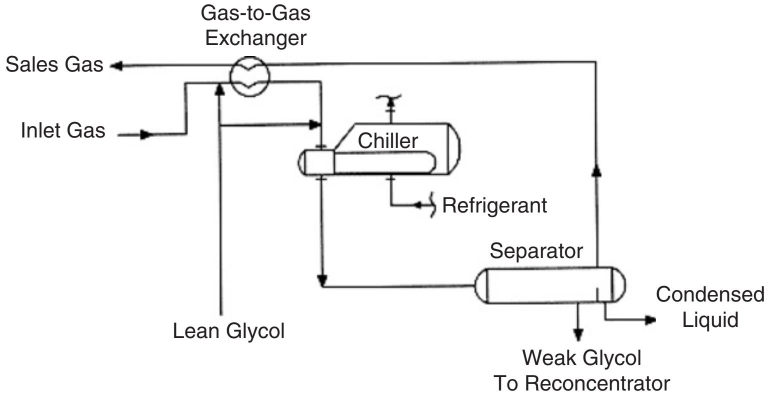


Figure 10-2. Flow sheet of a mechanical refrigeration process (Geist, 1985).

If not, injection of an inhibitor (usually methanol or glycols) upstream of the gas-to-gas heat exchanger is often the simplest and most economical solution (Rojey *et al.*, 1997).

In this case, ethylene glycol is injected at the inlet of the gas-to-gas exchanger and/or chiller to prevent hydrate formation, or freeze up, in these exchangers with the latter being more common. Freeze up will partially block exchanger tubes, thus increasing pressure drop and decreasing heat exchange. The weak glycol solution, containing absorbed water, is separated in the cold separator, reconcentrated, and recycled.

A mechanical refrigeration process is adopted when sizeable amounts of condensate are expected. This process may also lead to the recovery of liquified petroleum gas,[1] where for LPG recovery up to 90%, a simple propane refrigeration system provides refrigeration at temperatures to −40°F (Lee *et al.*, 1999). There are many straight refrigeration process schemes, which vary according to all the design variables, such as gas composition, process pressure, and LPG recovery objectives. A detailed discussion is given by Russell (2001).

10.2.1.1.1 Choice of Refrigerant Any material could be used as a refrigerant. The ideal refrigerant is nontoxic, noncorrosive, has Pressure-Volume-Temperature (PVT) and physical properties compatible with the system needs, and has a high latent heat of vaporization. The practical choice reduces to one, which has desirable physical properties and will vaporize and condense at reasonable pressures at the temperature levels desired. In general, the lower practical limit of any refrigerant is its atmospheric pressure boiling point. It is desirable to carry some positive pressure on the chiller to obtain better efficiency in the compressor, reduce equipment size, and avoid air induction into the system. Propane is by far the most popular refrigerant in gas processing applications. It is readily available (often manufactured on-site), inexpensive, and has a "good" vapor pressure curve. It is flammable but this is not a significant problem if proper consideration is given to the design and operation of the facility (Campbell, 1992).

10.2.1.1.2 Cascade Refrigeration Cascade refrigeration refers to two refrigeration circuits thermally connected by a cascade condenser, which

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[1] Liquified petroleum gas is the general expression for propane, butane and a mixture of the two, which are produced from two distinct sources.

is the condenser of the low-temperature circuit and the evaporator of the high-temperature circuit. A cascade system utilizes one refrigerant to condense the other primary refrigerant, which is operating at the desired evaporator temperature. This approach is usually used for temperature levels below −90°F, when light hydrocarbon gases or other low boiling gases and vapors are being cooled (Lee *et al.*, 1999). To obtain the highest overall efficiency for the system, the refrigerants for the two superimposed systems are different. Cascade refrigeration systems are not common in gas processing. Low-level refrigeration is typically provided using mixed refrigerants or expansion process (GPSA, 1998).

10.2.1.1.3 Mixed Refrigerants An alternative to cascade refrigeration is to use a mixed refrigerant. Mixed refrigerants are a mixture of two or more components. The light components lower the evaporation temperature, and the heavier components allow condensation at ambient temperature. The evaporation process takes place over a temperature range rather than at a constant temperature as with pure component refrigerants. The mixed refrigerant is blended so that its evaporation curve matches the cooling curve for the process fluid. Heat transfer occurs in a countercurrent exchanger, probably an aluminum plate fin, rather than a kettle-type chiller. Mixed refrigerants have the advantage of better thermal efficiency because refrigeration is always being provided at the warmest possible temperature (Mackenzie and Donnelly, 1985). The amount of equipment is also reduced to a cascade system. Disadvantages include a more complex design and a tendency for the heavier components to concentrate in the chiller unless the refrigerant is totally vaporized (Campbell, 1992).

10.2.1.2 Self-Refrigeration

As opposed to external refrigeration, in the self-refrigeration process (Figure 10-3), the inlet gas is precooled against the treated gas (through the gas to gas exchanger) and subsequently cooled further by isoenthalpic expansion (i.e., Joule–Thomson expansion) through a valve, resulting in heavy hydrocarbons and water to condense. In this process, the nonideal behavior of the inlet gas causes the gas temperature to fall with the pressure reduction, as shown by line ABC′ in Figure 10-1. The temperature change depends primarily on the pressure drop. The condensed liquids

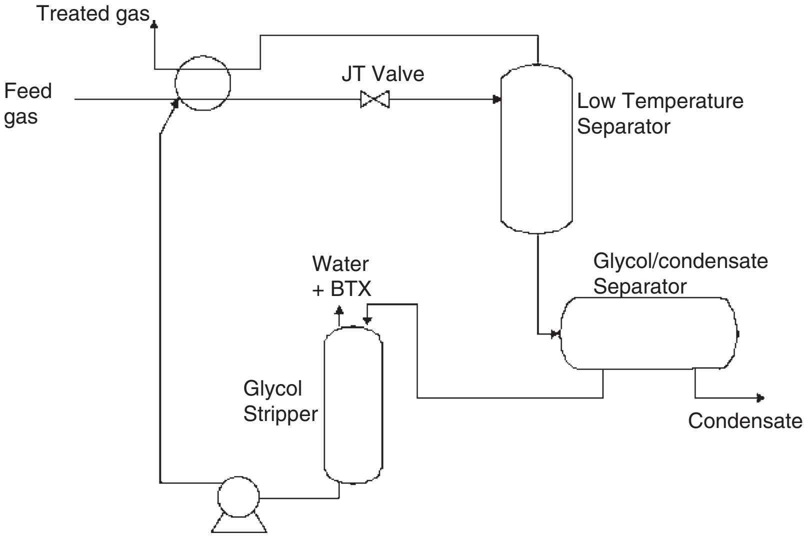

Figure 10-3. Flow sheet of a self-refrigeration system (Brands and Rajani, 2001).

are then removed in one or more separators to meet the vapor pressure and composition specifications. The gas from the low temperature separator (LTS), now satisfying the sales gas specification, is reheated against the incoming feed. Generally, treated gas must be recompressed to sales pipeline pressure if it has been expanded to a lower pressure. Thus, this process is most favored when the raw gas is produced at a very high pressure and can be expanded to sales-line pressure with no recompression. If the gas must be recompressed, the process is penalized by the recompression horsepower requirement.

Note that the extent of cooling in this process is often limited by the hydrate formation temperature at the prevailing pressure unless a hydrate inhibitor, such as methanol or glycol, is injected upstream of the gas-to-gas heat exchanger. In this case, the wet glycol and hydrocarbon condensate are heated and separated in a three-phase separator. Glycol can be regenerated in a stripper. Because glycol shows some affinity for hydrocarbons, the off-gas from the glycol stripper often contains BTX (Brands and Rajani, 2001).

If the objective is to recover ethane or more propane than obtainable by mechanical refrigeration, a good process can be self-refrigeration, which is particularly applicable for smaller gas volumes of 5 to 10 MMCFD (Russell, 2001). The self-refrigeration process may also is attractive if the inlet gas pressure is very high. It is important that the reservoir pressure remain high for the intended life of the plant. Low-pressure inlet gas favors a cryogenic refigeration plant or straight refrigeration process (if the gas is very rich).

10.2.1.3 Cryogenic Refrigeration

When insufficient pressure is available to attain the required dew point with the self-refigeration process, cryogenic refrigeration can be considered. Cryogenic refrigeration processes traditionally have been used for NGL recovery. These plants have a higher capital cost but a lower operational cost. Moreover, they contain numerous moving parts and are complicated to operate (Ewan *et al.*, 1975). In the cryogenic or turbo-expander plant, the chiller or Joule–Thomson (JT) valve used in two previous processes is replaced by an expansion turbine. As the entering gas expands, it supplies work to the turbine shaft, thus reducing the gas enthalpy. This decrease in enthalpy causes a much larger temperature drop than that found in the simple JT (constant enthalpy) process. The expansion process is indicated as line ABC″ in Figure 10-1. The turbine can be connected to a compressor, which recompresses the gas with only a small loss in overall pressure. This results in a higher treated gas pressure, which can be increased to the pipeline specification by a second compression step.

Although there are variations in the design of expander plants, most expander plants have the same basic process flow as shown in Figure 10-4. The inlet gas is first cooled in the high-temperature, gas-to-gas heat exchanger and then in the propane chiller. The partially condensed feed gas is sent to a separator. The liquid from the separator is fed to the demethanizer, and the gas is cooled further in the low-temperature gas-to-gas exchanger and fed into a second cold separator. Gas from the cold separator expands through the expansion turbine to the demethanizer pressure, which varies from 100 to 450 psia. The turbo expander simultaneously produces cooling/condensing of the gas and useful work, which may be used to recompress the sales gas. Typically 10 to 15% of the feed gas is condensed in the cold separator, which is usually at −30 to −60°F. The expander lowers the pressure from the inlet gas value (600 to 900 psia) to the demethanizer pressure of 100 to 450 psia. Typical inlet

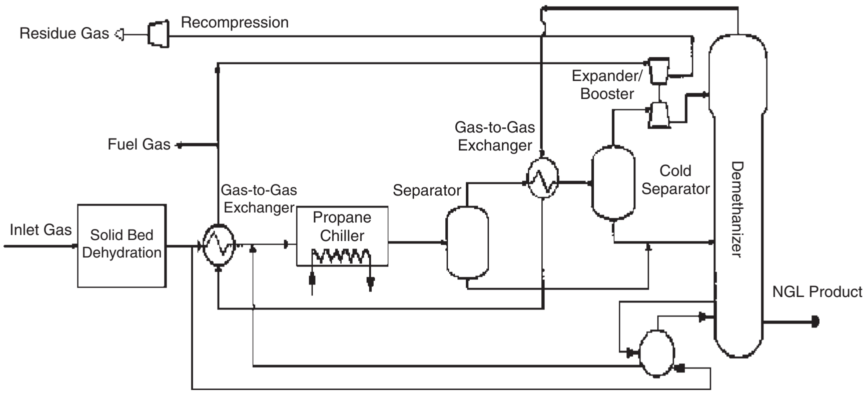

Figure 10-4. Typical flow sheet of a cryogenic refrigeration plant (Ewan *et al.*, 1975).

gas temperatures to the demethanizer are −130 to −150°F, sufficiently low that a great deal of the ethane is liquefied. The demethanizer is a low-temperature distillation column that makes a separation between methane and ethane. Methane and components lighter than methane, such as nitrogen, are the principal products in the vapor near the top of the column, whereas ethane and heavier components, such as propane, butanes, and heavier hydrocarbons, comprise the principal components in the bottom product of the column. The molar ratio of methane to ethane in the bottom product is typically 0.01 to 0.03. Because the outlet of the expander is usually two-phase flow, the liquid produced in the expander serves as reflux for the demethanizer (Elliot *et al.*, 1996). The bottom product from the demethanizer can be fractionated further to produce pure product streams of ethane, propane, butanes, and natural gasolin. The bottom product temperature is often below ambient so that feed gas may be used as the heat transfer medium for the reboiler. This provides additional refrigeration to the feed and yields higher ethane recovery, typically 80% (Holm, 1986). The top product from the demethanizer, after heat exchange with the inlet gas, is recompressed to pipeline pressure and is delivered as the sales gas.

Cryogenic processes can only be applied if the gas pressure after expansion is sufficiently high for condensation of the heavier components to take place. However, if the gas arrives at low pressure (say less than 725 psi), external mechanical refrigeration must be used to cool it to the specified temperature (Cranmore and Stanton, 2000). In addition, if the NGL content of the feed gas is relatively low [less than 2.5 to 3 gallons per Mcf (GPM[2]], self-refrigeration (gas-to-gas heat exchanger) usually suffices. However, for moderately rich feeds (>3 GPM), mechanical refrigeration should be considered to obtain high ethane recovery most economically (Ewan *et al.*, 1975).

Note that in this application, the attainable hydrocarbon dew point specifications are very good, but similar to the self-refrigeration process, upstream dehydration of the gas is required to prevent hydrate formation. Solid bed dehydration upstream of the plant is a proven technique for preventing hydrate formation. Sometimes, small quantities of methanol or glycol can also be injected upstream of the expander.

Cryogenic refrigeration process is generally the most technically advanced type of NGL recovery used today. This combines high recovery levels (typically allowing full recovery of all of the propane and heavier NGLs and recovery of 50% to more than 90% of the ethane) with low capital cost and easy operation (Lee *et al.*, 1999). This is less attractive on very rich gas streams or where the light NGL product (C_2 and C_3) is not marketable, whereas for gases very rich in NGL, simple refrigeration is probably the best choice.

Within the liquids recovery section of the gas processing plant, there are both operating cost and operating flexibility issues that directly impact the processing cost. While it is easily recognized that the efficiency of the selected liquids recovery process is an important factor in the processing cost, the flexibility of operating the process to either recover or reject ethane without sacrificing efficiency or propane recovery is often the critical factor (Pitman *et al.*, 1998). As the industry matured and the demand for more efficient ethane recovery increased, several new and very clever designs have been developed. Up until now, Ortloff's gas subcooled process (GSP) and residue split-vapor (RSV) process have been state of the-art for efficient NGL/LPG recovery from natural gas, particularly for those gases containing significant concentrations of carbon dioxide,[3] which are discussed next.

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[2] GPM is an indication of the potential recoverable NGLs contained in the gas stream. The NGL content of produced gas can vary from 0.2 GPM for very dry gas to 20 GPM for very rich gas.

[3] Many NGL recovery process require removal of the CO_2 to avoid solids formation (freezing) in the cold sections of the processing plant.

10.2.1.3.1 Ortloff Gas Subcooled Process To increase the ethane recovery beyond the 80% achievable with the conventional design, a source of reflux must be developed for the demethanizer. The GSP was developed to overcome this problem and others encountered with the conventional expander scheme. In this process, shown in Figure 10-5, a portion of the gas from the cold separator is sent to a heat exchanger where it is totally condensed and subcooled with the overhead stream. This stream is then flashed to the top of demethanizer, providing reflux to the demethanizer. The expander feed is sent to the tower several stages below the top of the column. Because of this modification, the cold separator operates at much warmer conditions well away from the system critical. Additionally, the residue recompression is less than with the conventional expander process.

The GSP design has several modifications. One is to take a portion of the liquid from the cold separator along with the gas to the overhead exchanger. Generally, this can help further reduce the horsepower required for recompression. Also, the process can be designed to just use a portion of the cold separator liquid for reflux. This modification is typically used for gases richer than 3 GPM (C_2^+). The GSP design is very CO_2 tolerant; many designs require no up front CO_2 removal to achieve high recovery. CO_2 levels are very composition and operating pressure dependent, but

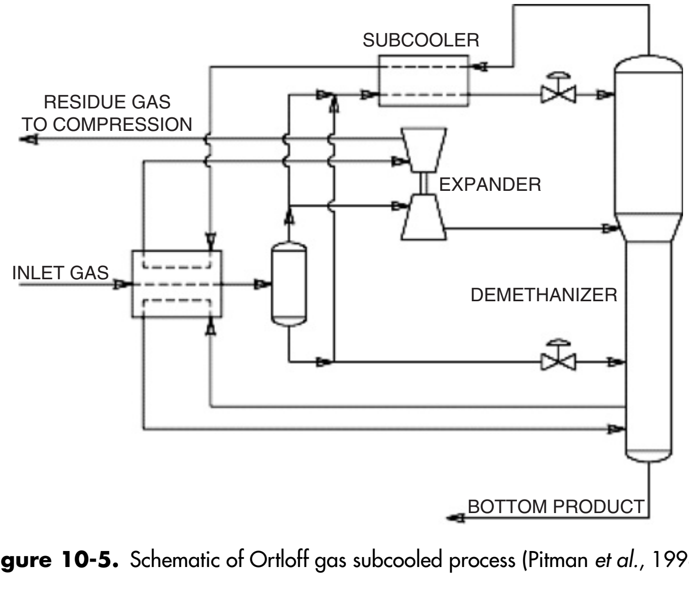

Figure 10-5. Schematic of Ortloff gas subcooled process (Pitman *et al.*, 1998).

levels up to 2% can usually be tolerated with the GSP design. When CO_2 is present in the feed gas, the higher concentrations of C_2^+ components in the cold liquids help reduce the amount of CO_2 concentrating in the upper, colder sections of the tower, allowing higher ethane recovery levels without CO_2 freezing. This same process can be operated to reject ethane, but propane recovery efficiency suffers significantly when operated in this mode due mainly to the higher concentration of propane present in the top feed (Pitman *et al.*, 1998).

Since the early 1980s, the Ortloff's GSP has been the pioneer process for high ethane recovery NGL plants, where all of the current "state-of-the-art" NGL recovery technologies are basically enhancements of the GSP concept (Nasir *et al.*, 2003).

10.2.1.3.2 Ortloff Residue Split-Vapor Process Another method of producing reflux is to recycle a portion of the residue gas, after recompression, back to the top of the column, the RSV process. As shown in Figure 10-6, the process flow is similar to the GSP design except that a portion of the residue gas is brought back through the inlet heat exchanger. At this point, the stream is totally condensed and is at the residue gas pipeline pressure. The stream is then flashed to the top of the demethanizer to provide reflux. The subcooled inlet gas split and the

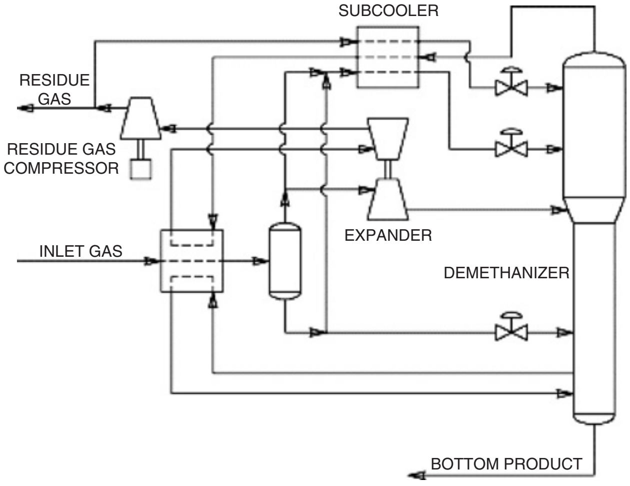

Figure 10-6. Schematic of Ortloff residue split-vapor process (Pitman *et al.*, 1998).

expander outlet stream are sent lower down in the tower rather than to the top of the column. The reflux provides more refrigeration to the system and allows very high ethane recovery to be realized. The recovery level is a function of the quantity of recycle in the design.

The RSV process has been used successfully in numerous facilities. It is CO_2 tolerant and the recovery can be adjusted by the quantity of recycle used. The RSV process can be used for very high ethane recoveries limited only by the quantity of horsepower provided.

10.2.2 Lean Oil Absorption

The absorption method of NGL recovery is very similar to using absorption for gas dehydration. The main difference is that, in NGL absorption, an absorbing oil is used as opposed to glycol. This absorbing oil has an "affinity" for NGLs in much the same manner as glycol has an affinity for water.

Lean oil absorption is the oldest and least efficient process to recover NGLs. In this process, as shown in Figure 10-7, the gas to be processed is contacted in a packed or tray absorption column (typically operated at the ambient temperature and a pressure close to the sales gas pressure) with an absorption oil (lean oil), which absorbs preferentially the most

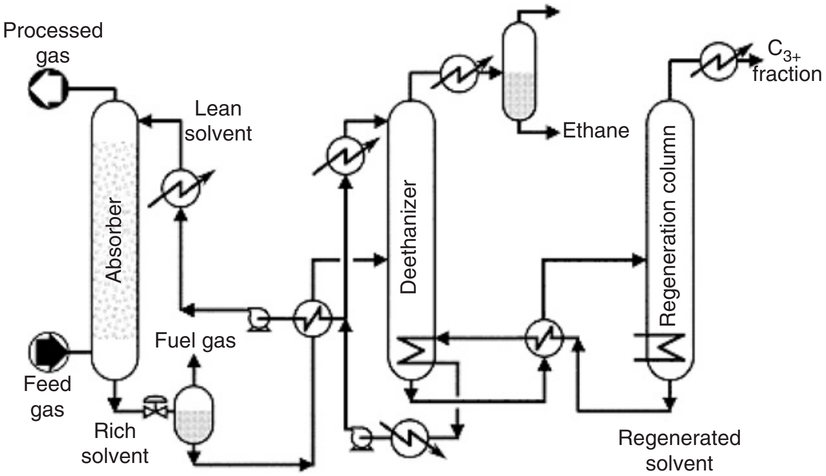

Figure 10-7. Simplified flow diagram of an oil absorption plant (Rojey *et al.*, 1997).

heavy hydrocarbons (C_3-C_7^+) from natural gas. The gas leaves the top of the absorber while the absorber oil, now rich in heavy hydrocarbons from the gas, leaves the bottom of the absorber and is expanded to liberate most of the absorbed methane. Rich oil is then sent to a deethanizer to reject all the methane and part of the ethane absorbed. At the top of this column, additional cold oil is injected to limit the desorption of ethane and to prevent the desoption of C_3^+. Rich oil then flows to a regeneration column where it is heated to a high enough temperature to drive the propanes, butanes, pentanes, and other natural gas liquid components to the overhead, and the regenerated solvent (lean oil) is recycled.

Note that an oil absorption plant cannot recover ethane and propane effectively, where it requires circulating large amounts of absorption oil, demands attendant maintenance, and consumes too much fuel. However, an oil absorption plant can be modified to improve its propane recovery by adding a propane refrigeration cycle for cooling. The liquid recovery possible in a lean oil plant is typically 99% of the butane and natural gasoline, 65–75% of the propane, and 15–25% of the ethane contained in the gas (Spletter and Adair, 2001).

Lean oil absorption plants are not as popular as they once were and are rarely, if ever, constructed anymore. They are expensive and more complex to operate, and it is difficult to predict their efficiency at removing liquids from the gas as the lean oil deteriorates with time (Arnold and Stewart, 1999). They are rarely designed as new plants. Existing lean oil plants are sometimes salvaged, refurbished, and moved to new locations.

10.2.3 Solid Bed Adsorption

This method uses adsorbents that have the capability to adsorb heavy hydrocarbons from natural gas. The adsorbent may be silica gel or activated charcoal. Activated alumina cannot be used in the presence of heavy hydrocarbons, which foul the adsorbent (Rojey *et al.*, 1997). Note that the design of an adsorbent-based system for heavy hydrocarbon removal is more complicated than that of a system for the removal of water only. For instance, different grades of adsorbent may be required, and the system must also be designed to accommodate the adsorption of more than one component (Daiminger and Lind, 2004).

The adsorption process required for removing heavy hydrocarbons is shown schematically in Figure 10-8. The process is continuous with

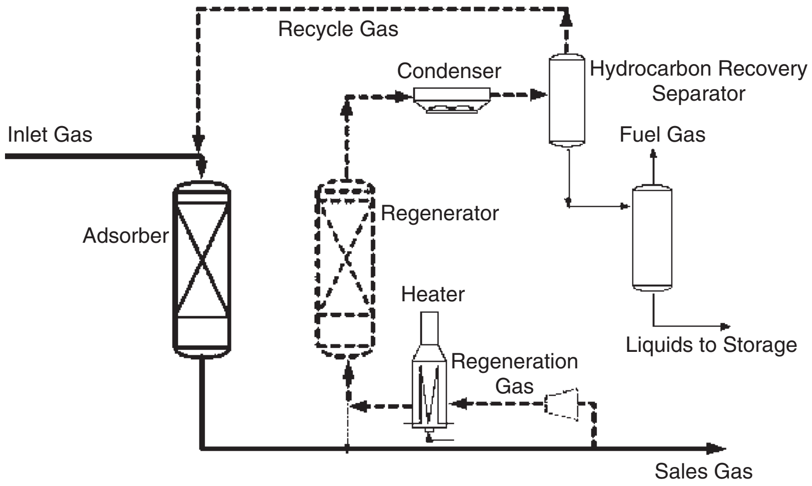

Figure 10-8. Schematic of a solid bed adsorption plant (Foglietta, 2004).

respect to the gas but cyclical with respect to the adsorbent bed because the latter must be regenerated when it becomes saturated with condensate. Regeneration is accomplished by passing heated recycle gas through the bed. The condensate is recovered from the regeneration gas by cooling, condensation, and phase separation. To recover a large fraction of the hydrocarbons, while limiting the volume of adsorbent, it is preferable to use a relatively short cycle time, about 1 hour. In practice, this cycle time may vary within a fairly wide interval, between 20 minutes and several hours, depending on how hydrocarbon rich the gas is (Ballard, 1965).

This process is appropriate for relatively low concentrations of heavy hydrocarbons. It can also be appropriate if the gas is at a high pressure, close to the cricondenbar. In this case, refrigeration processes become ineffective and separation by adsorption may offer the only way to obtain the required specifications (Parsons and Templeman, 1990).

Adsorption processes are easy to start up and to operate at high turn down (changes in throughput) and so are useful for variable and "on–off" operation. However, the adsorption beds are heavy in weight and expensive. This process option is not often used but may be considered in special applications, such as hydrocarbon dew point control in remote locations.

10.2.4 Membrane Separation Process

Refrigeration and cryogenic plants traditionally have been used for NGL recovery. These plants have high capital and operating costs. Moreover, they contain numerous rotating parts and are complicated to operate. The membrane separation process offers a simple and low-cost solution for removal and recovery of heavy hydrocarbons from natural gas. Figure 10-9 shows a schematic of a membrane separation process for NGL recovery (Foglietta, 2004). As shown, the separation process is based on a high-flux membrane that selectively permeates heavy hydrocarbons compared to methane. These hydrocarbons permeate the membrane and are recovered as a liquid after recompression and condensation. The residue stream from the membrane is partially depleted of heavy hydrocarbons and is then sent to a sales gas stream.

Gas permeation membranes are usually made with vitreous polymers that exhibit good diffusional selectivity. However, for separation to be effective, the membrane must be very permeable with respect to the contamination to be separated, which passes through the membrane driven by pressure difference, and it must be relatively impermeable to methane

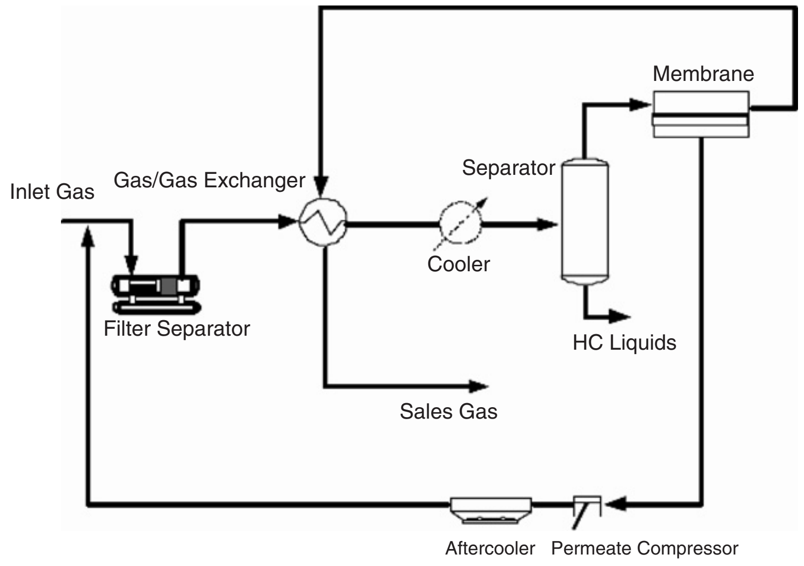

Figure 10-9. Schematic flowchart of the membrane separation process (Foglietta, 2004).

(Rojey *et al.*, 1997). Membrane Technology & Research Inc. has developed and comercialized a new membrane-based process for the gas processing industry. The enabling technology of this process is a unique type of rubbery membrane, which is now being applied to the separation of C_3^+ hydrocarbons from methane in gas processing. The new membrane-based process is well suited for NGL recovery and dew point control for associated gas and may also be used to debottleneck existing gas processing plants. A detailed discussion on this novel, membrane-based process and its applications in the gas processing industry is given by Lokhandwala and Jacobs (2000).

Membrane systems are very versatile and are designed to process a wide range of feed conditions. With very compact footprint and low weight, these systems are well suited for offshore applications.

10.2.5 Selection of NGL Recovery Processes

Choosing a cost-effective NGL recovery technology requires consideration of a broad range of factors (Mehra and Gaskin, 1999). The main variables that affect the choice of the most cost-effective process for a given application include inlet conditions (gas pressure, richness, and contaminants), downstream conditions (residue gas pressure, liquid products desired, and liquid fractionation infrastructure), and overall conditions (utility costs and fuel value, plant location, existing location infrastructure, and market stability). In addition to the feed gas composition and operation mode, the most decisive technical characteristics of any process are the feed gas pressure and permissible unit pressure drop. The following guidelines have been suggested for the selection of a NGL recovery process (Brands and Rajani, 2001).

1. In case of sufficiently high pressure, the self-refrigeration process requires the lowest capital investment. However, if the pressure differential between feed gas and treated gas is insufficient, additional compression is required.
2. When the feed gas pressure is close to the treated gas pressure, over a large pressure drop range, it may be more economical to employ a cryogenic refrigeration process.
3. When the feed gas pressure is clearly below the required pipeline pressure, it is usually most economical to apply mechanical refrigeration with additional compression to remove heavy hydrocarbons instead of

compression followed by the self-refrigeration process. This is due to the fact that compressors are capital intensive equipments.

4. When the feed gas pressure is equal to or lower than the required pipeline pressure, solid bed adsorption seems a good option, as it is quick to start up and is robust against changes in the feed gas composition and flow rate. Generally the solid bed process is only practical for gas that has small amounts of heavy hydrocarbons. Richer gases require refrigeration.

It is clear that the solid bed adsorption process will usually be competing against the self-refrigeration process. Specially, the solid bed adsorption unit is operated at lower differential pressure compared to self-refrigeration and thus no additional compression is required. In fact, at low feed gas pressure and for strict dew point specifications, economical analysis favors the solid bed adsorption process.

With reference to the membrane application to control the hydrocarbon dew point, there is no clear judgement. Current discussions look at this on a point-by-point base and compare the economics with other processes. The window of opportunity is still to be seen, although its use in lean fuel gas is more common.

10.3 NGL FRACTIONATION

The bottom liquid from the NGL recovery plant may be sold as a mixed product. This is common for small, isolated plants where there is insufficient local demand. The mixed product is transported by truck, rail, barge, or pipeline to a central location for further processing. Often it is more economical to fractionate the liquid into its various components, which have a market value as purity products. However, as the relative prices of natural gas and NGLs fluctuate, the relative incentive to extract the NGLs from the gas changes. The level of NGL extraction from natural gas is somewhat discretionary. Safety issues dictate the minimum extraction level, whereas a balance between the technology and the relative market value of the NGLs determines the maximum extraction level.

The process of separating a stream of NGLs into its components is called fractionation. At the fractionation plant, liquids will be separated into commercial quality products and then delivered to the market by tankers (exports) and tank trucks (domestic consumption). NGLs are fractionated by heating mixed NGL streams and passing them through a series of distillation towers. Fractionation takes advantage of the differing boiling points of the various NGL products. As the temperature of the NGL stream is increased, the lightest (lowest boiling point) NGL product boils off the top of the tower as a gas where it is then condensed into a purity liquid that is routed to storage. The heavier liquid mixture at the bottom of the first tower is routed to the second tower where the process is repeated and a different NGL product is separated and stored. This process is repeated until the NGLs have been separated into their components.

Fractionators are usually named for the overhead or top product, as depicted in the fractionation flow schematic of Figure 10-10. Therefore, a deethanizer implies that the top products is ethane; a depropanizer indicates that the top product is propane, etc. Natural gas liquids are normally fractionated by boiling the lighter products from the heavier products in the following order (Tuttle and Allen, 1976).

- Deethanizer: The first step in the fractionating sequence is to separate the ethane and propane, with the ethane going overhead and the propane and heavier components passing from the bottom of the fractionator.
- Depropanizer: The next step in the processing sequence is to separate the propane and the isobutane, with the propane going overhead and the isobutane and heavier components passing from the bottom of the depropanizer.
- Debutanizer**:** The next fractionating step is separation of butanes from pentanes plus (C_5^+) stream. The butanes (both iso and normal) pass overhead and the pentanes plus from the bottom of the fractionator.

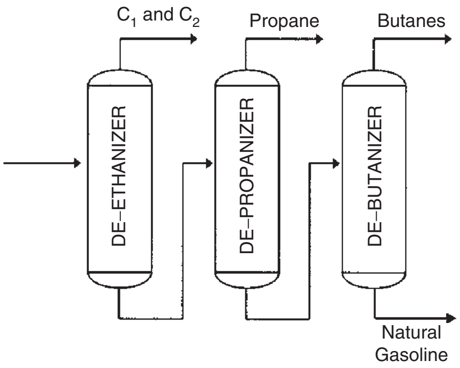

Figure 10-10. Simplified flow diagram of a fractionation plant.

- Butane Splitter or Deisobutanizer: When it is desirable to do so, the butanes that pass overhead from the debutanizer may be separated into iso- and normal butanes. The isobutane goes overhead and the normal butane is drawn from the bottom of the tower.

The system shown in Figure 10-10 most commonly produces commercial propane, comercial butane, and natural gasoline as products. In this system, the deethanizer must remove all methane, ethane, and other constituents not salable in one of the three products. Sometimes, depending on the market situation of each product, the number of fractionating columns may be varied. For example, when a market for liquified petroleum gas (LPG) only exists during a portion of the year, an LPG mixture could be produced overhead with natural gasoline produced as bottoms in the second tower, and the third tower would be shut down and not operated.

Note that fractionation of the NGLs to produce purity products such as petrochemical grade ethane feedstock or fuel grade propane may take place in a gas processing facility but occurs more commonly at another location, usually a regional market center (Spletter and Adair, 2001).

10.3.1 Fractionator Operation

The operation takes place in a vertical column where vapor and liquid mixtures flow countercurrent and are brought into repeated contact. During each contact, part of the liquid vaporizes and part of the vapor condenses. As the vapor rises through the column, it becomes enriched in the lighter or lower boiling components. Conversely, the downward flowing liquid becomes richer in heavier, higher boiling components. Figure 10-11 is a schematic view of a typical fractionating column. The liquid mixture that is to be processed is known as the feed, which is introduced usually somewhere near the middle of the column to a tray known as the feed tray. The feed tray divides the column into a top (enriching or rectification) section and a bottom (stripping) section. The feed flows down the column where it is collected at the bottom in the reboiler. Heat is supplied to the reboiler to generate vapor. The vapor raised in the reboiler is reintroduced into the unit at the bottom of the column. The liquid removed from the reboiler is known as the bottoms product or, simply, bottoms. The vapor moves up the column, and as it exits the top of the unit, it is cooled by a condenser. The condensed liquid is stored in a holding vessel known as the reflux drum (accumulator). Some of this liquid is recycled back to the top of the

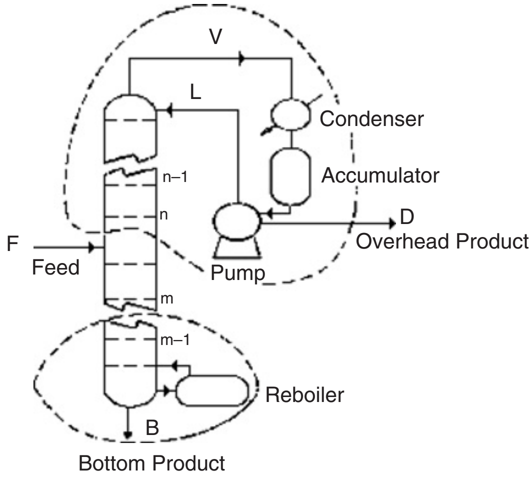

Figure 10-11. Schematic view of a fractionating column (Campbell, 1992).

column and this is called the reflux. The condensed liquid that is removed from the system is known as the distillate or overhead product.

To avoid reducing the operating pressure of the fractionator, which necessitates recompression, the fractionation process should be carried out with minimal loss of pressure and ideally at pipeline pressure if possible. The only limitation is that high-pressure fractionation must occur at a pressure safely below the critical pressure at every stage of the column to assure that both liquid and vapor phases will be present. The designer must also be concerned about the effect of pressure on the relative volatility of the critical components[4] in the fractionating column. As the operating pressure of the fractionating column increases, the relative volatility decreases, making it more difficult to get a clean separation between the two critical components.

The operating pressure of a fractionating column is normally set by a desired component separation and the temperature of the reflux condenser cooling medium (i.e. air, water, and refrigerant). This pressure is

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[4] The two components in a multicomponent mixture, between which a separation is desired, are spoken of as the "key" or "critical" components. The light-boiling key component will appear concentrated in the overhead product, and the heavy key component will appear concentrated in the bottom product. The choice of these components is a decision made by the designer.

the minimum pressure at which the column can operate at the chosen condenser temperature. The pressure will correspond to either the bubble point or the dew point of the column overhead product. The overhead product will be at bubble point conditions for a liquid product or at dew point conditions for a vapor product.

10.3.2 Types of Fractionators

There are two types of fractionating columns: (1) tray column, where trays of various designs are used to hold up the liquid to provide better contact between vapor and liquid, hence better separation, and (2) packed column, where instead of trays, "packings" are used to enhance contact between vapor and liquid phases.

There are many types of tray designs, but the most common ones are bubble cap trays, valve trays, and sieve trays. However, because of their efficiency, wide operating range, ease of maintenance, lower cost, and increased capacity (for a given diameter), valve and sieve trays have replaced the once highly thought of bubble cap trays in many applications. Many kinds of packing have also been invented, and several types are in common use. Packings are divided into those that are dumped at random into the column and those that must be stacked by hand. Dumped packings are used in the smaller column, whereas stacked packings are used only in the larger columns.

Traditionally, the majority of fractionating columns in gas processing plants were eqipped with trays; however, packed columns have become more common in recent years. Advantages of packed columns as compared to trayed columns include lower pressure drop (0.2 to 0.6 inches of water per foot of packed depth) and more capacity (at a given diameter) for high liquid-to-vapor ratio systems. Packed columns also allow the use of smaller equipment and thus lower capital cost. Offsetting these advantages are (1) column plugging, where packed towers will be more susceptible to plugging from dirt and other foreign materials, and (2) packing's more limited turndown (about 50%) and possible problems with liquid maldistribution (which has a serious effect on packed columns having a large number of theoretical stages) and channeling (which is the chief reason for the poor performance of large packed columns). A detailed discussion on design of packed columns, which is beyond the scope of this book, is given in GPSA (1998) and McCabe *et al.* (2001).

10.3.3 Fractionator Design

Important factors in the design and operation of fractionating columns are the number of trays required to obtain the desired separation, the diameter of the column, the heat input to the reboiler, and the heat output from the condenser. In accordance with general principles, analysis of the performance of the fractionating column is based on material and energy balances.

10.3.3.1 Overall Material Balance

In fractionator design, material balances around the column (as shown in Figure 10-11) should be made first. In steady state, two independent overall material balances can be written as follows.

$$F = D + B \tag{10-1}$$

$$F.X_F = D.X_D + B.X_B \tag{10-2}$$

where F is feed molar flow rate, D is distillate molar flow rate, B is bottoms molar flow rate, X_F is mole fraction of a component in the feed, X_D is mole fraction of a component in the distillate, and X_B is mole fraction of a component in the bottoms.

It is obvious from Equations (10-1) and (10-2) that at a given feed composition, the distillate (overhead) and bottoms product specifications set the overall material balance for column. In other words, in order to meet the distillate and bottoms specification exactly there is one (and only one) distillate or bottoms flow rate, which can be drawn from the column.

10.3.3.2 Minimum Theoretical Trays

The number of trays required to meet a given specification at the outlet is generally determined by first considering theoretical stages. A theoretical stage (tray) is an ideal stage from which the phases exist in equlibrium. However, an actual tray will not achieve equilibrium due to limitations of vapor–liquid contact time. Once the number of theoretical trays required is known, the number of actual trays is obtained by taking into account the overall efficiency of each of the trays.[5] Typical values for tray efficiency range from 0.5 to 0.7 and depend on a number of factors, such as the type of trays being used and internal liquid and vapor flow conditions. Detailed information on this subject is given in Perry (1997) and McCabe *et al.* (2001).

The minimum theoretical trays for a given separation will occur when the column is operating at a total reflux. The most rigorous method for determining minimum trays is a tray-to-tray calculation. However, the Fenske (1932) equation for minimum trays offers a rapid and fairly accurate method of computing minimum trays. It can be applied to any two components, i and j, in a column at infinite reflux ratio. When the keys have been chosen, components i and j are the light and heavy keys, respectively. The Fenske (1932) equation can be written in several forms. The most convenient form is as follows.

$$S_{\mathrm{m}} = \frac{\log\left[\left(\dfrac{X_{\mathrm{LK}}}{X_{\mathrm{HK}}}\right)_{\mathrm{D}}\left(\dfrac{X_{\mathrm{HK}}}{X_{\mathrm{LK}}}\right)_{\mathrm{B}}\right]}{\log(\bar{\alpha})} - 1 \qquad (10\text{-}3)$$

where S_{m} is minimum number of theoretical trays, X_{LK} is mole fraction of the light key component, X_{HK} is mole fraction of the heavy key component, and $\bar{\alpha}$ is relative volatility at average column temperature.

Subscripts D and B refer to the distillate and bottoms, respectively. Relative volatility is also defined as the K value[6] of the light key divided by the K value of the heavy key at a given condition. If the change in the value of relative volatility from the bottom of the column to the top is moderate, a geometric mean of the extreme values is recommended for relative volatility (McCabe *et al.*, 2001).

10.3.3.3 Feed Tray Location

The feed tray location should be determined so that feed enters at a temperature so that no sharp change occurs in the column temperature gradient at the feed tray. There is no exact way to locate a feed tray. There are several

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[5] Sometimes, additional trays are added (up to 10%) to accommodate the possibility that the column may be underdesigned.

[6] The distribution coefficient or the K value is defined as the ratio of mole fraction of a components in the gas phase to the mole fraction of that component in the liquid phase.

calculation techniques, which estimate feed tray location. A convenient empirical correlation is as follows (Kirkbride, 1944).

$$\log\left(\frac{N}{M}\right) = 0.206\log\left[\left(\frac{B.X_{\mathrm{HK_F}}}{D.X_{\mathrm{LK_F}}}\right)\left(\frac{X_{\mathrm{LK_B}}}{X_{\mathrm{HK_D}}}\right)^{2}\right] \qquad (10\text{-}4)$$

where N is number of theoretical stages in the rectifying section, M is number of theoretical stages in the stripping section, B is bottoms molar flow rate, D is distillate molar flow rate, $X_{\mathrm{HK_F}}$ is composition of heavy key in the feed, $X_{\mathrm{LK_F}}$ is composition of light key in the feed, $X_{\mathrm{LK_B}}$ is composition of light key in the bottoms, and $X_{\mathrm{HK_D}}$ is composition of heavy key in the distillate.

In Table 10-1 the actual number of trays are included. This is because complete equilibrium between vapor and liquid is normally not reached on each tray. For calculation purposes, the number of theoretical trays may be quite a bit less than the number of actual trays.

A McCabe-Thiele (1925) diagram may be used to estimate the feed tray location and is sometimes used because it is common practice to manifold

Table 10-1
Typical Fractionator Parameters/Number of Trays
(Arnold and Stewart, 1999)

Column	Approximate Ranges		
	Pressure range (psig)	Number of actual trays above main feed	Number of actual trays below main feed
Lean oil plant			
Absorber	200–1100	24–30	20–50
Rich oil Demethanizer	450–600	20–30	20–50
Rich oil Deethanizer	175–300	24–30	20–50
Rich oil still	85–160	12–60	16–60
Refrigeration plant			
Demethanizer	550–650	14–30	26–30
Deethanizer	350–500	10–70	20–70
Depropanizer	200–300	17–70	18–70
Debutanizer	70–100	18–70	15–70

a few trays on each side of the estimated location. Usually, as the feed stage is moved lower down the column, the top composition becomes less rich in the more volatile component while the bottoms contains more of the more volatile component. However, the changes in top composition are not as marked as the bottoms composition.

10.3.3.4 Minimum Reflux Ratio

The minimum reflux ratio[7] for a given separation will occur when the column is operating with an infinite number of trays, and it corresponds to the minimum reboiler heat load and condenser cooling load for the separation. The most rigorous method for determining minimum trays is a tray-to-tray calculation. However, such a procedure is a laborious trial-and-error solution as exemplified by Holcomb et al. (1942). Of the several methods of calculating the minimum reflux ratio, the most used is that of Underwood (1948), which involves use of two following equations:

$$\sum_{i=1}^{i=n} \frac{\alpha_i X_{Fi}}{\alpha_i - \varphi} = 1 - q \tag{10-5}$$

$$\sum_{i=1}^{i=n} \frac{\alpha_i X_{Di}}{\alpha_i - \varphi} = R_m + 1 \tag{10-6}$$

where α_i is relative volatility of component i at average column temperature, X_{Fi} is mole fraction of component i in feed, X_{Di} is mole fraction of component i in distillate, q is thermal condition of the feed, which is the total heat needed to convert 1 mole of feed into a saturated vapor divided by the molal latent heat of the feed (for boiling point feed, $q = 1.0$; for dew point feed, $q = 0$; for two phase feed, $0 < q < 1.0$), R_m is minimum reflux ratio, n is number of components, and φ is constant.

Use of Equations (10-5) and (10-6) to calculate the minimum reflux ratio involves determination of the value of φ that will satisfy Equation (10-5). This is a trial-and-error solution by the limitation that the only roots required are those in numerical value between the relative volatilities of

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[7] The analysis of fractionating column is facilitated by the use of a quantity called the reflux ratio, which is the ratio of the reflux to the overhead product.

the light and heavy keys. Because the relative volatilities for all components are based on the heavy key, this means that $1.0 < \varphi < \alpha_{LK}$. Thus, the value of φ that satisfies Equation (10-5) is used in Equation (10-6) to calculate the minimum reflux ratio. In this calculation, the distillate composition is obtained from the splits that were made for the column. Occasionally, the minimum reflux calculated by this method comes out a negative number. That, of course, is a signal that some other methods should be tried or it may mean that the separation between feed and overhead can be accomplished in less than one equilibrium stage (Walas, 1990).

With the simplifying assumptions of constancy of molal overflow and temperature independence of relative volatilities, the Underwood (1948) method is exact. Further corrections can also be made for variable overflow and temperature variations of relative volatilities when the simplifying assumptions are not applicabale.

The Underwood (1948) method gives only an approximate value for the minimum reflux ratio. However, its results are 5 to 20% high, and for this reason it can be used for column design.

10.3.3.5 Actual Reflux Ratio

As the reflux ratio is increased, more and more liquid that is rich in the more volatile components is being recycled back into the column. Separation then becomes better and thus less trays are needed to achieve the same degree of separation. The boundary conditions for a given separation are represented by minimum stages (total or infinite reflux) and minimum reflux (infinite stages). Obviously, neither of these conditions represents realistic operating conditions, but simply shows the limit between which column can be operated.

The Gilliland (1940) correlation is widely used for calculation of the actual reflux ratio. Once the minimum theoretical stages and minimum reflux ratio have been calculated, the Gilliland correlation can be used to establish an actual reflux ratio for a fixed number of stages or, conversely, the number of stages for a fixed reflux ratio. An approximate correlation of Gilliland for tray-type columns may be represented by the following equation (Eduljee, 1976):

$$Y = 0.75(1 - X^{0.5668}) \tag{10-7}$$

where $Y = (S - S_m)/(S + 1)$, $X = (R - R_m)/(R + 1)$, R is actual reflux ratio, R_m is minimum reflux ratio, S is actual number of theoretical trays, and S_m is minimum number of theoretical trays.

The operating reflux is an amount in excess of the minimum that ultimately should be established by an economic balance between operating and capital costs for the operation. Most columns are designed to operate between 1.2 and 1.5 times the minimum reflux ratio because this is approximately the region of minimum operating costs (more reflux means higher reboiler duty). However, the total cost is not very sensitive to reflux ratio in this range, and better operating flexibility is obtained if a reflux greater than the optimum is used (McCabe *et al.*, 2001).

Typical values of reflux ratio and tray efficiencies for various fractionating columns are given in Table 10-2. These are not design values but rather guidelines for typical values in previous applications. The actual selection depends on many factors, such as feed composition, energy cost, and capital cost.

Note that the actual number of trays required for a particular separation duty is determined by the efficiency of the tray and the packings if packings are used. Thus, any factors that cause a decrease in tray efficiency will also change the performance of the column. Tray efficiencies are affected by fouling, wear and tear, and corrosion, and the rates at which these

Table 10-2
Typical Fractionator Conditions (GPSA, 1998)

Tower	Reflux ratio[a]	Reflux ratio[b]	Tray efficiency (%)
Demethanizer	Top feed	Top feed	45–60
Deethanizer	0.9–2.0	0.6–1.0	50–70
Depropanizer	1.8–3.5	0.9–1.1	80–90
Debutanizer	1.2–1.5	0.8–0.9	85–95
Butane splitter	6.0–14.0	3.0–3.5	90–110
Rich oil fractionator (still)	1.75–2.0	0.35–0.40	Top 67 Bottom 50
Rich oil deethanizer	—	—	Top 25–40 Bottom 40–60

[a] Reflux ratio relative to overhead product, lbmole/lbmole.
[b] Reflux ratio relative to feed, ft^3/ft^3.

occur depend on the properties of the liquids being processed. Therefore, appropriate materials should be specified for tray construction.

Determining the number of stages required for the desired degree of separation and the location of the feed tray are merely the first steps in designing a fractionator. Other things that need to be considered are tray spacing, column diameter, internal configurations, and heating and cooling duties. All of these can lead to conflicting design parameters. Thus, fractionating column design is often an iterative procedure. If the conflicts are not resolved at the design stage, then the column will not perform well in practice.

10.3.3.6 Column Capacity

The capacity of a column is expressed in terms of the rate of either vapor or liquid flow per unit plate area. Depending on column diameter and the type of internals used, there are hydraulic limits to vapor and liquid handling capacity, both on high and low sides. At very high vapor or liquid loads, the column can flood due to excessive liquid entrainment or insufficient downcomer capacity. At very low throughputs, problems such as liquid weeping and vapor pulsation can occur. Loss of desired separation efficiency can also occur before column hydraulic limits are reached, resulting in product specifications not being met. Hence, the column needs to operate within the boundaries of the region designated "satisfactory performance."

To evaluate the column hydraulics, the diameter required for vapor flow is calculated for those stages where there is a significant change in the vapor and liquid flows. Such stages include top and bottom trays, pump-around stages, and the feed stage. The diameter profile along the column is then obtained by plotting the diameters for various stages, where this profile allows identifying the column bottlenecks that limit throughput enhanement (Gadalla *et al.*, 2003).

The column diameter required to handle a given vapor load can be found by applying the following equation at the point of greatest vapor load, generally just below the top tray (Souders and Brown, 1932):

$$V_{\max} = C\sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \tag{10-8}$$

Table 10-3
Value of Sizing Constant vs Tray Spacing
(Souders and Brown, 1932)

C, ft/sec	Tray spacing, inch
0.122	18
0.150	21
0.167	24

where V_{max} is maximum vapor velocity, ft/sec; ρ_L is liquid density, lb_m/ft^3; ρ_V is vapor density, lb_m/ft^3; and C is sizing constant, ft/sec.

The sizing constant (C) has been evaluated from plant data, and various correlations of "C" against operating variables have been made (Perry, 1997). Table 10-3 shows values indicative of those used commonly.

The column capacity as determined by the Souders and Brown (1932) equation is from 20 to 40% conservative, as pointed out by Brown *et al.* (1943).

The general method of column design is to choose a plate spacing, estimate the permissible vapor load, and calculate the column diameter as (GPSA, 1998):

$$D = \sqrt{\frac{\dot{Q}_V}{V_{max} \times 0.7854}} \tag{10-9}$$

where D is column diameter, ft; $\dot{Q}_V$ is volumetric vapor flow rate, ft^3/sec; and V_{max} is maximum vapor valocity, ft/sec.

This method was originally developed for bubble cap trays and gives a conservative diameter, especially for other types of trays.

In cases where a large liquid load is handled, the column diameter may be determined by the liquid load rather than the vapor load, and to obtain a safe liquid load the final design may be underloaded with respect to vapor flow (McCabe *et al.*, 2001).

If the diameter of a fractionator is 2 feet or less, it is more economical in many cases to use a packed column in place of a tray column. Manufacturers supply data for their packing material, which indicates the amount of feet of packing required to provide the same mass transfer as a standard bubble cap tray.

Some recent advances in structured packing are being used by some operators in larger diameter towers where they would have normally used trays. The structured packing is said to allow both smaller diameter and less height of tower. In addition, packings provide extra interfacial area for liquid–vapor contact, and the efficiency of separation is therefore increased for the same column height.

The height of the column is a function of the number of theoretical stages and the efficiency of the actual stages. Once the operating conditions are established for a column, its diameter and height can be chosen using data available from tray and/or packing manufacturers.

10.3.3.7 Overall Heat Balance

The last step in the basic calculation scheme of the process design for a fractionating column is to make an overall heat balance around the fractionator. This is a very important step in determining the desirability and economic feasibility of some of the arbitrary assumptions[8] that were made in the design of the tower. The overall heat balance has also value in evaluation or troubleshooting existing fractionators. The overall energy balance may point out a control problem that exists or some other reason that a tower is operating improperly.

The overall heat balance around the fractionator can be written as follows.

$$Q_{\mathrm{R}} + Q_{\mathrm{C}} = h_{\mathrm{D}}D + h_{\mathrm{B}}B - h_{\mathrm{F}}F \tag{10-10}$$

where Q_{R} is reboiler heat duty, Q_{C} is condenser heat duty, h_{D} is enthalpy of distillate product, h_{B} is enthalpy of bottom product, h_{F} is enthalpy of feed, and D, B, F is rate of flow of distillate and bottom and feed streams, respectively.

The heat duty of the reboiler given earlier should be increased by the heat loss from the fractionator. The heat loss is about 3% of the reboiler duty for light hydrocarbon fractionators but can be estimated more accurately by using a heat-transfer coefficient of 1.0 Btu/hr/ft^2.°F and the

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[8] The actual variation in vapor and liquid streams in a fractionating column is determined by an enthalpy balance, and the limitation imposed by the assumption of constant molal overflow can be removed by a rigorous enthalpy balance used in conjunction with material balances and phase equilibria.

temperature difference between the column insulation and the atmospheric temperature. At this point, the results of all calculations in the process design of the fractionator should be reviewed and studied carefully. If all results appear satisfactory, then final calculation on column sizing may be complete.

10.3.4 Design Procedure

In order to determine the design parameters for a fractionating column, the following procedure is recommended (Campbell, 1992; GPSA, 1998).

1. Establish feed composition, flow rate, temperature, and pressure.
2. Make product splits for the column based on feed and desired product needs.
3. Establish a condenser temperature (for the overhead stream) and column pressure, which will be used for calculation of the reboiler temperature.
4. Calculate minimum number of theoretical trays and the minimum reflux ratio needed to produce the desired products.
5. Calculate the combination of actual theoretical trays and actual reflux ratio needed.
6. Make a heat balance around condenser to determine condenser duty and around the column to determine reboiler duty.
7. Size the column.

Traditionally, fractionating columns have been designed using the equilibrium stage approach. However, the correct and general approach to column design is not to use the equlibrium stage but to change, head on, the complete nonequilibrium problem by taking into account the heat and mass transfer processes on trays. The nonequilibrium stage approach must incorporate the proper description of interphase mass transfer. Also, simultaneous heat transfer effects need to be incorporated into the model formulation. The incentive to adopt the nonequlibrium stage approach is that column profile predictions using this approach and the traditional approaches can be markedly different. Such differences could have a significant effect on column design. However, the major bottleneck in the use of this approach is the lack of generally applicable mass transfer correlations for trays and packings. Hence, new design methods are still on the conservative side.

10.4 GASOLINE AND LPG TREATING

Natural gasoline (condensate) and LPG streams are often contaminated with acidic compounds such as hydrogen sulfide (H_2S), carbon dioxide (CO_2), carbonyl sulfide (COS), carbon disulfide (CS_2), mercaptans, and elemental sulfur. Especially objectionable are H_2S, mercaptans, and elemental sulfur. Natural gasoline containing H_2S has objectionable odor and is corrosive. Mercaptans give an objectionable odor to gasoline and elemental sulfur makes the gasoline corrosive.

H_2S in the LPG can result in the formation of free sulfur or mercaptans when used as feedstock. Mercaptans will impart strong odors to an LPG as well as its combustion products if they are present in any significant quantity. The presence of significant quantities of CO_2 can also cause problems by raising the vapor pressure of the LPG and lowering the heating value. COS and CS_2, although not corrosive in LPG, will hydrolyze slowly to H_2S in the presence of free water and cause the product to become corrosive (Bullin *et al.*, 1995).

A product containing the objectionable materials can be treated to remove the H_2S, COS, and elemental sulfur and to either remove the mercaptans or convert them to less objectionable compounds. H_2S and CO_2 can be removed from LPG and gasoline by liquid–liquid contacting processes using a caustic solution, aqueous alkanolamines, or solid KOH. When quantities of H_2S and CO_2 components are small, a simple caustic wash is both effective and economical. However, as the quantity of contaminants rises, the caustic supply and disposal costs render this approach impractical. Amine treating is a very attractive alternative, especially when there is already an amine gas treating unit on site. The design considerations for sweetening LPGs with amines are discussed in more detail by Bullin *et al.* (1995).

Elemental sulfur is removed from the gasoline by contacting it with a polysulfide wash solution. COS can, of course, be removed sacrificially by MEA (GPSA, 1998). Mercaptans can be converted to disulfides by several methods. These disulfides will remain in the sweetened hydrocarbon product. The overall sulfur content, therefore, remains the same. However, the sulfur leaves as disulfide (no odor) rather than mercaptans. The method or combination of methods that can be used depends on the mercaptan content of the product to be treated and the specification that must be met (Fischer *et al.*, 1993). Two common treating methods are discribed next.

10.4.1 Doctor Sweetening Process

Among the processes that convert (oxidize) mercaptans to disulfides, the "Doctor sweetening" process is the oldest. Doctor treatment ordinarily will leave 0.0004% mercaptan at which level there is no or negligible effect on the tetraethyl lead susceptibility of the gasoline. In this process, an alkaline solution of lead oxide (usually sodium plumbite) contacts the hydrocarbon stream forming lead mercaptides (soluble in the oil) with the mercaptans. The mixture is then treated with powdered sulfur (which has a high affinity for lead) and a conversion of the mercaptide into a so-called disulfide (which remains in solution in the gasoline stream) occurs. The reactions of the "Doctor treating" process are considered to be

$$2RSH + Na_2PbO_2 \rightarrow (RS)_2Pb + 2NaOH \tag{10-11}$$

$$(RS)_2Pb + S \rightarrow R_2S_2 + PbS \tag{10-12}$$

Note that sulfur should be added in stoichiometric excess in order to get maximum conversion of the mercaptides to disulfides, where too much excess sulfur will cause formation of polysulfides (Maddox, 1974). Sometimes with no sulfur added, but in the presence of atmospheric oxygen and sodium hydroxide solution, the same conversion [Equation (10-12)] occurs, but only slowly and not completely (McBryde, 1991).

10.4.2 Merox Process

The extraction Merox[9] process is used to treat end product streams by rendering any mercaptan sulfur compounds inactive. This process can be used for treating LPG, gasoline, and heavier fractions. The method of treatment is the extraction reaction of the sour feedstock containing mercaptans (RSH) with caustic soda (NaOH) in a single, multistage extraction column using high-efficiency trays designed by Universal Oil Products process division. The extraction reaction is shown by the following equation:

$$RSH + N_aOH \leftrightarrow N_aSR + H_2O \tag{10-13}$$

After extraction, the extracted mercaptans in the form of sodium mercaptides (NaSR) are catalytically oxidized (by Merox WS catalyst, which is

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[9] Merox stands for mercaptan oxidation.

dispersed in the aqueous caustic solution) to water-insoluble disulfide oils (RSSR), as shown in the following equation.

$$4\text{NaSR} + \text{O}_2 + 2\text{H}_2\text{O} \rightarrow 2\text{RSSR} + 4\text{NaOH} \tag{10-14}$$

The disulfide oil is decanted and sent to fuel or to further processing in a hydrotreater. The regenerated caustic is then recirculated to the extraction column.

The Merox solution gives a very high degree of removal of mercaptans in a liquid stream. If a more complete removal is desired, Merox also provides a fixed-bed catalytic conversion of mercaptans to disulfides.

Note that for treatment of light feedstocks such as LPG, no sweetening is required, as mercaptans are nearly completely removed by extraction, whereas other feedstocks containing higher molecular weight mercaptans (i.e., NGL) may require a combination of Merox extraction and sweetening (UOP, 2003).

REFERENCES

Arnold, K., and Stewart, M., "Surface Production Operations, Vol 2: Design of Gas-Handling Systems and Facilities," 2nd Ed. Gulf Professional Publishing, Houston, TX (1999).

Ballard, D., How to operate quick-cycle plants. *Hydrocarb. Proc. Petr. Refiner.* **44**(4), 131 (1965).

Brands, D.S., and Rajani, J.B., "Comparison between Low Temperature Separation (LTS) and SORDECO Process for Hydrocarbon Dewpointing." Paper presented at the GPA Europe Annual Conference, Amsterdam, The Netherlands (Sept. 2001).

Brown, G.G., *et al.*, *Trans.AIChE J.* **39**(1), 63 (1943).

Bullin, J.A., Polasek, J., and Rogers, J., "Design Considerations for Sweetening LPGs with Amines." Paper presented at the 74th GPA Annual Convention, Tulsa, OK (1995).

Campbell, J.M., "Gas Conditioning and Processing," 3rd Ed. Campbell Petroleum Series, Norman, OK (1992).

Cranmore, R.G., and Stanton, E., "Modern Petroleum Technology" (R.A. Dawe, ed.), Chapter 9. Wiley, New York (2000).

Daiminger, U., and Lind, W., Adsorption-based processes for purifying natural gas. *World Refining* **14**(7), 32–37 (2004).

Eduljee, H.E., Hydrocarbon Processing, 120 (Sept. 1976).

Elliot, D.G., Chen, J.J., Brown, T.S., Sloan, E.D., and Kidnay, A.J., The economic impact of fluid properties research on expander plants. *Fluid Phase Equilibria* **116**, 27–38 (1996).

Ewan, D.N., Laurence, J.B., Rambo, C.L., and Tonne, R., "Why Cryogenic Processing?" Proc. 54th GPA Annual Convention, Houston, TX (March 1975).

Fenske, M.R., Fractionation of straight-run Pennsylvania gasoline. *Ind. Eng. Chem.* **24**, 482–485 (1932).

Fischer, E., Goel, R., and Saunders, D., "Preliminary Process Selection for Natural Gas Liquid (NGL) Treating." Proc. 72th GPA Annual Convention, 248–256, San Antonio, TX (March 15–17, 1993).

Foglietta, J.H., "Dew Point Turboexpander Process, a Solution for High Pressure Fields." Paper presented at the IAPG 2004 Gas Conditioning Conference, Neuquen, Argentina (Oct. 18, 2004).

Gadalla, M., Jobson, M., and Smith, R., Increase capacity and decrease energy for existing refinery distillation columns. *Chem. Eng. Prog*. **99**(4) 44–50 (2003).

Geist, J.M., Refrigeration cycles for the future. *Oil Gas J*. **83**(5), 56–60 (1985).

Gilliland, E.R., *Ind. Eng. Chem.* **32**, 1101 (1940).

GPSA Engineering Data Book, 11th Ed. Gas Processors Suppliers Association, Tulsa, OK (1998).

Holcomb, D.E., *et al.*, *Ind. Eng. Chem.* **34**, 590 (1942).

Holm, J., Turboexpanders in energy saving processes. *Energy Prog*. **6**(3), 187–190 (1986).

Kirkbride, C.G., *Petroleum Refiner* **23**(9), 321 (1944).

Lee, R.J., Yao, J., and Elliot, D., Flexibility, efficiency to characterize gas-processing technologies. *Oil Gas J.* **97**(50), 90–94 (1999).

Lokhandwala, K.A., and Jacobs, M.L., "New Membrane Applications in Gas Processing." Paper presented at the 79th GPA Annual Convention, Atlanta, GA (March 2000).

Mackenzie, D.H., and Donnelly, S.T., Mixed refrigerants proven efficient in natural gas liquids recovery process. *Oil Gas J*. **83**(9), 116–120 (1985).

Maddox, R.N., "Gas and Liquid Sweetening," 2nd Ed. Campbell Petroleum Series, Norman, OK (1974).

Manning, F.S., and Thompson, R.E., "Oil Field Processing of Petroleum," Vol. 1. Pennwell Publishing Company, Tulsa, OK (1991).

McBryde, W.A.E., Petroleum deodorized: Early Canadian history of the "Doctor sweetening process." *Ann. Sci.* **48**, 103–111 (1991).

McCabe, W.L., and Thiele, E.W., *Ind. Eng. Chem.* **17**, 605 (1925).

McCabe, W.L., Smith, J.C., and Harriott, P., "Unit Operations of Chemical Engineering," 6th Ed. McGraw-Hill, New York (2001).

Mehra, Y.R., and Gaskin, T.K., Guidelines offered for choosing cryogenic or absorption for gas processing. *Oil Gas J.* **97**(9), 62–67 (1999).

Nasir, P., Sweet, W., Elliot, D., Chen, R., and Lee, R.J., "Enhanced NGL Recovery Process Selected for Neptune Gas Plant Expansion." Paper presented at the 82ndGPA Annual Convention, San Antonio, TX (March 9–12, 2003).

Parasons, P.J., and Templeman, J.J., Models performance leads to adsorption-unit modifications." *Oil Gas J.* **88**(26), 40–44 (1990).

Perry, J.H., "Chemical Engineers Handbook," 7th Ed. McGraw-Hill, New York (1997).

Pitman, R.N., Hudson, H.M., Wilkinson, J.D., and Cuellar, K.T., "Next Generation Processes for NGL/LPG Recovery." Paper presented at the 77th GPA Annual Convention, Dallas, TX (March 16, 1998).

Rojey, A., Jaffret, C., Cornot-Gandolph, S., Durand, B., Jullin, S., and Valais, M., "Natural Gas Production, Processing, Transport." Editions Technip, Paris, France (1997).

Russell, T., "Gas Processing Basics." Paper presented at the 80th GPA Annual Convention, San Antonio, TX (March 11, 2001).

Souders, M., and Brown, G.G., Fundamental design of absorbing and stripping columns for complex vapors. *Ind. Eng. Chem.* **24**, 519 (1932).

Spletter, K.G., and Adair, L., US gas processing profitability statistics. *Oil Gas J.* **99**(21), 54–59 (2001).

Tuttle, R., and Allen, K., "Treating System Selection for Fractionation Plants." Proc. 55th Ann. Conv. GPA., San Antonio, TX (March 1976).

Underwood, A.J.V., Fractional distillation of multicomponent mixtures. *Chem. Eng. Prog.* **44**(8), 603–614 (1948).

UOP, "Merox Process for Mercaptan Extraction." UOP 4223-3 Process Technology and Equipment Manual, UOP LLC, Des Plaines, IL (2003).

Walas, S.M., "Chemical Process Equipment: Selection and Design." Butterworth-Heinemann, Boston (1990).

CHAPTER 11

SALES GAS TRANSMISSION

11.1 INTRODUCTION

Natural gas continues to play a great role as a worldwide energy supply. In fact, major projects are being planned to move massive amounts of high-pressure sales gas from processing plants to distribution systems and large industrial users through large-diameter buried pipelines. These pipelines utilize a series of compressor stations along the pipeline to move the gas over long distances. In addition, gas coolers are used on the discharge side of the compressor stations to maintain a specified temperature of the compressed gas for pipeline downstream pressure drop reduction and to protect gas pipeline internal and external coating against deterioration due to high temperatures. This chapter covers all the important concepts of sales gas transmission from a fundamental perspective.

11.2 GAS FLOW FUNDAMENTALS

Optimum design of a gas transmission pipeline requires accurate methods for predicting pressure drop for a given flow rate or predicting flow rate for a specified pressure drop in conjunction with installed compression power and energy requirements, e.g., fuel gas, as part of a technical and economic evaluation. In other words, there is a need for practical methods to relate the flow of gas through a pipeline to the properties of both the pipeline and gas and to the operating conditions, such as pressure and temperature. Isothermal steady-state pressure drop or flow rate calculation methods for single-phase dry gas pipelines are the most widely used and the most basic relationships in the engineering of gas delivery systems (Beggs, 1984; Smith, 1990, Aziz and Ouyang, 1995). They also form the basis of other more complex transient flow calculations and network designs.

11.2.1 General Flow Equation

Based on the assumptions that there is no elevation change in the pipeline and that the condition of flow is isothermal, the integrated Bernoulli's equation is expressed by Equation (11-1) (Uhl, 1965, Schroeder, 2001):

$$Q_{\mathrm{sc}} = C\left(\frac{T_{\mathrm{b}}}{P_{\mathrm{b}}}\right) D^{2.5}\left(\frac{P_1^2 - P_2^2}{f\gamma_{\mathrm{G}} T_{\mathrm{a}} Z_{\mathrm{a}} L}\right)^{0.5} E \tag{11-1}$$

where Q_{sc} is standard gas flow rate, measured at base temperature and pressure, ft^3/day; T_{b} is gas temperature, base conditions, 519.6°R; P_{b} is gas pressure, base conditions, 14.7 psia; P_1 is inlet gas pressure, psia; P_2 is outlet gas pressure, psia; D is inside diameter of pipe, inches; f is Moody friction factor; E is flow efficiency factor; γ_{G} is gas specific gravity; T_{a} is average absolute temperature of pipeline, °R; Z_{a} is average compressibility factor; L is pipe length, miles; and C is 77.54 (a constant for the specific units used).

Although the assumptions used to develop Equation (11-1) are usually satisfactory for a long pipeline, the equation contains an efficiency factor, E, to correct for these assumptions. Most experts recommend using efficiency factor values close to unity when dry gas flows through a new pipeline. However, as the pipeline ages and is subjected to varying degrees of corrosion, this factor will decrease (Campbell *et al.*, 1992). In practice, and even for single-phase gas flow, some water or condensate may be present if the necessary drying procedure for gas pipeline commissioning is not adopted or scrubbers are not installed. This puts compression equipment at risk of damage and also allows localized corrosion due to water spots (wetting of the pipe surface). The presence of liquid products in the gas transmission lines can also cause drastic reduction in the flow efficiency factor. Typically, efficiency factors may vary between 0.6 and 0.92 depending on the liquid contents of the pipeline (Ikoku, 1984). As the amount of liquid content in the gas phase increases, the pipeline efficiency factor can no longer account for the two-phase flow behavior and two-phase flow equations must be used (Brill and Beggs, 1991; Asante, 2002).

Pipelines are usually not horizontal; however, as long as the slope is not too great, a correction for the static head of fluid (H_c) may be incorporated into Equation (11-1) as follows (Schroeder, 2001).

$$Q_{sc} = C\left(\frac{T_b}{P_b}\right) D^{2.5}\left(\frac{P_1^2 - P_2^2 - H_c}{L\gamma_G T_a Z_a f}\right)^{0.5} E \tag{11-2}$$

where

$$H_c = \frac{0.0375\, g(H_2 - H_1)P_a^2}{Z_a T_a} \tag{11-3}$$

and H_1 is inlet elevation, ft; H_2 is outlet elevation, ft; and g is gravitational constant, ft/sec^2.

The average compressibility factor, Z_a, is determined from the average pressure (P_a) and average temperature (T_a), where P_a is calculated from Equation (11-4) (Campbell *et al.*, 1992):

$$P_a = \frac{2}{3}\left[(P_1 + P_2) - \left(\frac{P_1 P_2}{P_1 + P_2}\right)\right] \tag{11-4}$$

where P_1 and P_2 are the upstream and downstream absolute pressures, respectively. The average temperature is determined by Equation (11-5).

$$T_a = \left[\frac{T_1 - T_2}{\ln\left(\dfrac{T_1 - T_S}{T_2 - T_S}\right)}\right] + T_S \tag{11-5}$$

In the Equation (11-5), parameter T_S is the soil temperature, and T_1 and T_2 are the upstream and downstream temperatures, respectively.

Having obtained P_a and T_a for the gas, the average compressibility factor can be obtained using Kay's rule and gas compressibility factor charts (Campbell *et al.*, 1992).

11.2.2 Friction Factor Correlations

The fundamental flow equation for calculating pressure drop requires a numerical value for the friction factor. However, because the friction factor, f, is a function of flow rate, the whole flow equation becomes implicit. To determine the friction factor, fluid flow is characterized by a dimensionless value known as the Reynolds number [Equation (11-6)].

$$N_{\mathrm{Re}} = \frac{\rho V D}{\mu} \tag{11-6}$$

where N_{Re} is Reynolds number, dimensionless; D is pipe diameter, ft; V is fluid velocity, ft/sec; ρ is fluid density, $\mathrm{lb_m/ft^3}$; and μ is fluid viscosity, $\mathrm{lb_m}$/ft.sec.

For Reynolds numbers less than 2000 the flow is considered laminar. When the Reynolds number exceeds 2000, the flow is characterized as turbulent. Note that in high-pressure gas transmission pipelines with moderate to high flow rates, only two types of flow regimes are observed: partially turbulent flow (smooth pipe flow) and fully turbulent flow (rough pipe flow). For gases, the Reynolds number is given by Equation (11-7) (Kennedy, 1993):

$$N_{\mathrm{Re}} = \frac{0.7105\, P_{\mathrm{b}} \gamma_{\mathrm{G}} Q_{\mathrm{sc}}}{T_{\mathrm{b}} \mu_{\mathrm{G}} D} \tag{11-7}$$

where D is pipe diameter, inches; Q_{sc} is gas flow rate, standard $\mathrm{ft^3}$/day; μ_{G} is gas viscosity, cp; P_{b} is base pressure, psia; T_{b} is base temperature, °R; and γ_{G} is gas specific gravity, dimensionless. For the gas industry, Equation (11-7) is a more convenient way to express the Reynolds number, as it displays the value proportionally in terms of the gas flow rate.

The other parameter in the friction factor correlation is pipe roughness (ε), which is often correlated as a function of the Reynolds number and the pipe relative roughness (absolute roughness divided by inside diameter). Pipe roughness varies considerably from pipe to pipe, and Table 11-1 shows the roughness for various types of new (clean) pipes. These values

Table 11-1
Pipe Roughness Value (Norsok Standard, 1996)

Type of pipe (new, clean condition)	ε (inches)
Carbon steel corroded	0.019685
Carbon steel noncorroded	0.001968
Glass fiber reinforced pipe	0.0007874
Steel internally coated with epoxy	0.00018 to 0.00035

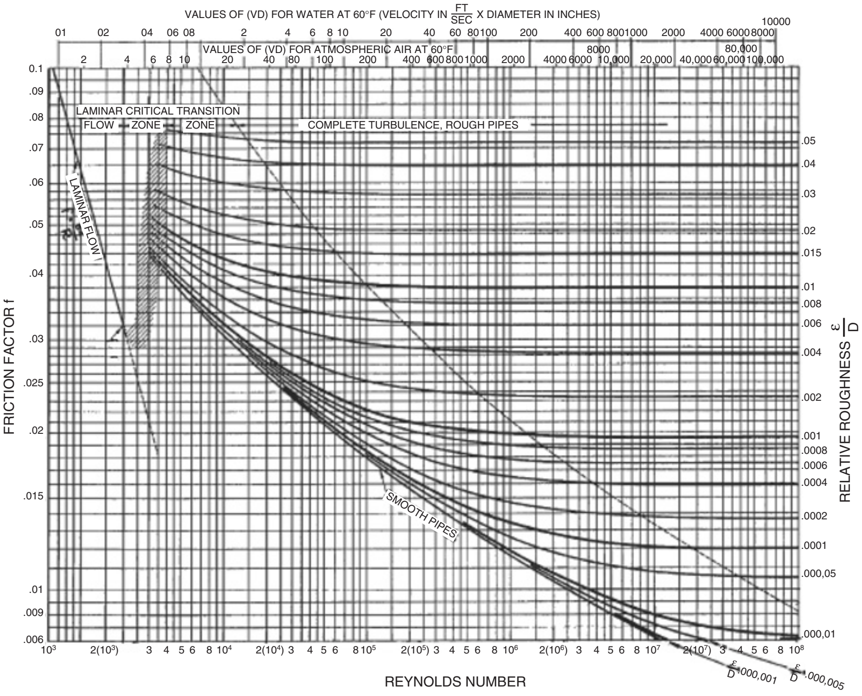


Figure 11-1. Moody friction factor diagram (Streeter and Wylie, 1979).

should be increased by a factor ranging between 2 and 4 to account for age and use.

The Moody (1944) friction factor, f, in Equation (11-1) is determined from the Moody diagram. The Moody correlation is shown in Figure 11-1. The Moody diagram consists of four zones: laminar, transition, partially turbulent, and fully turbulent zones.

The laminar zone, the left side, is the zone of extremely low flow rate in which the fluid flows strictly in one direction and the friction factor shows a sharp dependency on flow rate. The friction factor in the laminar regime is defined by the Hagen–Poiseuille equation (Streeter and Wylie, 1979):

$$f = \frac{64}{N_{\mathrm{Re}}} \tag{11-8}$$

The fully turbulent zone, the right side, describes fluid flow that is completely turbulent (back mixing) laterally as well as in the primary direction.

The turbulent friction factor shows no dependency on flow rate and is only a function of pipe roughness, as an ideally smooth pipe never really exists in this zone. The friction factor to use is given by the rough pipe law of Nikuradse (1933) [Equation (11-9)].

$$\frac{1}{\sqrt{f}} = 2\log\frac{D}{\varepsilon} + 1.14 \tag{11-9}$$

Equation (11-9) shows that if the roughness of the pipeline is increased, the friction factor increases and results in higher pressure drops. Conversely, by decreasing the pipe roughness, lower friction factor or lower pressure drops are obtained. Note that most pipes cannot be considered ideally smooth at high Reynolds numbers (Schlichting, 1979); therefore, the investigations of Nikuradse (1933) on flow through rough pipes has been of significant interest to engineers.

The partially turbulent (transition) zone is the zone of moderately high flow rate in which the fluid flows laterally within the pipe as well as in the primary direction, although some laminar boundary layer outside the zone of roughness still exists. Partially turbulent flow is governed by the smooth pipe law of Karman and Prandtl (Uhl, 1965):

$$\frac{1}{\sqrt{f}} = 2\log\left(N_{\mathrm{Re}}\sqrt{f}\right) - 0.8 \tag{11-10}$$

This correlation has received wide acceptance as a true representative of experimental results. However, a study by Zagarola (1996) on the flow at high Reynolds numbers in smooth pipes showed that the relevant correlation was not accurate for high Reynolds numbers, where the correlation was shown to predict too low values of the friction factor.

Consensus on how the friction factor varies across the transition region from an ideal smooth pipe to a rough pipe has not been reached. However, Colebrook (1939) presented additional experimental results and developed a correlation for the friction factor valid in the transition region between smooth and rough flow. The correlation is as follows:

$$\frac{1}{\sqrt{f}} = -2\log\left(\frac{\varepsilon/D}{3.7} + \frac{2.51}{N_{\mathrm{Re}}\sqrt{f}}\right) \tag{11-11}$$

This equation is universally accepted as standard for computing friction factor of rough pipes. Moody (1944) concluded that the Colebrook (1939) equation was adequate for friction factor calculations for any Reynolds number greater than 2000. Certainly the accuracy of the equation was well within the experimental error (about ±5% for smooth pipes and ±10% for rough pipes). However, the transcendental nature of the Colebrook (1939) equation is not conducive to simulation codes, as iteration is required. In addition, this equation is problematic to program, as convergence is dependent on a reasonable initial guess for the friction factor. This is not insurmountable, of course, but the complication should be avoided if possible. It is recommended that the modified Colebrook equation be used in practical engineering calculations for pipe flow in place of the classical equation at no loss and significant gain. Garland *et al*. (1999) provide additional details on this subject. Note that an explicit correlation for friction factor was presented by Jain (1976). This correlation is comparable to the Colebrook (1939) correlation and does not require iteration. For relative roughness between 0.000001 and 0.01 and the Reynolds number between 5×10^3 and 1×10^8, the errors were within 1% when compared with the Colebrook correlation. Therefore, it is usually recommended for all calculations requiring friction factor determination of turbulent flow.

The friction factor is sometimes expressed in terms of the Fanning friction factor, which is one-fourth of the Moody friction factor. Care should be taken to avoid inadvertent use of the wrong friction factor.

11.2.3 Practical Flow Equations

The Moody friction factor, f, is an integral part of the general gas flow equation. Because it is a highly nonlinear function, it must be either read from a chart or determined iteratively from a nonlinear equation. Approximations to the Moody friction factor have been widely used because they allow the gas flow equation to be solved directly instead of iteratively. The four most widely published friction factor approximations are Weymouth, Panhandle A, Panhandle B, and IGT (Beggs, 1984, Ikoku, 1984). The Weymouth equation approximates the Moody friction factor by Equation (11-12), and the remaining three equations approximate the friction factor by Equation (11-13), where “m” and “n” are constants. These constants are given in Table 11-2.

$$f = \mathrm{m}(D)^{-\mathrm{n}} \tag{11-12}$$

$$f = \mathrm{m}(N_{\mathrm{Re}})^{-\mathrm{n}} \tag{11-13}$$

Table 11-2
Constants in Equations (11-12) and (11-13)

Equation	m	n
Weymouth	0.032	0.333
Panhandle A	0.085	0.147
Panhandle B	0.015	0.039
IGT	0.187	0.200

The Reynolds number, N_{Re}, can be approximated by Equation (11-7). In addition to the Reynolds number, the pipe roughness also affects the friction factor for turbulent flow in rough pipes. Hence, the efficiency factor is chosen to correctly account for pipe roughness (Ouyang and Aziz, 1996).

These approximations can then be substituted into the flow equation for f, and the resulting equation is given by Equation (11-14) (Kennedy, 1993; Towler and Pope, 1994):

$$Q_{\mathrm{sc}} = \frac{\mathrm{a}_1 \left(\frac{T_{\mathrm{b}}}{P_{\mathrm{b}}}\right) E\left(P_1^2 - P_2^2\right)^{0.5} D^{\mathrm{a}_2}}{(\gamma_{\mathrm{G}})^{\mathrm{a}_3} (T_{\mathrm{a}} Z_{\mathrm{a}} L)^{\mathrm{a}_4} (\mu_{\mathrm{G}})^{\mathrm{a}_5}} \tag{11-14}$$

In Equation (11-14) the values for a_1 through a_5 are constants that are functions of the friction factor approximations and the gas flow equation. These constants are given in Table 11-3.

Inspection of Table 11-3 shows that the gas flow rate is not a strong function of the gas viscosity at high Reynolds numbers because viscosity is of importance in laminar flow, and gas pipelines are normally operated in the partially/fully turbulent flow region. However, under normal

Table 11-3
Constants in Equation (11-14)

Equation	a_1	a_2	a_3	a_4	a_5
Weymouth	433.46	2.667	0.5000	0.5000	0.0000
Panhandle A	403.09	2.619	0.4603	0.5397	0.0793
Panhandle B	715.35	2.530	0.4900	0.5100	0.0200
IGT	307.26	2.667	0.4444	0.5556	0.1111

conditions, the viscosity term has little effect because a 30% change in absolute value of the viscosity will result in only approximately a 2.7% change in the computed quantity of gas flowing. Thus, once the gas viscosity is determined for an operating pipeline, small variations from the conditions under which it was determined will have little effect on the flow predicted by Equation (11-14) (Huntington, 1950).

Note that all of the equations noted earlier have been developed from the fundamental gas flow equation; however, each has a special approximation of the friction factor to allow for an analytical solution. For instance, the Weymouth (1912) equation uses a straight line for f, and thus, its approximation has been shown to be a poor estimation for the friction factor for most flow conditions (Kennedy, 1993). This equation tends to overpredict the pressure drop, and thus provides a poor estimate relative to the other gas flow equations. The Weymouth equation, however, is of use in designing gas distribution systems in that there is an inherent safety in over predicting pressure drop (Maddox and Erbar, 1982). In practice, the Panhandle equations are commonly used for large-diameter, long pipelines where the Reynolds number is on the flat portion of the Moody diagram. The Panhandle "A" equation is most applicable for medium to relatively large diameter pipelines (12″–60″ diameter) with moderate gas flow rate, operating under medium to high pressure (800–1500 psia). The Panhandle "B" equation is normally appropriate for high flow rate, large-diameter (>36″), and high-pressure (>1000 psi) transmission pipelines (Maddox and Erbar, 1982; Kennedy, 1993).

Because friction factors vary over a wide range of Reynolds number and pipe roughness, none of the gas flow equations is universally applicable. However, in most cases, pipeline operators customize the flow equations to their particular pipelines by taking measurements of flow, pressure, and temperature and back calculating pipeline efficiency or an effective pipe roughness.

11.3 PREDICTING GAS TEMPERATURE PROFILE

Predicting the pipeline temperature profile has become increasingly important in both the design and the operation of pipelines and related facilities. Flowing gas temperature at any point in a pipeline may be calculated from known data in order to determine (1) location of line heaters for hydrate prevention (Towler and Mokhatab, 2004), (2) inlet gas temperature at each compressor station, and (3) minimum gas flow rate required to maintain a specific gas temperature at a downstream point. To predict the temperature profile and to calculate pressure drop accurately, it is necessary to divide the pipeline into smaller segments. The temperature change calculations are iterative, as the temperature (and pressure) at each point must be known to calculate the energy balance. Similarly, the pressure loss calculations are iterative, as the pressure (and temperature) at each point along the pipeline must be known in order to determine the phase physical properties from which the pressure drop is calculated. Thus, generating a usable temperature profile requires a series of complex, interactive type calculations for which even the amount of data available in most cases is insufficient. Additionally, the pipeline outer environment properties such as soil data and temperatures vary along the pipeline route and therefore play a very important role and require a consistent modeling to provide a reliable temperature profile evaluation. A simple and reasonable approach is to divide the pipeline into sections with defined soil characteristics and prevailing soil temperatures for summer and winter time and then to calculate the overall heat transfer that will be highly influenced by the outer conditions. The complexity of this method has led to the development of approximate analytical methods for the prediction of temperature profile, which in most situations are satisfactory for engineering applications.

Basic relationships needed for these calculation methods are thermal, mechanical energy balances, and mass balance for the gas flow in pipelines. The general or thermal energy balance can be written as follows (Buthod *et al*., 1971):

$$C_{\mathrm{P}}\left(\frac{dT}{dx}\right) - \eta C_{\mathrm{P}}\left(\frac{dP}{dx}\right) + \left(\frac{VdV}{g_{\mathrm{c}}}\right)\left(\frac{dV}{dx}\right) + \left(\frac{g}{g_{\mathrm{c}}}\right)\left(\frac{dH}{dx}\right) = -\frac{dq}{dx} \tag{11-15}$$

where T is gas temperature, P is absolute pressure of gas, V is gas linear velocity in the pipeline, q is heat loss per unit mass of flowing fluid, C_{P} is constant pressure-specific heat, η is Joule–Thomson coefficient, H is height above datum, x is distance along pipeline, g_{c} is conversion factor, and g is gravitational acceleration. The major assumption in the development of Equation (11-15) is that the work term is zero between the compressor stations.

To calculate heat transfer from the pipe to the ground (soil), per unit of pipeline length, the Kennelly equation, Equation (11-16), is used (Neher, 1949):

$$\frac{dq}{dx} = [2\pi K(T - T_{\mathrm{S}})]/m_{\mathrm{G}}\ \mathrm{Ln}\left[(2\mathrm{H}' + \sqrt{4\mathrm{H}'^{2} - \mathrm{D}_0^2})/\mathrm{D}_0\right] \qquad (11\text{-}16)$$

where K is thermal conductivity of soil, T_{S} is undisturbed soil temperature at pipe centerline depth, m_{G} is mass flow rate of gas, H' is depth of burial of pipe (to centerline), and D_0 is outside diameter of pipe.

A basic assumption in Equation (11-16) is that the temperature of the gas is the same as the temperature of the pipe wall (the resistances to heat transfer in the fluid film and pipe wall are negligible).

The mechanical energy balance is given by Equation (11-17) (Streeter and Wylie, 1979):

$$\frac{1}{\rho}\frac{dP}{dx} + \frac{VdV}{g_{\mathrm{c}}dx} + \frac{2fV^2}{g_{\mathrm{c}}\mathrm{D}_{\mathrm{i}}} = 0 \qquad (11\text{-}17)$$

where ρ is density of gas, f is Fanning friction factor, and D_{i} is inside diameter of pipe.

The continuity equation, Equation (11-18), relates velocity to the pressure and temperature (Buthod *et al.*, 1971):

$$\frac{dV}{dx} = \frac{-m_{\mathrm{G}}}{\rho^2\mathrm{A}}\left[\frac{\rho}{P} - \frac{\rho}{Z}\left(\frac{\partial Z}{\partial P}\right)_{\mathrm{T}}\right]\frac{dP}{dx} - \left[\frac{\rho}{T} + \frac{\rho}{Z}\left(\frac{\partial Z}{\partial T}\right)_{\mathrm{P}}\right]\frac{dT}{dx} \qquad (11\text{-}18)$$

where A is inside cross-sectional area of pipe and Z is gas compressibility factor.

Equations (11-15) to (11-18) are the basic equations that must be solved simultaneously for calculation of the gas temperature and pressure profiles in pipelines. Details for solving this set of equations can be found in textbooks on numerical analysis, such as Constantinides and Mostoufi (1999).

The typical equations used to determine pipeline temperature loss are the integrated form of the general equations. However, assumptions and simplifications must be made to obtain the integrated equations, even though the effects of these assumptions or simplifications are not always known. A major advantage of numerical integration of the differential equation is that fewer assumptions are necessary. Considering this fact, Coulter and Bardon (1979) have presented an integrated equation as follows:

$$T_{\mathrm{x}} = \left\{T_1 - \left[T_{\mathrm{S}} + \left(\frac{\eta}{\mathrm{a}}\right)\left(\frac{dP}{dx}\right)\right]\right\} e^{-\mathrm{ax}} + \left[T_{\mathrm{S}} + \left(\frac{\eta}{\mathrm{a}}\right)\frac{dP}{dx}\right] \qquad (11\text{-}19)$$

where T_1 is the inlet gas temperature and the term "a" is defined as

$$\mathrm{a} = \frac{2\pi \mathrm{R} U}{m_{\mathrm{G}} C_{\mathrm{P}}} \qquad (11\text{-}20)$$

where R is pipe radius and U is overall heat transfer coefficient.

Equation (11-19) can be used to determine the temperature distribution along the pipeline, neglecting kinetic and potential energy and assuming that heat capacity at constant pressure, C_{P}, and the Joule–Thomson coefficient remain constant along the pipeline. For most practical purposes, these assumptions are close to reality and generally do give quite good results. Moreover, for a long gas transmission pipeline with a moderate to small pressure drop, the temperature drop due to expansion is small (Buthod *et al.*, 1971) and Equation (11-19) simplifies to Equation (11-21):

$$T_{\mathrm{x}} = T_{\mathrm{S}} + (T_1 - T_{\mathrm{S}})\exp(-\mathrm{ax}) \qquad (11\text{-}21)$$

Equation (11-21) does not account for the Joule–Thomson effect, which describes the cooling of an expanding gas in a transmission pipeline. Hence, it is expected that the fluid in the pipeline will reach soil temperature later than that predicted by Equation (11-19).

While there has been extensive effort in the development of such pipelines, little attention has been paid to the fact that the Joule–Thomson coefficient and heat capacity at constant pressure are not constants. However, a new analytical technique for the prediction of temperature profile of buried gas pipelines has been developed by Edalat and Mansoori (1988), while considering the fact that η and C_{P} are functions of both temperature and pressure. Readers are referred to the original reference for a detailed treatment of this method.

Example 11-1

A 0.827 specific gravity gas flowing at 180 MMscfd is transferred through a 104.4-mile horizontal pipeline with a 19-inch internal diameter. The inlet pressure and temperature are 1165 psia and 95°F, respectively. The desired exiting pressure is 735 psia. Assuming an overall heat transfer coefficient of 0.25 $\text{Btu/hr.ft}^2.°\text{F}$, Joule–Thomson coefficient of 0.1093°F/psi, gas heat capacity (C_P) of 0.56 $\text{Btu/lb}_m°\text{F}$, and soil temperature of 60°F, how far will the gas travel before its temperature reaches the hydrate formation temperature? (Note that these conditions are very rare, as all export gas is dried to avoid hydrates. Hydrates could form in an export gas pipeline only when the dehydration unit breaks or the inhibitor pump breaks.)

Solution

Using the given data:

$$\rho_G = \frac{P \times MW}{ZRT} = \frac{1165 \times 0.827 \times 28.96}{0.776 \times 10.7316 \times 554.6} = 6.041 \frac{\text{lb}_m}{\text{ft}^3}$$

$$Q = Q_{SC}\left(\frac{P_b}{P}\right)\left(\frac{T}{T_b}\right)Z = \left(180 \times 10^6\right)\left(\frac{14.7}{1165}\right)\left(\frac{554.6}{519.6}\right)(0.776) \times \left(\frac{1}{86400}\right) = 21.773 \frac{\text{ft}^3}{\text{sec}}$$

$$m_G = (6.041)\,(21.773) = 131.530 \frac{\text{lb}_m}{\text{sec}}$$

$$a = \frac{(\pi \times 1.5833 \times 0.25) \times 5280}{(131.530 \times 0.56 \times 3600)} = 0.0247\,\text{miles}^{-1}$$

Solving Equation (11-19) gives $T_x = 41.82 + 53.18\,\exp[-0.0247\,x]$

By simultaneously using the just-described equation and the Katz gravity chart (Figure 2-14), for prediction of hydrate formation temperature, the location of hydrate formation is determined. In this case, the hydrate formation temperature is 70.5°F. The first location where hydrate formation occurs is 24.94 miles along the pipeline. However, Equation (11-21) is often used for most practical gas transmission pipelines. Using the solution with these parameters becomes $T_x = 60 + 35\,\exp\,[-0.0247\,x]$. Based on this equation, the first location of hydrate formation is 48.62 miles along the line, which means Equation (11-21) predicts that the fluid in the pipeline will form hydrate much later than that predicted by Equation (11-19).

It should be noted that in this example the temperature profile is strongly affected by both the heat transfer coefficient, U, and the Joule–Thomson coefficient, η. According to Equation (11-19), heating would be required from 24.94 miles until 104.4 miles to prevent hydrate formation. This result depends strongly on U, η, and the soil temperature T_S. If the pipe was insulated more strongly and U was reduced to 0.1 Btu/hr.ft^2, the point where line heating is required would be extended out to 36.67 miles but heating would still be required. It is also of note that the outlet temperature would drop to 45.87°F when $U = 0.25$ Btu/hr.ft^2.°F and to 43.27°F when $U = 0.1$ Btu/hr.ft^2.°F, both of which are below the outside soil temperature. This result may seem surprising to the reader but it occurs due to the large pressure drop in the pipeline. In this case, line insulation is exacerbating the problem because it does not allow the soil to retard the Joule–Thompson cooling effect. Equation (11-21) is not able to account for this effect at all.

In order to prevent the formation of hydrates, the temperature must be such that no point in the pipeline is in the region where a hydrate will form. A common technique to avoid hydrate formation in onshore gas transmission pipelines is thermal stimulation. Thermal stimulation involves the use of a source of heat applied directly in the form of injected steam or hot liquids or indirectly via electric means. The addition of heat raises the temperature of the pipeline and forces hydrates to decompose. The direct approach works well during steady-state conditions but is of no benefit in certain transient or shut-in scenarios (Lervik, 1998). However, indirect heat, such as the installation of line heaters for onshore gas transmission systems, is feasible for these transient conditions. This technology is particularly applicable to transmission and distribution systems that operate in cold climates. The other method to avoid/prevent hydrate formation is insulation. In fact, the proper use of insulation may, in some case, negate the requirement for a heater altogether (Carroll, 2003).

11.4 TRANSIENT FLOW IN GAS TRANSMISSION PIPELINES

Pipeline operation is such that transmission pipeline gas flow exists in the unsteady state, primarily due to variations in demand, inlet and outlet flows change, compressor start and stop, control setpoints, etc. In fact, steady-state operation is a rarity in practice. The unsteady nature of the gas flow indicates the need for a useful transient flow pipeline model to represent such conditions. In other words, the model should solve time-dependent flow equations. When modeling lines, however, it is sometimes convenient to make the simplifying assumption that flow is isothermal and steady state as long as we incorporate a load or swing factor contingent to a latter transient design checking to prevent inadequate pipeline sizing (Santos, 1997). The steady-state models are widely used to design pipelines and to estimate the flow and line pack. However, there are many situations where an assumption of steady-state flow and its attendant ramifications produce unacceptable engineering results. Unsteady state flow of gas in transmission lines can be described by a one-dimensional approach and by using an equation of state, continuity equation, momentum, and energy equations. In practice, the form of the mathematical relationships depends on the assumptions made based on the operating conditions of the pipelines. For the case of slow transient flows due to fluctuations in demand, it is assumed that the gas in the line has sufficient time to reach thermal equilibrium with its constant temperature surroundings. Similarly, for the case of rapid transient flows, it is assumed that the pressure changes occur instantaneously, allowing no time for heat transfer between the gas in the pipeline and the surroundings. However, for this case, heat conduction effects cannot be neglected. Streeter and Wylie (1970) have presented different methods that provide an accurate means of simulating unsteady state flow in gas transmission lines. For a detailed description of these method variations, which are beyond the scope of present discussion, readers are referred to the original paper.

11.5 COMPRESSOR STATIONS AND ASSOCIATED PIPELINE INSTALLATIONS

Compressor stations are installed along the pipeline to boost the gas pressure in the pipeline to increase the pipeline capacity in order to meet the gas demand made by the users. Compressor stations comprise one or more compressor units, each of which will include a compressor and driver together with valves, control systems and exhaust ducts and noise attenuation systems. Each compressor station will have inlet filters or knock out vessels to protect the compressors from damage due to liquids and entrained particles. In addition to compressor stations, there may be gas injection and delivery points along the line where the pressure and flow will have to be monitored and controlled. Each of these locations will include pressure control facilities and flow measurement.

11.5.1 Compressor Drivers

Transmission systems have high volume flow and the compressor stations generally have low head. Centrifugal compressors are ideal for these low-head and high-volume applications. Centrifugal compressors are also high-speed machines and ideally should have high-speed drivers. Choices for drivers can be gas turbines, gas engines, and electric motors. The selection is usually based on considerations of cost, both capital and maintenance, fuel or energy cost, reliability, and availability. Dealing with each of these in turn, gas engines are low speed and therefore will require a gearbox to connect to the compressor. Gas engines are really not competitive with other drivers in terms of installed cost at the power levels demanded by large-diameter, high-pressure pipeline transmission systems. Gas turbines are high-speed machines and can be directly coupled to the compressor and of course are available speed drives. Electric motors can be of several types with both fixed and variable speed options. Variable speed drivers (VSD) with electric motors present an overall performance much better than gas turbines and their selection depends on the site logistics and availability of electric energy at a reasonable price. They are very competitive in installed cost with comparable gas turbines. VSDs offer very low maintenance costs, rapid starting, lower noise levels, and no CO_2 emission.

The decision to use gas turbine or electric motor drivers is almost always based on the site logistics, cost, availability, and reliability of the energy source. There will always be gas in the pipeline so the question of reliability and availability of the energy source for the gas turbine does not enter into the question. For the electric drive, there has to be a reliable electric grid within a short distance from the compressor station, transmission lines are costly, and there should be a backup independent feed to guard against line failure. Given that these conditions are satisfied, the decision then comes down to a matter of fuel versus electricity cost, overall performance, maintenance, and operation cost, which is usually demonstrated with a life cycle cost calculation and economic evaluation. The life cycle cost must examine and test the results for sensitivity to cost escalation in power prices and gas price, taking into account the correlation between these two commodities. A long-term power supply agreement would be required to mitigate risk. The question of using a diesel engine as a power source as not been considered as it offers no advantages over a similar gas engine and it introduces another fuel, which invokes additional costs for transportation and storage.

11.5.2 Compressors Configurations

Because gas pipeline projects demand high amounts of capital expenditure and therefore are involved with investments risks, the project sponsor will try to maximize the capacity usage and minimize investment so as to have a competitive transportation rate to offer to the market. At the same time it will avoid the pipeline to operate with spare capacity. The decision on the compressors arrangement whether series or parallel is mostly based on economics and on simulation of failure analysis (Santos, 2000).

11.5.3 Reduction and Metering Stations

Each reduction and metering station branches off the pipeline and is used to reduce pressure and meter the gas to the various users. For the reduction and metering stations the main equipment includes filters, heaters, pressure reduction and regulators, and flow metering skids. In addition, each station is generally equipped with drains collection and disposal, instrument gas system, and storage tanks.

11.5.3.1 Filters

Natural gas filter units are installed at each station to remove any entrained liquids and solids from the gas stream. The filters may comprise cyclonic elements to centrifuge particles and liquids to the sides of the enclosing pressure vessel. These particles and liquids will then drop down for collection in a sump, which can be drained periodically.

11.5.3.2 Heaters

Natural gas heaters are installed to avoid the formation of hydrates, liquid hydrocarbons, and water as a result of pressure reduction. The gas heater is designed to raise the temperature of the gas such that after pressure reduction, the temperature of the gas will be above the dew point temperature at operating conditions and maximum flow. The heater is a water bath natural circulation type maintained at a temperature between 158 and 176°F. Where gas cost is high, an alternative is to use high-efficiency or condensing furnaces for the purpose of preheating the gas rather than the water bath heater.

11.5.3.3 Pressure Reduction and Regulation System

The pressure reduction system controls the supply pressure to the gas users at a regulated value. Each system consists of at least two trains of pressure reduction, one operating and the other standby. Each train will normally comprise two valves in series, one being the “active valve” and the other the “monitor valve.” Each valve will be equipped with a controller to operate the valve to maintain the preset discharge pressure.

11.5.3.4 Metering System

The flow rate of the gas has to be measured at a number of locations for the purpose of monitoring the performance of the pipeline system and more particularly at places where “custody transfer” takes place, which is where gas is received from the supply source and gas is sold to the customer for distribution. Depending on the purpose for metering, whether for performance monitoring or for sales, the measuring techniques used may vary according to the accuracy demanded. Typically, a custody transfer metering station will comprise one or two runs of pipe with a calibrated metering orifice in each run.

11.6 DESIGN CONSIDERATIONS OF SALES GAS PIPELINES

The typical design of a gas transmission pipeline involves a compromise among the pipe diameter, compressor station spacing, fuel usage, and maximum operating pressure. Each of these variables influences the overall construction and operating cost to some degree, hence an optimized design improves the economics of the construction and operation of the system and the competitiveness of the project.

11.6.1 Line Sizing Criteria

The pipe size generally is based on the acceptable pressure drop, compression ratio, and allowable gas velocities. Acceptable pressure drop in gas transmission pipelines must be one that minimizes the size of the required facilities and operating expenses such as the pipe itself, the installed compression power, the size and number of compressors, and fuel consumption. In fact, a large pressure drop between stations will result in a large compression ratio and might introduce poor compressor station performance. Experience has shown that the most cost-effective pipeline should have a pressure drop in the range between 3.50 and 5.83 psi/mile (Hughes, 1993). However, for those pipelines (short ones) in which pressure drop is of secondary importance, the pipe could be sized based on fluid velocity only. The flow velocity must be kept below maximum allowable velocity in order to prevent pipe erosion, noise, or vibration problems, especially for gases that may have a velocity exceeding 70 ft/sec. In systems with CO_2 fractions of as low as 1 to 2%, field experiences indicate that the flow velocity should be limited to less than 50 ft/sec because it is difficult to inhibit CO_2 corrosion at higher gas velocities (Kumar, 1987).

In most pipelines, the recommended value for the gas velocity in the transmission pipelines is normally 40 to 50% of the erosional velocity (Mohitpour *et al.*, 2002). As a rule of thumb, pipe erosion begins to occur when the velocity of flow exceeds the value given by Equation (11-22) (Beggs, 1984):

$$V_e = \frac{C}{\rho_G^{0.5}} \tag{11-22}$$

where V_e is erosional flow velocity, ft/sec; ρ_G is density of the gas, lb/ft^3; and C is empirical constant.

In most cases, C is taken to be 100. However, API RP 14E (1984) suggested a value of $C = 100$ for a continuous service and 125 for a noncontinuous service. In addition, it suggests that values of C from 150 to 200 may be used for continuous, noncorrosive or corrosion controlled services if no solid particles are present.

After selecting the appropriate inside diameter for a pipe, it is necessary to determine the pipe outside diameter (wall thickness), which would result in the minimum possible fabrication cost while maintaining pipeline integrity.

Example 11-2

Given the following data for a pipe segment of a gas transmission line, calculate the erosional velocity, neglecting the gas viscosity effect.

$Q_{sc} = 25.7\,\text{MMscfd}$ $P_1 = 425\,\text{psia}$

$T_a = 90°\text{F}$ $L = 8280\,\text{ft}$

$\gamma_G = 0.7$ $D = 12\,\text{inch}$

$Z_a = 0.925$ $E = 1.0$

Solution

1. The outlet gas pressure is computed using the Panhandle "A" equation:

$$25.7 \times 10^6 = \frac{\left[403.09\left(\frac{519.6}{14.7}\right)\left(P_1^2 - P_2^2\right)^{0.5}(12)^{2.619}\right]}{\left(0.7^{0.4603}\right)\left(549.6 \times 0.925 \times \frac{8280}{5280}\right)^{0.5397}}$$

$$P_1^2 - P_2^2 = 7062.765, \text{ therefore } P_2 = 416.608 \text{ psia}$$

2. Calculate the average pressure and gas density:

$$P_a = \frac{2}{3}\left[(425 + 416.608) - \frac{425 \times 416.608}{(425 + 416.608)}\right] = 420.817\,\text{psia}$$

$$\rho_G = \frac{P_a \times MW}{Z_a R T_a} = \frac{420.817\,(0.7 \times 28.96)}{0.925 \times 10.731 \times 549.6} = 1.563\,\frac{\text{lb}_m}{\text{ft}^3}$$

3. The erosional velocity is computed using API RP 14E equation [Equation (11-22)]. Continuous service ($C = 100$) is assumed.

$$V_e = \frac{100}{(1.563)^{0.5}} = 79.971\,\frac{\text{ft}}{\text{sec}}$$

4. Check the gas velocities through the line to ensure that excessive erosion will not occur.

The actual gas flow rate is calculated as

$$Q = Q_{\mathrm{SC}}\left(\frac{P_{\mathrm{b}}}{P_{\mathrm{a}}}\right)\left(\frac{T_{\mathrm{a}}}{T_{\mathrm{b}}}\right)Z_{\mathrm{a}}$$

$$= 25.7 \times 10^6\left(\frac{14.7}{420.817}\right)\left(\frac{549.6}{519.6}\right)(0.925) = 10.166\ \frac{\mathrm{ft}^3}{\mathrm{sec}}$$

Therefore, the gas velocity is

$$V_{\mathrm{G}} = \frac{10.166}{\left(\dfrac{\pi D^2}{4}\right)} = 12.950\ \mathrm{ft/sec}$$

The gas velocity is below the erosional velocity, so erosion should not be expected. However, it is high enough to prevent solids from settling.

11.6.2 Compressor Station Spacing

In long-distance gas transmission systems with a number of operating compressors, there is a definite need to optimize the spacing between compressor stations. Compressor station spacing is fundamentally a matter of balancing capital and operating costs at conditions, which represent the planned operating conditions of the transmission system. The process can become somewhat involved and lengthy, particularly as the selection of spacing needs to be designed in such a way to address a capacity ramp-up scenario that will cover not only the initial condition, but also the future years associated with the economics of the pipeline. In case of unexpected growth opportunities we can also rely on loop lines, which may be a better additional choice to increase capacity even more.

For a given pipe diameter, the distance between compressor stations may be computed from the gas flow equation, assuming a value of pipeline operating pressure (station discharge pressure) and a next compressor station suction pressure limited to the maximum compression ratio adopted for the project. Ideally, the pipeline should operate as close to maximum allowable operating pressure (MAOP) as possible, as high density in the line of the flowing gas gives best efficiency. This would point to the selection of close compressor station spacing, but this approach would not be the best economical decision. A decision based on the pipeline economics is the recommended one.

Based on the required gas flow, an initial diameter is assumed that results in a reasonable compression ratio (usually around 1.3–1.4 for transmission lines) and gas velocity, and the compressor station spacing is established by setting the maximum discharge pressure at the MAOP. Other diameters are tested, and compressor station spacing calculations are performed again. The optimum diameter is determined based on minimizing capital cost and operating cost, resulting in a chart (the so-called J curves because of their shape) that will plot transportation ratio in US$/MMBTU against transportation capacity based on predefined economic assumptions and risks (Santos and Saliby, 2003). Such assumptions include the design life of the facility, the required rate of return on capital employed, and the discount factor used to express the annual operating expenses incurred over time to a present value. The total cost is then plotted against compressor-discharge pressure, and a point of discharge pressure corresponding to the minimum total cost is picked as the best operating pressure.

Note that a good approach is to design for maximum capacity and define the required number of compressor stations and their spacing and after that going down on capacity on each operating year and take stations off as required. This approach will allow a better design and helps defining equipment that would equal. This also allows defining the installation schedule for the compressor stations and compressor units as well.

Capital expenditure (CAPEX) includes costs such as pipe, valves, fittings, compressors, turbines (or electric motors), control and construction, and assembly costs. Operating expenditure (OPEX) includes all maintenance and supervision and fuel or energy cost. CAPEX can be derived from past experience and databases. OPEX has to be estimated based on the specific project and past experience. The most significant part of operating costs is fuel or energy and equipment overhaul. Fuel cost is directly related to compressor horsepower. In order to illustrate how compressor station spacing influences the economics of pipeline operation, a simple model can be set up. This hypothetical pipeline model is based on a system 1000 miles long, operating at a maximum pressure of 1000 psig and flowing 1000 MMscfd. Assuming a uniform pressure loss per unit length of the pipe and station spacing, the inlet pressure at the first station downstream can be calculated and the horsepower needed to bring the pressure up to the discharge pressure setpoint. The process is repeated and the total power needed is the sum of all the stations. Figure 11-2 shows the manner in which total power required increases with spacing of stations. Figure 11-3 based on the same data shows horsepower in relation to number of stations. We need to keep in mind that even if we have an increase in required power for a pipeline with less

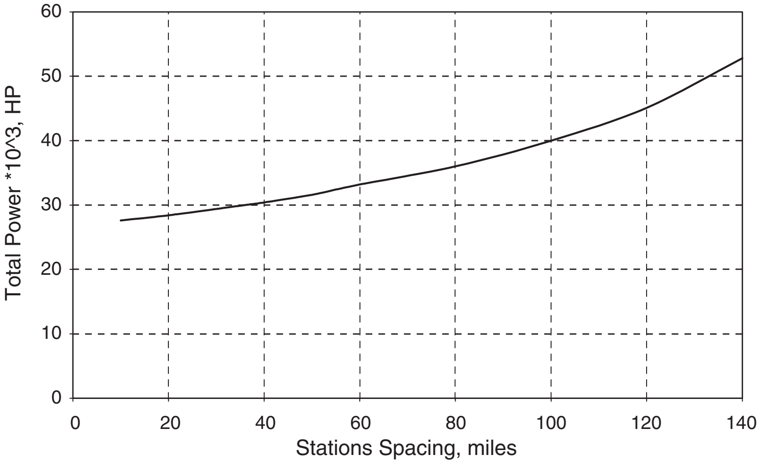


Figure 11-2. Effect of the compressor station spacing on total power (Cleveland and Mokhatab, 2005).

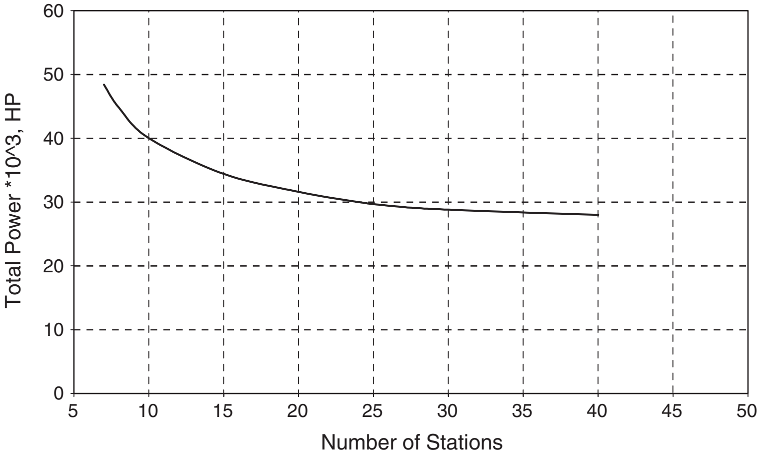


Figure 11-3. Effect of the number of compressor stations on total power (Cleveland and Mokhatab, 2005).

compressor stations, the overall cost tends to be lower than many compressor stations with lower horsepower requirements. The installed cost per horsepower will be lower for larger compressor units, and thermodynamic performance will be much better pointing to the direction of having fewer compressor stations with larger compressor units. This explains why an economic evaluation has to be done for each project configuration, taking into consideration all related information in terms of CAPEX and OPEX.

Because fuel use is related to horsepower, a minimum operating cost is associated with close compressor station spacing, which is logical, as maximum transmission efficiency is obtained at the highest mean line pressure, although the pipeline will have a larger diameter, requiring higher CAPEX. However, the optimum is influenced by two factors, the first that small turbine compressor units are less efficient than large ones (have a higher specific fuel consumption), although this effect is small and almost negligible at unit powers above 20,000 horsepower. A much greater impact though lies in the cost of the stations, and this capital cost declines as the number of stations is reduced (not linear as large stations are proportionately more costly than small ones). This tends to move the optimum spacing away from the minimum distance shown in Figure 11-2.

Every project has to be considered individually because of specific factors and the relationship between CAPEX and OPEX that will differ, but the general conclusion that close spacing results in greater transmission efficiency although may not be the best economic selection.

Experience shows that large units (compressor stations) are more efficient than smaller ones, as larger centrifugal compressors and gas turbines have better efficiency. However, the impact of unit outage or failure must be simulated under a transient analysis (Santos, 1997) so that we can define the remaining capacity and therefore establish a maintenance criterion in terms of having standby units or spare equipment to allow quick replacing without affecting the contractual obligation in terms of transportation capacity.

When the preferred solution is found, it must be tested for robustness not only over the range of throughput anticipated, but also for all credible upset conditions. Having developed the optimum solution the result should then be applied to the practical case with elevation changes and other local factors, including the availability of sites, all of which may result in adjustments and minor changes.

11.6.3 Compression Power

The next step in the design of a pipeline system is to calculate the maximum power required at the stations and set the design point(s). Typically, a new pipeline system will grow from a low flow condition to the maximum over a period of several years, and the decisions on compressors and drivers have to take these changing conditions into account. Growth of flow can be accommodated in several ways. Initially, compressors may only be installed at alternate stations and the intermediate stations built as the growth of flow dictates. Another option is to install one unit at each station location and then add units at the stations as flow increases. On the design phase the capacity ramp up will determine the installation schedule for the compressor station and also the additional units that will be necessary at the stations. Hydraulic simulators in both steady and transient states will help make an accurate design and will guarantee that the project will have good performance along the operating years without any unexpected situation. A preselection of the compressor units can also be performed while at the design phase (Santos, 1997). Another important job that can be checked in transient analysis is the pipeline operating points inside the performance maps of the compressors on an early basis so as to allow proper selection of impellers and number of compressor units as the capacity increases yearly. Different compressor manufacturers can be modeled to check performance and fuel usage. Operations close to surge line or operations that would require recycling would also be identified during transient analysis underlying the importance of doing this kind of simulation on the design phase of a pipeline.

When the compressor design point is decided, the power required from the driver can be calculated using Equation (8-28).

11.7 PIPELINE OPERATIONS

In the industry supply chain, pipeline operations is an integral part of the transportation between exploration and production (E&P) or the “upstream end” precedes it and distribution or the “downstream end” follows it. Pipeline operations evolved from being prescriptive (i.e., defined by mandatory requirements) to its current stance of being performance-based driven by risk management principles.

These trends stemmed from competitive forces that decrease operating costs; they also have been evolved because of the experience gained from several decades of pipeline operations along with the technologies and applications that developed along the way. These evolutions have given pipeline operators the tools they need to survive under these conditions. The pipeline facilities are mature to the point that many of them have exceeded their originally intended design life of approximately 25 years at the time of conception. Today, most of these facilities continue to operate, partly for economic reasons as they are too costly to replace and also partly because these facilities still remain worthy of continued use (i.e., they are still deemed to be safe). Recognizing this, operating companies continue to extract value from these facilities, but under tremendous scrutiny and heightened awareness of their existence and vulnerabilities.

Current pipeline operation activities have taken on a new dimension of performance. While the basic activities continue, such as mechanical operations and maintenance of the facilities, including line pipe, valves, and valve actuators, corrosion prevention and control, and pipeline monitoring as well as the focus on safety, the optimization of resources is being considered while still achieving safety, reliability, and efficiency. These challenges become more daunting given the fact that these pipeline systems have expanded and merged, often acquiring systems built by others under different design, construction, and operating philosophies. Further, staff reorganization and attrition saw much of the corporate knowledge and information misfiled or discarded. Some companies remained as free-standing, whereas many became a part of a larger corporate entity. To overcome these developments, pipeline operators now strive for standardization in their procedures and compare their performance to industry benchmarks to gauge their performance and identify areas for improvement.

Certain time-dependent defects such as corrosion and environmental concerns started to manifest themselves in unplanned incidents. Development of other infrastructure at or near pipeline right-of-way saw an increase in third-party incidents and close calls. Pipeline regulators too evolved over these times and increased their vigil over the industry but allowed them to formulate their own facility management programs. Industry sponsored research programs to better understand the consequence effects of pipeline incidents for risk evaluation.

Consequently, pipeline operation has been transformed toward the following areas of focus.

a. Making effective choices among risk-reduction measures.
b. Supporting specific operating and maintenance practices for pipeline subject to integrity threats.
c. Assigning priorities among inspection, monitoring, detection, and maintenance activities.
d. Supporting decisions associated with modifications to pipelines, such as rehabilitation or changes in service.

These focus points require that pipeline operations activities include the following elements (Mokhatab and Santos, 2005).

1. Baseline assessment and hazard identification
2. Integrity assessment by

    a. In-line inspection
    b. Hydrostatic testing
    c. Direct assessment
    d. Defect management and fitness for service
    e. Information management and data integration
    f. Risk management

3. Integrity management programs
4. Operator qualification and training
5. Operating procedures, including handling abnormal operating conditions
6. Change management
7. Operating excellence

These elements constitute a broad makeup of pipeline operations. Not only must operators be aware of them but they must also be well versed in their application, improve them continuously, and incorporate them into a comprehensive and systematic integrity management plan. Combined, these elements form the basis for directing a prevention, detection, and mitigation strategy for their system.

Pipeline operations will be the longest phase of the life cycle of a pipeline when cost management becomes a high priority. This priority will see pipeline operators performing many scenarios of life extension of the existing assets for enhancing value to its stakeholders.

REFERENCES

API RP 14E, “Design and Installation of Offshore Production Platform Piping Systems,” 4th Ed. Production Department, API, Dallas, TX (April 1984).

Asante, B., “Two-Phase Flow: Accounting for the Presence of Liquids in Gas Pipeline Simulation.” Paper presented at 34th PSIG Annual Meeting, Portland, Oregon (October 23–25, 2002).

Aziz, K., and Ouyang, L.B., Simplified equation predicts gas flow rate, pressure drop. *Oil Gas J.* **93**(19), 70–71 (1995).

Beggs, H.D., “Gas Production Operations.” OGCI Publications, Oil and Gas Consultants International, Inc., Tulsa, OK (1984).

Brill, J.P., and Beggs, H.D., “Two-Phase Flow in Pipes,” 6th Ed. Tulsa University Press, Tulsa, OK (1991).

Buthod, A.P., Castillo, G., and Thompson, R.E., How to use computers to calculated heat pressure in buried pipelines. *Oil and Gas J.* **69**, 57–59 (1971).

Campbell, J.M., Hubbard, R.A., and Maddox R.N., “Gas Conditioning and Processing,” 3rd Ed. Campbell Petroleum Series, Norman, OK (1992).

Carroll, J.J., “Natural Gas Hydrates: A Guide for Engineers.” Gulf Professional Publishing, Amsterdam, The Netherlands (2003).

Cleveland, T., and Mokhatab, S., Practical design of compressor stations in natural gas transmission lines. *Hydrocarb. Eng.* **10**(12), 41–46 (2005).

Colebrook, C.F., “Turbulent Flow in Pipes with Particular Reference to the Transition Region between the Smooth and Rough Pipe Laws.” J. Inst. Civil Engineers, 11, 133–156, London (1939).

Constantinides, A., and Mostoufi, N., “Numerical Methods for Chemical Engineers with MATLAB Applications.” Prentice Hall, New Jersey (1999).

Edalat, M., and Mansoori, G.A., Buried gas transmission pipelines: Temperature profile prediction through the corresponding states principle. *Energy Sources* **10**, 247–252 (1988).

Garland, W.J., Butler, M., and Saunders, F., “Single-Phase Friction Factors for MNR Thermalhydraulic Modeling.” Technical Report, McMaster University, Ontario, Canada (Feb. 23, 1999).

Goulter, D., and Bardon, M., Revised equation improves flowing gas temperature prediction. *Oil Gas J.* 77, 107–108 (1979).

Hughes, T., “Optimum Pressure Drop Project.” Facilities Planning Department Internal Reports, Nova Gas Transmission Limited, Calgary, Alberta, Canada (1993).

Huntington, R.L., “Natural Gas and Natural Gasoline.” McGraw-Hill, New York (1950).

Ikoku, C.U., “Natural Gas Production Engineering.” Wiley, New York (1984).

Jain, A.K. “An Accurate Explicit Equation for Friction Factor.” J. Hydraulics Div. ASCE 102, HY5 (May 1976).

Kennedy, J.L., “Oil and Gas Pipeline Fundamentals,” 2nd Ed. Pennwell, Tulsa, OK (1993).

Kumar, S., “Gas Production Engineering.” Gulf Professional Publishing, Houston, TX (1987).

Lervik, J.K., *et al.*, “Direct Electrical Heating of Pipelines as a Method of Preventing Hydrates and Wax Plugs.” Proceeding of the International Offshore Polar Engineering Conference, 39–45, Montreal, Canada (May 24–29, 1998).

Maddox, R.N., and Erbar, J.H., “Gas Conditioning and Processing: Advanced Techniques and Applications.” Campbell Petroleum Series, Norman, OK (1982).

Mohitpour, M., Golshan, H., and Murray, A., “Pipeline Design and Construction: A Practical Approach.” ASME Press, American Society of Mechanical Engineers, New York (2002).

Mokhatab, S., and Santos, S., Fundamental principles of the pipeline integrity: A critical review. *J. Pipeline Integr.* **4**(4), 227–232 (2005).

Moody, L.F., Friction factors for pipe flow. *Trans. ASME* **66**, 671–684 (1944).

Neher, J.H., The temperature rise of buried cables and pipes. *Trans. AIEE* **68**(1), 9 (1949).

Norsok Standard, “Common Requirements: Process Design.” P-CR-001, Rev. 2, Norwegian Petroleum Industry, Norway (Sept., 1996).

Nikuradse, J., “Stromungsgesetze in Rauhen Rohren.” Forschungsheft, Vol. B, VDI Verlag, Berlin (July/Aug., 1933).

Ouyang, L.-B., and Aziz, K., Steady state gas flow in pipes. *J. Petr. Sci. Eng.* **14**, 137–158 (1996).

Santos, S.P., "Transient Analysis: A Must in Gas Pipeline Design." Paper presented at 29th PSIG Annual Meeting, Arizona (Oct. 15–17, 1997).

Santos, S.P., "Series or Parallel: Tailor Made Design or a General Rule for a Compressor Station Arrangement?" Paper presented at 32nd PSIG Annual Meeting, Savannah, GA (Oct. 28–30, 2000).

Santos, S.P., and Saliby, E., "Compression Service Contract: When Is It Worth?" Paper presented at 35th PSIG Annual Meeting, Berne, Switzerland (Oct. 15–17, 2003).

Schilichting, H., "Boundary Layer Theory," 7th Ed. McGraw-Hill, New York (1979).

Schroeder, D.W., "A Tutorial on Pipe Flow Equations." Paper presented at 33rd PSIG Annual Meeting, Salt Lake City, UT (Oct. 17–19, 2001).

Serghides, C.K., Estimate friction factors accurately. *Chem. Eng.* 63–64 (1984).

Smith, R.V., "Practical Natural Gas Engineering," 2nd Ed. Pennwell, Tulsa, OK (1990).

Streeter, V.L., and Wylie, E.B., Natural gas pipeline transient. *SPE J.* **10**, 357–364 (1970).

Streeter, V.L., and Wylie, E.B., "Fluid Mechanics." McGraw-Hill, New York, (1979).

Towler, B.F., and Pope, T.L., New equation for friction factor approximation developed. *Oil Gas J.* **92**(14), 55–58 (1994).

Towler, B.F., and Mokhatab, S., New method developed for siting line heaters on gas pipelines. *Oil Gas J.* **102**(11), 56–59 (2004).

Uhl, A.E., NB-13 Committee, "Steady Flow in Gas Pipelines." Institute of Gas Technology, Report No. 10, American Gas Association, New York (1965).

Weymouth, T.R., "Problems in Natural Gas Engineering." Trans. ASME, Reference No. 1349, 34 (1912).

Zagarola, M.V., "Mean Flow Scaling of Turbulent Pipe Flow." Ph.D. thesis, Princeton University, Princeton (1996).

CHAPTER 12

GAS PROCESSING PLANT CONTROLS AND AUTOMATION

12.1 INTRODUCTION

Automation has become an increasingly important aspect of gas processing. The amount of specific types of automation is a major decision now in the construction of new plants. With many existing plants having debottlenecked and improved process efficiency through mechanical means, automation is now a focus for further improvements. Automation provides the means for fully utilizing the mechanical capabilities of the equipment at all times and to run the process at its most efficient points in a stable and reliable fashion. A good automation platform can be leveraged to provide the right information at the right time to the right personnel to make the right decisions in a timely manner. Historical data can be collected in virtually any time frame and analyzed statistically. With this historical data, upset situations can be reconstructed and production reports can be automatically generated just to name a couple of the numerous uses for these data. Many processors have upgraded their plants to higher level control systems such as distributed control systems (DCS). Some of these processors ask, “Where are the benefits?” The correct questions to ask are, “‘How do I best leverage by automation equipment to maximize benefits?” and “What are the best methods for measuring these benefits?”

This chapter discusses the elements of automating today's gas processing plants, including considerations for instrumentation, controls, data collection, operator information, optimization, and management information. The advantages and disadvantages of various approaches are analyzed in this chapter. Also, strategies for identifying and quantifying the benefits of automation are discussed.

12.2 EARLY METHODS OF GAS PLANT AUTOMATION

The earliest gas processing plants were typically controlled manually by opening, pinching, and closing valves to meet their operating requirements. Pneumatic control systems were adopted quickly. These pneumatic control systems allowed the use of proportional, integral, and derivative (PID) controllers to send analog outputs to control valves to change their opening. As long as a sensor is available as a process variable for the controller, then a set point given by the operator can be targeted automatically. Discrete control could also be accomplished with pneumatic control systems by employing devices to fail with or without an air signal. In most cases, compressed air is used as the pneumatic conveyor; however, natural gas is used in some remote operations and hydraulic oils are also employed. A reliable source of clean, oil-free air is quite important in the operation of pneumatic air systems.

As electronic controls were developed, these systems became the standard, although some elements of the pneumatic control systems are still in use. Most control valves today are still pneumatically operated for new and existing plants even though electronic control valves have been on the market for decades. Electronic controllers were accepted due to their lower price and greater reliability. These controllers included fewer moving parts to maintain. For new installations, wire was now run instead of pneumatic tubing with electronic to pneumatic (I/P for current to pressure) transducers added at the control valves. Process sensors such as temperature, pressure, level, and flow indicators were converted to electronic types as available and justified. Electronics also brought widespread use of safety systems such as vibration sensors, burner management, and emergency shutdown systems. Other special controllers, such as dedicated surge controllers for rotating equipment and triple modular redundant shutdown devices, were developed.

12.3 MICROPROCESSOR-BASED AUTOMATION

12.3.1 Programmable Logic Controllers (PLC)

Next came development of programmable logic controllers. This type of controller was first intended for discrete factory floor applications such as those found in the auto manufacturing industry. These controllers employ programming of a microprocessor to mimic electronic relays. As PLCs gained wider acceptance, the functionality expanded to include PID control and other nondiscrete capabilities. PLCs are still quite effective for batch-type operations in the gas processing industry, such as solid bed gas dehydration and start-up, as well as shutdown sequencing for rotating equipment. This automation platform is often used for smaller new facilities and smaller scope retrofits of existing facilities. A human machine interface (HMI) is highly recommended to monitor the activities of the PLC.

12.3.2 Distributed Control Systems (DCS)

The definition of a distributed control system varies somewhat, but a rather simple definition is a control system method that is spread, or distributed, among several different unit processes on a common computer platform. These systems are typically hardwired and exist within finite boundaries, although wireless means are becoming accepted, as data security issues are resolved. DCS offers the advantage of centralized control, while retaining the capability of local control. True distributed control systems use localized control, which is in turn controlled by the operator located at a central location. A DCS consists of the following:

- Remote control panel or device
- Communications medium
- Central control panel or facility
- Control, interface, and database software

A DCS may be as simple as one PLC remotely connected to a computer located in a field office. Larger systems may be PLC based, but will most likely consist of specially designed cabinets containing all of the equipment necessary to provide I/O and communication. One point to consider during the design and specification of a control system is the level of autonomy each node will have in the event of a network or system failure. A true distributed system will allow most remote nodes to operate independently of the central control facility should the facility go off line or lose communication capability. Each remote node should be able to store the minimum process data required to operate in the event of such a failure. In this manner, costly and potentially disastrous process upsets can be avoided. If the system is performing both monitoring and control of a process or facility, it is referred to as a supervisory control and data acquisition (SCADA) system. Most systems will transfer data and commands using communication protocols such as Ethernet or some other open standard, depending on the DCS vendor (Capano, Dan, *Distributed Control Systems Primer*, DTS, Inc.).

12.3.2.1 Remote Control Panel

The remote panel for a DCS or SCADA is typically referred to as a remote transmission unit (RTU). A typical RTU contains terminal blocks, input/output modules (both analog and digital), a computer or proprietary processor, and a communications interface. A RTU, depending on where it is located in the world, can perform both monitoring and control of a given process or processes.

12.3.2.2 Communications Medium

The communications medium is a cable or wireless link that serves to connect the RTU to the central control facility. There are several methods of doing this; typically a cable, either a coaxial or twisted pair, is connected between the central control computer and the remote unit, or RTU. It is considered prudent to run two cables, on different routes, between the two in order to increase the reliability of the system. A network operates by taking data from the sending station, or node, and packaging and routing the information to the proper receiving station. The possibility of electrical noise, physical abuse, and software bugs should be considered when implementing a DCS.

12.3.2.3 Central Control

The control room is the center of activity and provides the means for effectively monitoring and controlling the process or facility. The control room contains the HMI, a computer that runs specialized software designed for that purpose. There may be multiple consoles, with varying degrees of access to data. In most cases, each operator or manager is given specific rights to allow more or less access and control of the system. The plant superintendent, for instance, may have complete control over his facility, while a technician may only have access to specific data on a particular process. This is done to avoid accidents and process upsets. This scheme also affords a degree of security, ensuring that only properly trained and authorized personnel can operate the various parts of the facility. The HMI presents the operator with a graphical version of the remote process. Depending on the skill of the operator and the level of sophistication of the interface, the process may be represented by anything from simple static graphics and displays to animation and voice alerts. Most packages afford the operator wide latitude on the design of the interface. The common thread to each system is the input–output (I/O) database.

The database contains all of the I/O defined for that DCS. This does not mean that all process data will be monitored and controlled; it means that only data defined by the designers to be monitored and controlled will be available to the DCS. This database is a product of detailed evaluation of the process by the designer who typically has the responsibility, with operator input, to design the most effective control schemes for a particular facility. The database is the reference the control software uses to correctly address each remote I/O point. Each database entry corresponds to an entity on the system, whether it is a physical point or an internal, or "soft" point such as an alarm, timer, or screen entity. A disadvantage of the early generations of PLC and DCS was the proprietary communications protocols employed. Some protocols such as MODBUS were adopted as a defacto standard, but had limitations as communications complexity increased.

12.3.3 Standards and Protocols

More recently, control systems have standardized on protocols made popular through the wide use of personal computers. Ethernet and Object Linking and Embedding (OLE) for Process Control (OPC) are two examples of highly accepted protocols. Ethernet is used mainly for device-to-device communications, whereas OPC is used primarily for application-to-application communications. Various versions of field buses have been developed to further standardize device-to-device communications. Early DCS was also based on legacy computing platforms such as VMS and the many different UNIX versions. Although Solaris UNIX is still used widely, Microsoft's Windows platform is gaining popularity (Poe and Harris, 2005).

12.4 CONTROL OF EQUIPMENT AND PROCESS SYSTEMS

12.4.1 Gas Gathering

Gas-gathering systems are typically controlled through pressure control. Because most gathering systems employ primarily reciprocating compressors (employing a piston or screw), the discussion will focus on control of this type of compression. Rod loadings, maximum discharge temperature, and liquid entry as well as minimum and maximum speeds must be considered when operating these machines. It is quite important to prevent liquids from entering the compressor. An adequately sized upstream scrubber with a mesh pad should be installed. A level controller to automatically dump liquids is required. The basic purpose of the gas-gathering compressors is to keep the wellhead pressures down and the pressure to the downstream facilities, whether gas conditioning systems, liquids recovery facilities, or transmission pipelines, up to a minimum pressure. The speed of the reciprocating machine is the main manipulation to control the suction pressure while the discharge pressure is dictated by flow rate and downstream resistance. Speed range is rather limited for these compressors, but can be manipulated from about 90 to 100% of maximum speed. Other forms of capacity control include pockets and valve unloaders. These may be manual or automated. Rod loading, which is a function of the pressure differential across the compressor (discharge pressure less suction pressure), must not be exceeded or damage will result. Because there is no direct way to limit discharge pressure, a recycle line is installed to allow gas to recycle to the suction and raise the suction pressure. This line is often controlled by a minimum flow controller, but is more effective from an efficiency perspective when controlled off a differential pressure or even a rod load calculation where the control platform can be configured accordingly. The recycle line should be routed after the discharge cooler and discharge scrubber to increase its effectiveness. The temperature rise across the compressor is dictated by the ratio of discharge pressure to suction pressure. High temperatures can warp piping and destroy packings. Discharge temperature should be monitored and the speed reduced or recycle rate increased to keep the temperature below the maximum allowable. The primary medium for provision of cooling the gas discharged from gas gathering systems is air. The following typically controls these coolers:

- louvers
- on–off fans
- multispeed fans
- variable pitch fans
- variable speed fans, or gas recirculation

In most cases, lower temperatures of gas exiting the cooler are preferred for downstream processes. However, temperatures below the hydrate or freezing point are not desired. In these cases, a temperature controller manipulating any of the aforementioned coolers can maintain the desired temperature. Logic may be employed to determine when fans are turned on, off, or selected to run at a discrete speed.

12.4.2 Gas Treating

The primary method of gas treating is with chemical absorbents. This process is analogous to gas dehydration with absorbents. Some differences are noted here.

1. The objective is acid gas removal and, therefore, the contactor outlet gas should be analyzed for H_2S and/or CO_2.
2. Reflux temperature is set lower to maximize the retention of water.
3. In some cases, selective treating of H_2S versus CO_2 is desired, and the contactor outlet analyzer can drive a stripper overhead temperature setpoint, which in turn drives the ratio of heat medium used for the reboiler to absorbent flow.

Physical absorption is sometimes employed that utilizes a series of flash tanks, which yield better absorbent regeneration at lower temperatures. A vacuum is often pulled on the final stage of flash. In this case, the vacuum driver may need to be on speed control or eductor on flow control to prevent implosion of the vessel. Hydraulic turbines that use the energy of depressuring to drive recirculation pumps are quite effective. A “helper” pump is always required to make up the horsepower deficiency.

12.4.3 Sulfur Recovery

The most common sulfur recovery process, is the Claus process, where one-third of the H_2S must be converted to SO_2 for proper stoichiometry. The theoretical sulfur recovery efficiency drops sharply when the stoichiometry is off the 2 moles of H_2S to 1 mole of SO_2. A proper amount of oxygen, typically in the form of air, must be introduced to the reaction furnace. A tail gas analyzer is installed after the final reaction stage and before incineration. Two parallel valves, a main valve and a trim valve, are usually available on the discharge of the combustion air blower. Ideally, the main combustion air is manipulated on a feed-forward basis as acid gas feed rate and H_2S percentage varies. Feedback from the tail gas analyzer will control the trim valve. Because the time delay between air introduction and tail gas analysis may be significant, a model predictive scheme may improve the ability to maintain proper stoichiometry.

Reaction furnace temperature must be maintained between a minimum and a maximum temperature. Infrared measurements are quite effective in this service. Air preheat can be increased or decreased to raise or lower the reaction furnace temperature. Otherwise, air input requires adjustment. Conversion in the reactors is a trade-off between equilibrium favored at lower temperatures and kinetics favored at higher temperatures. Equilibrium usually dictates when the catalyst is fresh, whereas kinetics typically dictates when the catalyst is near the end of life. Another factor when the temperatures are low is that the converter outlet temperature must be maintained above the sulfur dew point. Converter outlet temperature is controlled primarily by the reheat, which directly controls the converter inlet temperature. Reheat can be classified as direct or indirect. Direct reheat is when hot gas bypasses the waste heat boiler at the outlet of the reaction furnace and is injected at the inlet of each converter. Lowering the converted inlet temperature when kinetics allow causes less gas to bypass and gives more shots at additional stages of conversion. Indirect reheat is with a heat medium such as steam or hot oil. Control of the reheat is straightforward with the indirect methods. Condensers are air cooled or water cooled and operated at their lowest temperature to achieve minimum dew point.

12.4.4 Gas Dehydration

This section covers two types of gas dehydration typically employed in gas processing operations. The first is absorption, typically with a glycol, and the second is fixed bed absorption, typically with mol sieve.

12.4.4.1 Absorption

Dehydration by absorption has several aspects, which require control. These include the following:

- Lean absorbent flow rate and temperature
- Contactor pressure
- Flash tank pressure (where applicable)
- Stripper pressure
- Stripper reboiling
- Stripper reflux

Because the objective of gas dehydration is removal of water from the gas stream, the outlet gas stream should be monitored continuously with a moisture indicator. This indication should be monitored to adjust the flow of lean absorbent and the heat input to the stripper reboiler. The flow of absorbent should be based on a ratio of the gas flow that is corrected based on the moisture indicator reading. This adjustment should take precedence when the moisture is lower than required, as sensible heat will be saved in the stripper in addition to latent heat. The flow of heating medium to the stripper should be on ratio control and corrected based on the moisture indicator reading. This adjustment should take precedence when the moisture is higher than required, as the ability to dry is to driven largely by the water content of the lean absorbent. If the lean absorbent is increased without increasing the ratio of heat to flow, then the moisture content of the lean absorbent may not be reduced. Bottom temperature is not a good control basis, as it will only indicate the boiling temperature of water at the pressure encountered at the bottom of the stripper. A better indication is a pressure-compensated top tray (or above the packing height) temperature. This has direct correlation with the water content of the lean absorbent. A low absorbent temperature improves its ability to hold water. However, too low a temperature may lead to condensation of hydrocarbons into the absorbent, causing foaming. An ideal strategy is to control the temperature of the absorbent about 5–10°F above the temperature of the inlet gas.

Feed-forward strategies that take into account the water content, flow rate, and temperature of the inlet gas to the absorber can also be employed. Model predictive strategies can account for the relative effects of lean absorbent temperature, flow, and stripper heat. A high pressure on the contactor is desirable to increase the contactor capacity and enhance the absorption of water. A back pressure controller should be employed to maintain a high pressure without causing the relief valve to function. The pressure should be maintained as low as possible when a flash tank exists between the contactor and the stripper. A minimum pressure is required to "push" liquids into the contactor while decreasing the demand for stripper reboiler duty. One strategy will raise or lower the flash tank pressure when the liquid level exceeds a desired dead band around the set point. If the liquid rises above the dead band, the pressure is increased. Once the liquid falls within the dead band, the pressure can be decreased slowly. This strategy works best with nonlinear level control, which is always recommended for flash or feed tanks.

The stripper pressure should also be maintained as low as possible to lower the boiling point of the water stripped and allow the flash tank, if installed, to run at a lower pressure. Differential pressure measurements should be installed to indicate the onset of column flooding. When the differential pressure approaches set point the stripper pressure should be raised to alleviate flooding. The reflux condenser should be set at a temperature to attain maximum recovery of entrained glycol without condensing excessive water unless required by environmental considerations.

12.4.4.2 Adsorbents

Because drying with fixed bed adsorbents is a multibed process, the main control is cycle and bed switching. The dryer modes are drying, regenerating, cooling, and standby. Typically, each cycle is set for a fixed time and the beds are cycled through the use of switching valves per this timer.

12.4.5 Liquids Recovery

12.4.5.1 Condensate Stabilization

Condensate is stabilized by stripping light hydrocarbon components in a fractionation tower. Nonlinear level control is recommended for the upstream flash or feed tank to provide a steady feed rate. Due to the nature of the condensate, online analysis is very difficult. Typically a bottom temperature, preferably pressure compensated, is used to control the input of heat to the reboiler. A laboratory analysis is required to verify the adequacy of the bottom temperature set point. An inferential property predictor can be added to drive the temperature set point in between laboratory updates. The reflux temperature, when employed, should be controlled by a sensitive tray above the feed tray. Tower pressure should be driven as low as possible to enhance separation subject to constraints on an overhead compressor.

12.4.5.2 Refrigeration

Refrigeration is used to achieve the bulk condensation of natural gas liquids. Propane refrigerant is the primary medium used in gas processing. The main control aspects are compression, compression driver, refrigerant condenser, economizers, and chillers. Both centrifugal and reciprocating compressors are commonly employed in this service with turbine, electric motor, or gas engine drivers.

Lower temperatures are achieved at lower compressor suction pressures subject to surge conditions on centrifugal compressors and rod loading of reciprocating compressors. The suction pressure directly affects the pressure on the chillers.

Level control of the chillers is cascaded to the flow of refrigerant to the chiller. This control is critical to assure that the chiller tubes are covered without carryover of liquids. A scrubber or economizer before each compression stage is necessary to dump liquids whenever encountered. Economizers should be employed on multistage systems. The economizer pressure should be set to accommodate the compressor load and minimize kickback of high-stage vapor to lower stages. Refer to the sections in this chapter for compressor and driver control considerations.

12.4.5.3 Cryogenic Recovery (Turboexpander Processes)

Expansion with turboexpanders is now the main process employed for recovering natural gas liquids. Turboexpanders can be controlled for various objectives: inlet pressure, demethanizer pressure, or residue pressure are the most common. Guide vanes are manipulated to control the speed of the expander. A Joules–Thompson (JT) valve is always included to allow rapid unloading of the expander. One split range controller typically operates the guide vanes and JT valve so that the JT valve will open when the manipulation of the guide vanes has been exhausted. The compressor driven by the turboexpander in either inlet compression or residue compression mode requires a recycle valve to maintain a minimum flow for surge protection. Depending on the exact cryogenic recovery processing scheme, additional controls may be required for heat exchanger circuit flow splits, chillers, separator levels, and pressure profiles. Heat exchanger flow splits are typically configured as flow ratios. This ratio may be overridden to prevent “cold spins” or prevent temperatures below the critical temperature in the cold separator upstream of the turboexpander.

12.4.5.4 Demethanizer

The demethanizer is integral to the turboexpansion process. The various feeds to the column are created at multiple points in the process. Side reboil heat sources and sometimes the bottom reboiler heat are integral to the heat exchanger circuit. Seldom are the side reboiler temperatures controlled. Manipulating the heat to the bottom reboiler controls a bottom temperature, preferably pressure compensated. An online chromatograph monitors the methane and/or carbon dioxide content.

Ideally, this output would reset the bottom temperature. Demethanizers are good candidates for model predictive control due to the disturbances caused by the side reboilers, inlet flow rates, and inlet compositions. Minimizing the pressure of the demethanizer based on turboexpander constraints and residue compression constraints is a major opportunity for increasing liquids recoveries.

12.4.6 NGL Fractionation

NGL fractionation consists of deethanization, depropanization, debutanization, and butane splitting (or deisobutanization). The control schemes for each are analogous. The major control points for these fractionators are reboiling heat, reflux, and pressure. Again, nonlinear level control is recommended for feed tanks and bottom surge levels. Reboiling heat is manipulated to control the bottom composition. The composition is cascaded to a sensitive temperature below the feed tray. Preferably the temperature is pressure compensated.

Reflux flow is manipulated to control the bottom composition. The composition is cascaded to a sensitive temperature above the feed tray. Preferably the temperature is pressure compensated. Minimum reflux schemes are employed to assure that reboiling load is not increased due to excessive reflux and, conversely, excessive reboiling leading to greater reflux rates for a given separation. Internal reflux calculations and multivariable control schemes can achieve minimum energy consumption for a given separation.

Pressure should be minimized on these towers subject to constraints such as flooding, condenser temperature, and bottom hydraulics. Flooding is indicated by delta pressure measurements across the tower. Reflux is more difficult at lower temperatures, as the available duty of the condenser may be limited. There must be sufficient head on the bottom of the tower to allow liquids to feed downstream towers or satisfy minimum head requirements for pumps. Again, multivariable control schemes handle the pressure minimization issue elegantly.

12.4.7 Centrifugal Compressors

Centrifugal compressors (utilizing an impeller to increase the kinetic energy of a vapor) are gaining wider acceptance in a variety of gas processing services, including feed, residue, and refrigeration compression. These compressors are typically driven by gas turbines or electric motors, but can sometimes be steam turbines. There are several control considerations for this unit operation. Upstream liquids removal, surge prevention, suction pressure, discharge pressure, and driver speed control are the primary issues. Upstream liquid removal is accomplished with vessels upstream of the compressor that remove any entrained liquids and automatically dump liquids based on a liquid level. A simple on/off level control scheme is adequate for this purpose.

Surge is caused by excessive head requirements for a given suction pressure. The horsepower delivered by the compressor driver can be reduced to prevent surge or gas can be recirculated from a higher stage of compression. Slowing the compressor driver is usually the most energy-efficient means. However, some drivers are limited in their speed range. There are many schemes that use antisurge or kick-back valves to quickly increase the suction pressure by recirculating gas. The most sophisticated will take gas density and head curve characteristics into account for a wide variety of operating conditions. The simplest schemes assure a minimum flow or minimum suction pressure for the compressor. These simpler schemes yield horsepower inefficiencies due to a more conservative approach. Kick-back schemes that use cooler gases and minimize the number of stages that gas is recycled are also the most energy efficient.

Suction and discharge pressures can be controlled by adjusting the driver speed, recycling gas, or with throttling valves. Driver speed adjustments are the most energy efficient. Gas turbines will typically have a wide speed range. Steam turbines have a moderate speed range. Electric motors may be constant speed; however, variable speed and variable frequency drives are becoming more popular.

12.4.8 Centrifugal Pumps

Centrifugal pumps are analogous to centrifugal compressors, but are seldom driven by a gas turbine. The same control considerations exist except that the requirement to remove liquids upstream is replaced by a need to remove entrained vapors upstream. The net positive suction head dictates the surge point. Therefore, the level and density of the liquid at the suction of the pump are important.

12.4.9 Reciprocating Pumps

Reciprocating pumps are typically driven by electric motors, but some applications employ gas engines or steam turbines. These pumps are rather forgiving in their operation and require minimal control. The most common control required is capacity control, as these are positive displacement machines. Variable speed and liquid recycle based on minimum flow or upstream level considerations are the main forms of capacity control.

12.4.10 Utilities

The most common utilities found in gas processing plants for process purposes are refrigeration systems, heating (hot oil or steam) systems, and cooling water systems. Refrigeration systems have been covered previously. Hot oil systems employ heaters, mixing tanks, and headers. It is not uncommon for a plant to employ at least two levels of hot oil temperatures. The various temperatures are distributed through separate header systems. Some processes, such as amines and glycols, will degrade when high skin temperatures are encountered. Lower temperature heat medium minimizes reboiler skin temperatures. A common hot oil temperature scheme is to accumulate all heat medium returning from the process in a surge tank. A portion of the hot oil is routed through the heater to the temperature target for the high temperature header. Enough of the liquid in the surge tank bypasses the heater and is mixed with the right amount of heated oil to achieve the low-temperature header target. Steam systems also typically employ several levels of temperature (or pressure). Boilers produce sufficient steam to satisfy a high-pressure header. Steam users, heat exchangers and steam turbines, discharge the exhausted steam into a lower pressure header to be reused or into the condensing system for collection and reuse. Steam turbines that exhaust steam into a lower pressure header for reuse at a lower temperature are called topping turbines. Otherwise, the steam turbines are total condensing turbines.

In order to satisfy the balance for the entire steam system for all the temperature (or pressure) levels required, let-down valves are employed to route higher pressure steam into lower pressure headers based on pressure control. The exhausted system is collected in the lowest pressure header, typically at atmospheric pressure, deaerated, replenished with make-up boiler feed water to replace losses, and boiled again. Process temperature control is usually achieved by simply regulating the flow of heat medium to the heat exchanger.

12.5 AUTOMATION APPLICATIONS

With a central control room using electronic means to transmit data came the advent of applications to reside on the automation platform. These applications are focused on collecting information and using it to operate more cost effectively (Poe and Harris, 2005).

12.5.1 Data Historians

The ability to collect and store a large amount of data on a disk is a key advantage of microprocessor-based automation platforms. Even so, several vendors have specialized in developing historians to store and analyze data more efficiently and effectively. Data compression techniques are used to store a maximum amount of information in a minimum space while maintaining resolution of data. These historian packages come with tools to assist in mining data, graphing, tabulating, and analyzing statistically. The historical database can be manipulated to automatically generate reports as well.

12.5.2 Asset and Performance Management

Asset and performance management software has been developed to also tap into the wealth of information that is now available with microprocessor-based automation systems. These solutions include the following.

- Computerized maintenance management
- Work order generation
- Predictive maintenance
- Control loop performance and tuning
- Online equipment health monitoring
- Process performance monitoring

Asset management focuses on maintaining the plant equipment. Inventory management, work order generation, predictive maintenance programs, and turnaround planning can be accomplished with these tools. Many of these packages include hooks into enterprise planning systems. Control loop performance monitoring and tuning packages are available. These applications can determine whether a control loop is experiencing problems with a valve, positioner, or controller tuning, for example. Other applications trend the vibration, temperatures and other key parameters for rotating equipment to determine when a failure is expected to occur.

Process models can be run online to determine how well a plant is performing compared to an expected performance. Heat exchanger fouling, expander, and compressor efficiencies, as well as tower efficiencies, can be calculated and monitored. These packages include data reconciliation features to overcome the problem of how to adjust process models for inconsistent, missing, or bad data.

The opportunities for application of the process performance audit initiative for increased plant profitability in the gas processing and NGL fractionation industries can be defined by analysis of the individual facilities in order to pinpoint the control loops that are the economic drivers of each facility. The following is a list of applications that require minimum control variability and tight adherence to hard spec limits in order to maximize economic performance (Kean, 2000).

- Distillation towers: feed and reflux flow control loops, reflux temperature control loop, reboiler temperature control loop, reboiler level control loop, and pressure control loop.

- Gas compressors: flow and pressure control loops, surge control loops, station recycle control loops, and gas temperature control loops.
- Acid gas treating systems: stripper reboiler temperature control loop, stripper overhead temperature control loop, stripper reflux flow control loop, contactor and flash drum level control loops, and hot oil heater fuel and air flow control loops.
- Steam boiler systems: steam drum level/feedwater flow control loops, steam pressure and fuel flow control loops, feedwater heater train control loops, and combustion air/o_2 control loops,
- Plant utility systems: cooling water flow control loops, fired heater fuel and air flow control loops, and refrigeration chiller level control loops.

Control loop optimization through the employment of a formal process performance audit by skilled process consultants and control engineers can be an effective route to increased plant profitability. This economic improvement initiative could make the difference between a plant being economically viable or one that is considered for temporary shutdown or asset disposition (Kean, 2000).

As discussed earlier, the basic control loop affects plant performance greatly due to the following facts.

- Process optimization requires optimization of the entire process, both hardware and software (Rinehart, 1997).
- The final control loop plays a significant role in process optimization (Rinehart, 1997).
- Control loop optimization reduces process variability and also increases process reliability.
- Optimization of the control loop is an essential step for successful application of advanced control.
- Large economic returns result from proper sizing, selection, and maintenance of the process control equipment.
- Continuous, online monitoring of both loop equipment and loop performance is a key element for achieving lowest cost of production, while minimizing the life cycle cost of the processing facility.

12.5.3 Statistical Process Control

Data from the automation system can be interfaced to statistical process control packages. This software is used to generate run charts, process capability analyses, process characterization, experimental design, and cause-and-effect diagrams. This type of information is quite valuable to determine the causes of plant instability and off-specification products. It is also an excellent tool when baselining the plant performance and determining the benefits of improved control. Statistical process control concepts form a foundation for many of the Six Sigma and other quality initiatives when applied to continuous processes.

12.5.4 Advanced Regulatory Control

Advanced regulatory control was made much easier with the advent of microprocessor-based controllers. This control methodology basically turns single input, single output control into multiple input, single output control through the use of cascading controllers, selectors, feed forwards, ratios, etc. Shinskey (1996) discusses the variety of control strategies that can be employed in this way. Although not impossible with single loop pneumatic and electronic controllers, the software configuration approach with microprocessor-based controllers superseded the tubing runs, wiring, and other devices necessary to accomplish these strategies with earlier controller forms.

12.5.5 Multivariable Predictive Control

A more elegant and robust form of control is multivariable predictive control. This form of control has been used in the petroleum refining industry since the 1970s and provides true multiple input, multiple output control. Multivariable predictive process control provides a structured approach to managing process constraints, such as limits on valves and rates of change of temperature and pressures. A model for long-range prediction is used to ensure that the constraints upon these variables are not violated. This enables the maintenance of an operating envelope within which the process is constrained. Recently introduced technology that enhances capability in this area includes constrained quadratic programming. In order to determine the optimal set points and constraint values for the controller, an outer optimization can be performed. This optimization can be described as a linear programming technique that is combined with a steady-state model and a cost function, determining the optimum operating point to be derived from a strategy based on minimum energy usage, maximum throughput, or a balance between these or other objectives.

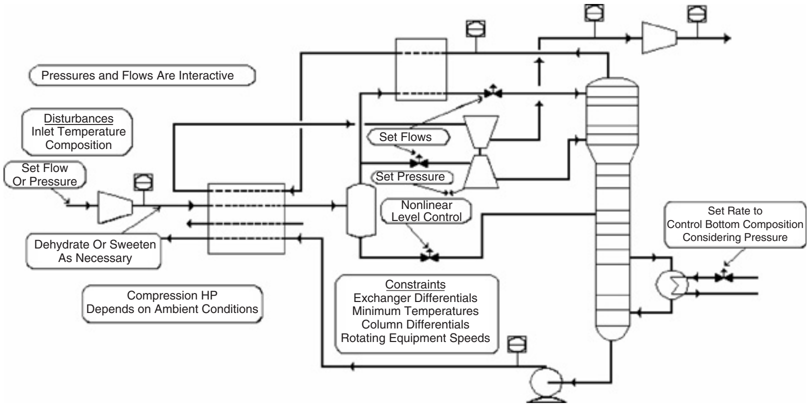


Figure 12-1. Multivariable control on a typical cryogenic demethanizer.

Figure 12-1 shows a multivariable control strategy for a typical cryogenic demethanizer. A multivariable controller takes advantage of the interactive nature of the process. Key controlled variables are modeled dynamically as a function of key manipulated and disturbance variables. Flows depend on the pressure profile, the compression horsepower available, and the efficiency of the turboexpander. When more horsepower is available, such as at night and in cooler weather, then the flow can be increased or the pressure on the demethanizer can be lowered to increase NGL recovery for a given flow demand. Control of the NGL quality, typically for methane or carbon dioxide in ethane, becomes more difficult as tower pressure is adjusted. The multivariable controller can determine the correct heat input as tower pressure is adjusted to maintain maximum recovery and product quality. Pressure-compensated temperature is a key element of this strategy.

Other controlled variables can be minimum and maximum flows, pressures, temperatures, levels, speeds, etc. Demethanizer pressure, flow splits, reboiler flow, and plant inlet pressure are some manipulated variables, while disturbances such as residue pipeline pressure, inlet flow, and inlet composition are considered.

In the aforementioned manner, a “team” of key controllers is pushing the plant to its optimum operating point at all times. Controllers are effectively decoupled, thereby making adjustments similar to an automobile

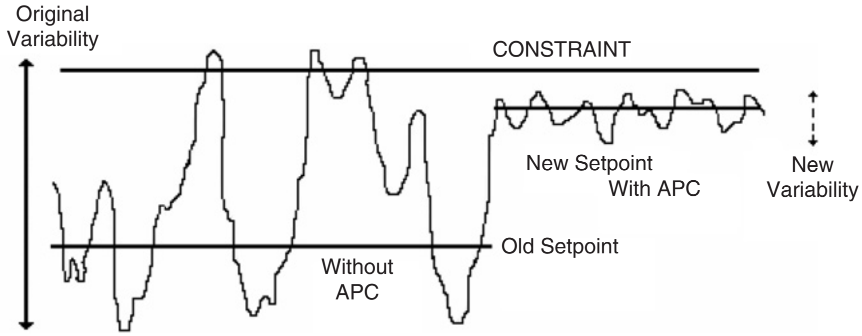


Figure 12-2. Impact of advanced controls. Benefits include operation closer to constraint and reduced variation.

on cruise control. In many regards, the objectives of statistical process control are achieved.

Figure 12-2 shows the beneficial effects of advanced controls. First the variability of the process is reduced by the ability to control closer to set points with the model predictive capability. Once variability is reduced, then the process can be pushed closer to the operating constraints where maximum profitability is attained. This is an area where operators hesitate to operate due to the possibility of exceeding a constraint. With advanced process control, set points can be put closer to these constraints without fear of exceeding the constraints.

12.5.6 Optimization

The next level of automation is optimization. For simpler optimization problems, linear programming (LP) techniques can be employed. Nonlinear techniques may be warranted for more complex optimization opportunities where linear methods would miss significant benefits. With the advances in computing power and optimization mathematics, online, rigorous optimization has become a reality. Again, petroleum refining is leading the way, but several gas processing applications have been developed. Optimization systems must rely on a multivariable predictive control system. As the robustness of the multivariable predictive control system improves, so does the effectiveness of the optimization system. With the business focus that optimization brings, multivariable

Table 12-1
Recent Developments in Multivariable Predictive Control Technology (Canney, 2004)

Development	Technology
Nonlinear controllers	Neural nets and other empirical modeling techniques
Automated process testing	Multistep, random frequency testing
Performance monitoring	Metrics for model adequacy
Remote implementation and support	Web-enabled applications
adaptive	Background testing and model identification
Control system embedded	Powerful processors coupled with programming efficiencies

predictive control developments should address the requirements of operations management. Table 12-1 details recent technology developments that have improved operations management and the benefits of multivariable predictive control applications. These advancements have enhanced basic process operations. This enables staff to be reployed and provide more valuable operations functions that provide increase efficiency, reliability, reduced cost, improved quality, ideal staffing, and responsiveness to changing business requirements.

Networking and Web capabilities continue to impact advanced applications. Some companies have central support centers or have contracted application service providers. Automated testing can be easily monitored remotely. Performance monitoring technologies and tools provide clear benchmarks of expected performance. Updates and improvements can be applied remotely. These capabilities reduce project and support costs and further improve the economics of multivariable predictive control initiatives. Mixed integer programming and enhanced state-space methods are now employed to improve models and decrease the impacts of unmeasured disturbances. New developments in sensor technology supply new valuable, low-cost process information. Methods for property estimation and predicting the behavior of multiphase and complex reaction systems are also improving. Multivariate statistical methods continue to progress. Visualization technology may be a key to making the models transparent to the end user.

12.5.7 Leveraging Automation

In order to get the most benefit, the automation platform should be leveraged to its maximum capability. The platform itself is about 80% of the cost for about 20% of the potential benefits. A multitude of additional benefits are there for a fraction of the platform cost by adding applications. Include adequate instrumentation upfront for grassroot plants and during retrofits for existing plants. It is much more cost effective to add instrumentation during construction rather than adding it later. Use control system consultants rather than relying solely on equipment manufacturers. Many equipment manufacturers are penalized for including more instrumentation than their competitors. For control system upgrades, do not just replace in kind; replace and enhance. A DCS gives limited additional value when used simply as a panel board replacement (Poe and Harris, 2005).

12.5.7.1 Automation Upgrade Master Plans

Many gas processors have developed automation upgrade master plans for existing facilities where the relative benefits of each unit and the value of enhancements are estimated. In this way the area with the most benefits can bear the majority of the initial cost of the platform. Additions to the platform, once the infrastructure is in place, typically cost less per I/O than the initial installation. Master plans are quite effective when the gas processor teams with an automation consultant. The consultant can share their past experience and advise of future developments that should be considered.

12.5.7.2 Determining the Benefits

The benefits of a fully leveraged automation system are numerous and include the following.

- Control room consolidation (reduced manpower)
- Plant reliability (uptime)
- Plant stability (better efficiency due to fewer process upsets)
- Maintenance management (reduced inventories)
- Product quality (fewer off-specification penalties or give away of overprocessed product)

- Continuous constraint pushing (increased throughput and recoveries as well as energy savings)
- Optimization (energy savings and increased recoveries)

When approached appropriately, these benefits are quantified easily. The key is measuring through baselining and monitoring the results through proper metrics.

12.5.7.2.1 Baselining A good baseline is a crucial element of determining benefits. Historical data are necessary to derive a good baseline. Historical data should include the primary measurement as well as any factors that should be used to normalize the result. For instance, ethane recovery should be normalized for inlet gas flow and composition as well as ambient temperature. Data during upset periods and outliers should be discarded. True process variability is measured best with frequent data capture on the order of every minute. High-quality baselining requires an excellent understanding of the operation in addition to statistical analysis. In addition to measuring mean performance of a process, baselining will reveal the amount of variability in the process and the source of the variability. Reducing variability is one of the main benefits of automation. With reduced variability, the process has freedom to shift to a more beneficial operating point.

12.5.7.2.2 Statistical Analysis Statistical analysis is crucial in determining the benefits of automation. A good statistical analysis will give the most accurate assessment of process performance under changing conditions. Mean or average performance is always the final measure between before and after performance. Again, the mean should be evaluated on a normalized basis so that the performance is assessed on a fair or equivalent basis. Total liquids or even total ethane recovered is not a normalized evaluation. Barrels per MMCF of inlet is a better metric, but does not take into account changes in inlet composition. Recovery levels take into account inlet composition, but not process capability under variable inlet rates or ambient conditions that provide more mechanical horsepower availability. Standard deviation gives a measure of the amount of variability for a normal (bell-shaped) response curve. Depending on the process and the sources of variability, anywhere from one-half to two standard deviation improvement is reasonably expected. One way to check this is to compare best performance or process capability (entitlement) to the mean performance. This difference can be compared to the standard deviation to give an idea of the maximum standard deviation improvement possible. Because standard deviation gives an absolute number and is difficult to compare to the mean, coefficient of variation, which is the standard deviation divided by the mean, gives a relative measure of standard deviation. In other words, the coefficient of variation gives a percentage improvement potential. All responses are not normal or bell shaped. Kurtosis is a measure of the skewness of the response curve and should be considered when evaluating standard deviation. A common example of an abnormal response is when a process runs close to its physical limits to the upside or downside and far from its physical limits to the opposite side. Elimination of upset conditions can sometimes shift a response curve close to normality. Multivariable control packages also include the capability to further identify the dynamic performance of a process with tools such as autocorrelation, power spectrum, and Fourier transforms. Determining financial benefits in the face of uncertainty is always a challenge. Even though we may have a good gauge on process improvement potential from the standpoint of increased product, the value of these improvements is dependent on economic conditions such as pricing, feedstock availability, and product demand. Risk management tools, such as Monte Carlo simulations, can be used to establish the range and certainty within these ranges of financial benefits over the variety of conditions anticipated.

12.5.7.2.3 Performance Improvement Initiatives

Many companies are adopting Six Sigma and other statistically founded performance improvement initiatives. Automation can greatly enhance the effectiveness of Six Sigma and other performance improvement initiatives. Conversely, the rigor of these performance improvement methodologies is useful in documenting the benefits of automation. Take Six Sigma as an example of how performance improvement methodologies and automation complement each other. The four phases of Six Sigma include measure, analyze, improvement, and control (Breyfogle, 1999).

Measurement can sometimes be difficult and painful without automation. With automation and data historians, the measurement and analysis task becomes much easier. After data are collected, the process capability assessment, run charts, cause-and-effect matrices, and so on can be developed to assess the performance improvement opportunity and where to focus the effort. The many elements of automation can be the enabling technology for the implementation and control phases of Six Sigma.

Stabilizing the process is always a key element of performance improvement and is enabled by automation, especially advanced controls. The control phase pretty well speaks for itself.

12.6 CONDENSATE STABILIZER CASE STUDY

The following case study reviews an application of some of the advanced control features and the benefits. This case study is for an actual condensate stabilization process. The main reason for advanced process control on this unit was for quality purposes. Condensate product must meet a Reid vapor pressure (RVP) specification as defined by the customer. The previous method of production relied on laboratory sampling to verify the RVP, which was infrequent. Online analyzers are available but can be expensive. Instead of an online analyzer, the important quality measurement was derived through inferential means. The APC solution provided a stable plant and reduced the RVP variation.

BG Tunisia implemented multivariable predictive control on their gas condensate production system at the Hannibal plant in Sfax, Tunisia. The aims of the project were to maximize condensate yields, improve the stability of the condensate stabilization process, and ensure that quality limits for the product are adhered to at all times. BG Tunisia's Hannibal terminal processes gas at 5.4 MMSCM/D with the condensate column operating at a typical rate of 550 l/m controlling RVP to a maximum limit of 12 psia. Condensate can be sold as crude oil and is therefore more valuable than the alternative natural gas liquids products (Hotblack, 2004).

Figure 12-3 shows a detailed schematic for the condensate process. The liquid hydrocarbon condensate is brought into the plant from the slug catcher and is preheated in the stabilizer feed/bottoms exchanger E-0201. The preheated feed is flashed in the stabilizer feed drum D-0201. Liquid from the feed drum is further preheated through the feed/bottoms exchanger E-0250 and fed to the stabilizer T-0201 on local flow control FC-02016. The feed drum level controller LC-02007 maintains the level by manipulating the feed to the drum. The flashed vapor is sent to the high-pressure fuel system. A boot is provided on the feed drum to separate any entrained water, which is sent to the warm flare drum under level control. Condensed hydrocarbon liquid from the warm separator (D-0701) in the nitrogen rejection unit (NRU) is fed to the NRU liquids flash drum D-0202 under level control via the warm gas/liquid exchanger E-0702 in the NRU.

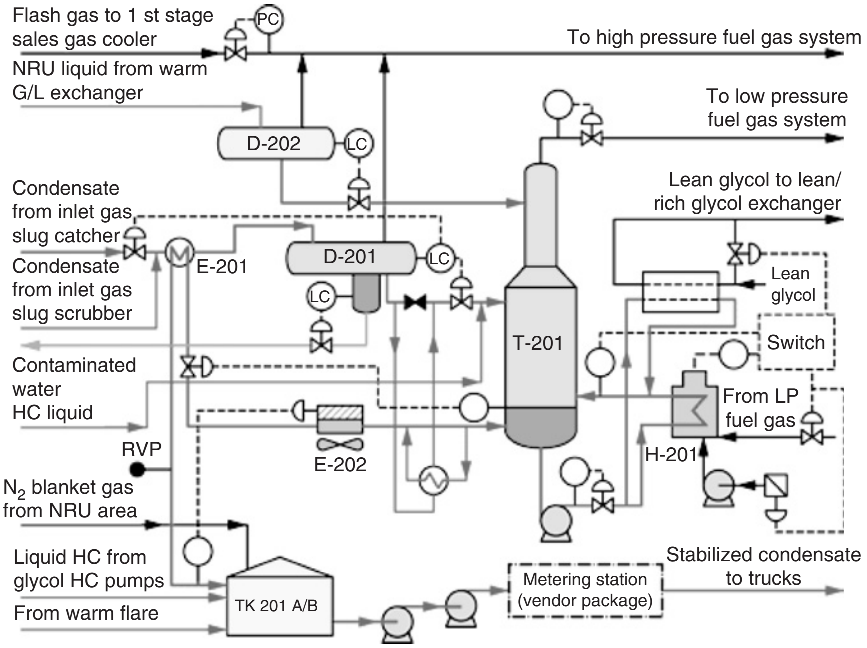


Figure 12-3. Schematic for BG Tunisia's Hannibal terminal condensate process (Hotblack, 2004).

The flashed vapor is sent to the high-pressure fuel system, and liquid from the drum is fed to the stabilizer on level control LC-02026.

The condensate stabilizer T-0201 reduces the vapor pressure of the condensate by removing the lighter components. It is a stripper column with 24 trays. Liquid from the stabilizer feed drum is fed at the midpoint on tray 9, and liquid from the NRU liquids flash drum is fed to the top tray. Overhead vapor from the stabilizer is sent to the low-pressure fuel system through a back-pressure control valve that maintains the tower pressure to set point under the action of PC-02045.

The bottom part of the tower is divided into two sections by a baffle; the baffle does not extend to the very bottom of the tower so there is some mixing between the two sections. Liquid from the bottom tray flows into the section that is preferentially pumped through the fired stabilizer reboiler H-0201. The two-phase stream from the reboiler is returned to the other compartment where the liquid is separated as stabilized product and the vapor flows up the tower to provide stripping action.

The glycol/condensate exchanger E-0203 provides additional heat duty if required.

Stabilized condensate leaves the stabilizer on level control LC-02001, is air cooled in the condensate cooler E-0202 and by exchange with cold inlet feed condensate in the exchanger E-0201, and is sent to the condensate storage tanks TK-0201 A/B.

The main objectives of the condensate stabilizer connoisseur controller are as follow.

- Control RVP to an operator-specified target value
- Enforce any specified unit operating constraints at all times
- Stabilize the unit operation

The controller uses a real-time online estimate (inferential) of RVP. By controlling more tightly to the target value, excessively high RVP and the associated increased flashing of stabilized condensate in the storage tanks and during tanker transfer can be minimized, thus reducing losses. The product specification for RVP is 10–12 psia. Figure 12-4 shows the before-and-after trends of key process variables. The baseline assessment before APC was a mean RVP of 12.25 psia with 1 SD of 0.43. After

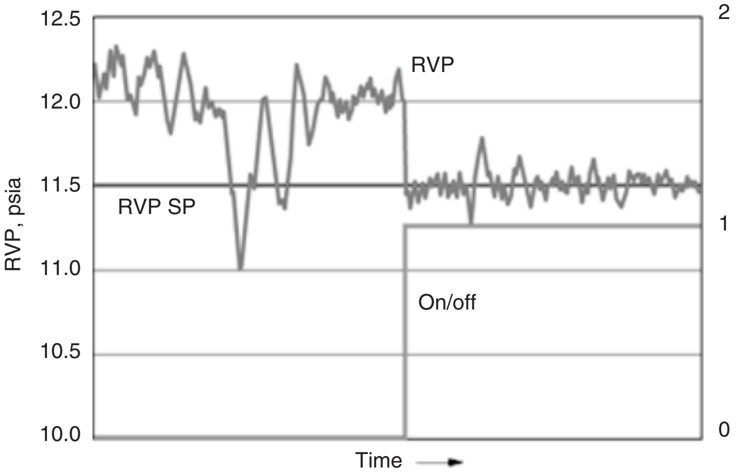


Figure 12-4. Reid vapor pressure (RVP) at BG Tunisia's plant before-and-after trends of key process variables (Hotblack, 2004).

APC was installed, the standard deviation dropped to 0.26, enabling control to 11.5 psia with 95% confidence on the 12-psia limit (Hotblack, 2004).

Constrained optimization drives the process to an optimum without violating process constraints such as the RVP limit imposed by the condensate purchaser.

A hybrid RVP sensor using both first principle and neural network technology provides a continuous measurement for control. Laboratory samples taken on a periodic basis are integrated into the control scheme, improving accuracy and ensuring that product quality is maintained.

The inferential estimate is implemented directly within the controller as a hybrid of a first principles model based on the Antoine vapor pressure equation and a radial basis function neural network. Both the first principles model and the neural network require the stabilization column base temperature (TI-02035) and top vapor pressure (PC-02045). Compensation for process drift and process measurement error is via feedback from the laboratory analysis. A weekly laboratory sample is inserted as a minimum for this purpose.

REFERENCES

Breyfogle, F.W., "Implementing Six Sigma." Wiley, New York (1999).

Canney, W., Advanced process control powers developments in operations management. *Oil Gas J.* **102**(42), 50 (2004).

Capano, Dan, *Distributed Control Systems Primer*, DTS, Inc.

Hotblack, C., BG Tunisia's advanced process control improves condensate product stability. *World Oil Mag.* **225**, 9 (2004).

Kean, J., "Maximize Plant Profitability through Control Loop Optimization." Paper presented at the 79th GPA Annual Convention, Atlanta, GA (March 13–15, 2000).

Poe, W.A., and Harris, S., "Gas Processing Plant Automation A-Z." Paper presented at the 84th GPA Annual Convention, San Antonio, TX (March 13–16, 2005).

Rinehart, N.F., "The Impact of Control Loop Performance on Process Profitability." Paper presented at AspenWorld-97, Boston, MA (Oct. 15, 1997).

Shinskey, F.G., "Process Control Systems: Application, Design and Tuning." 4th Ed. McGraw-Hill, New York (1996).

SUGGESTED READING

Aguilar, J., Poe, W.A., Gamez, J.P., Clay, R.M., and Stegall, R.P., "Compressor Multivariable Control." Paper presented at the ISA Conference, Philadelphia, PA (Oct., 1994).

Alexandre, M.C., Owen, M., Basu, U., and Poe, W.A., "Multivariable Control of Texaco's Eunice South Gas Plant." Paper presented at the 77th GPA Annual Convention, Dallas, TX (March 16–18, 1998).

Al Rougha, S.Y.B, Ni, H., Viswanathan, S., and Gejji, S., "Multivariable Control of ATHEER'S HABSHAN Gas Plant." Paper presented at the 79th GPA Annual Convention, Atlanta, GA (March 13–15, 2000).

Berkowitz, P.N., Papadopoulos, M.N., and Poe, W.A., "Multivariable Control and Optimization for Sulfur Plants." Paper presented at the 47th Annual Laurence Reid Gas Conditioning Conference, Norman, OK (March 2–5, 1997).

Clay, R.M., Chou, K., Everett, G.J., Poe, W.A., and Gamez, J.P., "PC-Based Amine Treating and Sulfur Recovery Unit Optimization." Paper presented at the National Petroleum Refiners Association (NPRA), New Orleans, LA (Nov.1993).

Clay, R.M, Chou, K., Everett, G.J, Poe, W.A., and Gamez, J.P., "Optimization of Amine/Claus Units." Sulphur Magazine, 231 (March–April 1994).

Colwell, L.W., Ma, S., Moran, M.M., Poe, W.A., Wu, P., and Gamez, J.P., "Multivariable Control of Amine Treating and Sulfur Recovery Operations." Paper presented at the 45th Annual Laurence Reid Gas Conditioning Conference, Norman, OK (Feb. 26–March 1, 1995).

Colwell, L.W., Moran, M.K., and Poe, W.A., "How to Justify and Implement Advanced Process Control Projects." Paper presented at the ISA Conference, New Orleans, LA (Oct., 1995).

Colwell, L.W., Papadopoulos, M.N., Poe, W.A., and Gamez, J.P., "What's New in Multivariable Predictive Control." Paper presented at the 74th GPA Annual Convention, San Antonio, TX (March 13–15, 1995).

DeVries, S., Dearwater, G., and Poe, W.A., "Maximize The Profitability of Your Gas Plant Assets –The Integrated Gas Plant of The Future." Paper presented at the 16th International Venezuelan GPA Conference, Caracas, Venezuela (2002).

Ferber, P., Poe, W.A., Basu, U., and Venkatarmanan, G., CNGT installs fuel minimization system to reduce operating cost. *Pipeline Gas Indus.* **82**, 3 (March 1999).

Kennedy, J.M., Saunders, A., and Poe, W.A., "Reducing Human Intervention in Gas Processing." Paper presented at the 81st GPA Annual Convention, Dallas, TX (March 11–13, 2002).

Khalil, A.M., Taha, O., Dunn, M., Al-Yousif, M., and Dhaliwal, S., "Application of Multivariable Predictive Control on a Liquid Recovery Process." Paper presented at the 83rd GPA Annual Convention, New Orleans, LA (March 14–17, 2004).

Lyne, M., Pidlisny, S., Treiber, G, Powley, R., Toffolo, D.S., and Bell, M., "Natural Gas Plant Advanced Control and Optimization." Paper presented at the 75th GPA Annual Convention, Denver, CO (March 1996).

Munsif, H., and Poe, W.A., "A Unique and Improved Approach to Multivariable Control." Paper presented at the AIChE National Meeting, New Orleans, LA (March 8–12, 1998).

Poe, W.A., "Developments in the Application of Multivariable Control and Optimization to Gas Process Operations." VII Encuentro Técnico ISA Occidente, Maracaibo, Venezuela (Sept., 2001).

Poe, W.A., and Berkowitz, P.N., Gas plant advanced process control using personal computers. *Petr. Technol. Q.*, 99–106 (Winter 1997/1998).

Poe, W.A., and Papadopoulos, M.N., "Optimal Allocation of Compressors Using Mixed Integer Algorithm." International Federation of Operations Research Societies, Fourth Specialized Conference: OR and Engineering Design, St. Louis, MO (Oct., 1995).

CHAPTER 13

DYNAMIC SIMULATION OF GAS PROCESSING PLANTS

13.1 INTRODUCTION

Modeling has been used for a very long time for the design and for improved operation of gas processing and transmission facilities. The use of steady-state models is universally accepted in all stages of the design and operation of gas processing plants. Dynamic simulation has been used for a long time, but rigorous first principles dynamic simulation has been confined to use by specialists and control engineers were using models based on transfer functions that were incapable of representing the nonlinearities in systems and the discontinuities in start-up cases, for example. Only since the late 1990s has dynamic simulation become a more generally accepted tool to be used by process engineers and control engineers alike. The software available today enables process engineers with some process control knowledge and control engineers with some process knowledge to build dynamic models fairly easily. The constraint to using dynamic simulation is no longer that dynamic simulation is difficult to employ, but rather that the implementation time for a dynamic model is in the order of two to four times as long as the time needed to implement a steady-state model. Frequently a consultant is employed to develop the model and one or more engineers of an operating company or engineering company would use the model to run the needed studies.

This chapter discusses the areas of application of dynamic process modeling and modeling considerations, both general and for specific equipment used frequently in gas processing plants. The use of dynamic models in specific gas processing units is analyzed. Some case studies are presented to illustrate the use and impact of dynamic simulation on the design and operation of gas processing and transmission plants.

13.2 AREAS OF APPLICATION OF DYNAMIC SIMULATION

The areas of application have been divided into two large groups. The plant design group highlights applications that are used most frequently by engineering companies, whereas the plant operation group is used mostly by operating companies.

13.2.1 Plant Design

Dynamic models have several applications in plant design. Quite often it is difficult to quantify the benefits associated with dynamic simulation. It is important to realize that a dynamic model can be reused for various applications in the design of a plant. The dynamic model will evolve as the design and the project evolve. The most detailed and rigorous model will be used close to plant commissioning and beyond (Brown and Hyde, 2001).

13.2.1.1 Controllability and Operability

Decisions taken very early in the design phase of a new plant or a revamp of a plant can have a significant impact on the controllability or operability of that plant. If the design calculations only use steady-state process simulation, the decision to employ a novel design is often a trade-off between the advantage the novel design brings and the potential controllability and operability issues. A dynamic model will shed more light on the problems that can be expected and lets control engineers devise adapted control strategies to mitigate or remove controllability problems. The use of dynamic simulation will increase the adoption of novel designs and ultimately the efficiency of new or revamped plants.

As these issues need to be analyzed in the early development of a process, the models will necessarily be simpler than the models used when the design has been completed. At this stage concepts are tested and one would not expect quantitative answers from a dynamic model but rather indications of process stability or control feasibility.

13.2.1.2 Safety Analysis

Much goes into ensuring that a plant will be safe for the operators and the people living in the neighborhood of the plant. Huge liabilities are associated with the safety of a plant. Some dynamic models are standard practice for any design, but surprisingly in other areas the use of dynamic models is still limited.

For virtually any plant operating at high pressure depressuring studies will be a standard part of the engineering phase. Depressuring studies are performed to analyze the behavior of pressure and temperature when the pressure of a plant that has been stopped is let down. The depressuring study defines the necessary flaring capacity that is needed and the results may also impact the choice of materials of vessels. Typically, if the pressure letdown of a vessel results in too low temperatures, carbon steel will have to be replaced with stainless steel to prevent the metal becoming too brittle. The other class of depressuring studies is related to the consequences of a fire in the area where a vessel is located. The main goal here is to determine the minimum vent rate that is needed to keep the vessel pressure under control and to bring down the vessel pressure to prevent a failure of the vessel wall as temperature rises.

All plants have emergency shutdown systems (ESD). The design of a system to safely stop the operation of part of or a complete plant can be very complex and it is often difficult to foresee all the consequences of everything that happens during a shutdown. A dynamic simulation model of the plant is an invaluable tool to set up an ESD system properly. Far too often it is seen that a dynamic model is used when an incident has taken place to analyze the exact causes of it when the use of such a model during the design phase might have prevented the occurrence of the incident all together.

Modeling the behavior of the plant under emergency shutdown conditions requires the model to be quite detailed and the simulation is usually more challenging than other applications. Under ESD conditions, much of the equipment is shut down, many flows are stopped, and any engineer knows that mathematics and zeros do not go very well together.

13.2.1.3 Start-Up Procedure Definition

Modeling the start-up procedure also requires a detailed model. Although generally the dry start condition also involves lots of stopped equipment and zero flows, it tends to be somewhat easier to model than an ESD scenario.

The use of a dynamic model to verify the start-up procedures of a plant can reduce the commissioning time by weeks. This exercise consists of adding the start-up logic to the model and to run this start-up logic while observing the behavior of the plant model. When problems occur, the model can be stopped and the start-up logic can be reviewed and rerun. Not only does the start-up procedure become streamlined, the engineers that have worked on it acquire a detailed understanding of the behavior of the plant, which allows them to make better decisions during plant commissioning and subsequent operation of the plant.

13.2.1.4 Distributed Control System (DCS) Checkout

DCS checkout alone will not warrant the construction of a dynamic model of a plant, but if a model is available, the modifications needed to be able to run a DCS checkout are relatively small. The purpose of the DCS checkout is to verify that all the cabling connecting the DCS to the plant and the DCS internal TAG allocations are hooked up properly. Obviously a dynamic model will not be able to help with checking the physical cabling, but the signals from a dynamic model can replace the plant signals. This will help tremendously in verifying the logical connections inside the DCS. If a wrong measurement is routed to a particular controller, this will be seen quite readily as the dynamic model provides realistic numbers for these. It is much easier to discern an erroneous number among realistic numbers than to match quasi-random numbers.

13.2.1.5 Operator Training

A classical use of dynamic simulation has been for operator training systems. Nowadays this is just one of the applications that can be built on top of a detailed dynamic model. In most new projects an operator training system is becoming a standard requirement and it is yet another driver to start using dynamic simulation early in plant design, as part of the earlier work can be reused in an operator training system.

In addition to the dynamic model, operator training systems are composed of various other parts.

- The operator stations that mimic the real DCS operator stations or that serve as spare operator consoles.

- An instructor station to allow the instructor to monitor the student's progress and to let him introduce a selection of failures or other problems the operator may encounter in the real plant.
- Possibly an automated system to assign a score to the performance of the operator and/or to let the operator run predefined training scenarios.
- Software and hardware for the communication between the various modules.

13.2.1.6 Advanced Process Control

Advanced process control, particularly multivariable predictive control (MPC), normally requires access to plant data and step test results from the operating plant. Hence, MPC is usually only implemented once the plant has been commissioned. With a dynamic model available, this is no longer a limitation and the necessary information can be obtained from the step tests on that model. This part is covered in more detail later.

13.2.2 Plant Operation

A dynamic model in plant operation will usually require justification on a single application. Although future uses of a dynamic model are certainly possible, these will usually not be considered to be a tangible reason or additional justification to create such a model. However, it is quite often easier to quantify the benefits that will be gained from the use of a dynamic model.

13.2.2.1 Troubleshooting

Issues in the control or operability of the plant can be resolved easier, safer, and with no loss of production using a dynamic model. A dynamic model can be exercised at will where the engineer has very little freedom to test things out on the real plant. Maintaining production on spec will almost always override the need for testing to solve a problem. The solution imagined by the plant engineer is not necessarily the right one and implementing an untested solution may lead to unsafe operation.

With a dynamic model, the worst that can happen is that the model fails. The engineer can test a large range of operating conditions to ensure that the implemented solution will hold up in abnormal conditions, for example.

13.2.2.2 Plant Performance Enhancement

Most plant engineers will accept it as a given that many of the controllers that are in a plant are operated in manual. This is frequently a source of added operating cost for the plant. A typical example would be the reflux rate of a distillation column that is set to a fixed relatively high value to ensure that the product is always meeting the specifications. Most of the time, say 85% of the time, the reflux could be run significantly lower and only during less than 5% of the time the fixed reflux is needed to maintain product quality. It is fairly easy to compute the financial benefit associated with keeping the controller in automatic.

The reason that controllers are put in manual is often related to the trust operators have in controllers. There are generally two possible reasons for distrust. Either the operator does not understand properly how the controller will cope with upsets or the controller has shown in the past that it is incapable of dealing with those upsets.

Even without a full operator training system, a dynamic model can be used to show the operators how a series of typical upsets will be handled by the control system. Some of the responses may appear illogical to the operators at first, and with a model the operator will not feel the pressure to act prematurely in order to ensure product quality. This can help instill more confidence in the automated control and let it run in automatic mode.

Of course, the distrust of the operator may be well founded, but in this case the dynamic model can be used as described in Section 13.2.2.1 to improve the controller behavior and subsequently illustrate to the operator that the problem has been solved to restore trust.

13.2.2.3 Incident Analysis

Although this is by no means the best use of a dynamic model, it is all too often the first step toward using dynamic simulation. After an incident there is always the need to know why it happened. If the incident resulted in damages, there will be legal requirements to determine the root cause of the incident. A dynamic process simulation model will often be used in this analysis to determine the sequence of events on the process side that led up to the incident and how adequate the emergency shutdown procedures were to mitigate the consequences of the incident.

13.2.2.4 Operator Decision Support

Operator decision support is an emerging use of dynamic simulation models. In this type of application the dynamic simulation model is run in real time and is receiving the same input signals as the real plant. It is impossible to cover the entire plant with instrumentation to provide all the information one would like to obtain. The real-time model provides the operators and engineers with simulated measurements throughout the entire plant to better appreciate current operation. Typical parts of a plant that do not have all the instrumentation one would like are long pipelines and high temperature outlets of reactors or furnaces. A second use of the online model is its predictive capability. Assuming the dynamic simulation model is fast enough, it can be used to predict events minutes or even hours ahead of the actual event. This information can be used to improve the handling of the event and to keep the plant operating within specifications.

13.2.2.5 Operator Training

It is important for the operators to keep their knowledge of the plant operation up to date. Especially with highly automated plants it is important that operators are confronted with abnormal situations using the simulator. New operators will also benefit from the use of an operator training system.

If is therefore very important to keep the operator training system that was installed as part of the plant commissioning up to date. This means that any change to the DCS screens and systems must also be made to the operator training system and that any change made to the plant must be made in the dynamic process model as well.

13.2.2.6 Advanced Process Control (APC)

Implementation of an advanced process control system requires a significant investment and such a decision is not taken lightly. A dynamic simulation model can assist in determining the relevance of an APC implementation and it can help streamline the implementation itself.

It is fairly straightforward to run the necessary step tests for the implementation of a multivariable predictive controller (MPC) on a dynamic process simulation. The results can be used to design the MPC and to run the MPC on the dynamic model. A comparison of the plant performance using the existing control system with the MPC can provide the necessary information to decide if an MPC implementation is an attractive investment. Running step tests on a model has a number of advantages over step tests on the real plant.

- No disturbance of the plant operation
- Step tests can use a broader range of conditions
- Step tests do not depend on the availability of plant personnel, plant incidents, and other events that are not compatible with step tests on the plant
- The dynamic model does not suffer from valves that get stuck and other incidents that make life difficult when performing step tests
- The dynamic model can be run faster than real time and hence step tests that would otherwise take days can be run in an hour or less

When the final MPC controller has been designed and implemented, it can first be put online using the dynamic simulation model. This setup can be used to discover a significant part of the practical problems that would otherwise only surface during the commissioning of the MPC on the real plant. Although the MPC models should be verified for actual plant operation, this accelerates MPC commissioning and lowers the risk of production loss that may be experienced during commissioning of the MPC.

13.3 MODELING CONSIDERATIONS

13.3.1 Level of Detail in the Model

The level of detail required for a dynamic simulation model is very dependent on the application. Most of the time a model will contain components that are modeled in great detail (high fidelity) while other components only capture the overall dynamic behavior. A typical example is a model of a gas compressor.

An initial application is to use the model to analyze the behavior of the antisurge control logic. In this case it will be important to properly model all the gas volumes in the main gas lines and also in the antisurge system. The control logic used in the model will be an exact replication of the commercial system that will be installed. As compressor surge is a very fast phenomenon, these controllers have sampling times on the order of 50 milliseconds, the model will need to run using a time step that is capable of capturing these phenomena and hence have a time step that is even smaller than 50 milliseconds. As a consequence, the model may run slower than real time. However, as in this case, the time span of interest is a few minutes at the most, the slower model performance is not really a problem.

A second application for this dynamic model of a compressor station may be for operator training. It is not relevant to model events that happen so fast that it is impossible for an operator to respond during the event. On the one hand, the step size and hence the speed of the model can be increased. On the other hand, it will be important to include manually operated purge valves in the dynamic model to allow the operator to perform all actions he deems necessary. This is a bit of detail that is of no use for the initial application.

The level of detail required should be assessed based on the objectives for the model. This assessment is not a global assessment for the complete model, but the assessment should consider the objectives for each section of the plant down to each piece of equipment.

13.3.2 Model Speed

The speed of a model is expressed most frequently as the real-time factor of the model. This is the ratio of the simulated time divided by the real time. The speed requirements vary dependent on the application. For an operator training simulator, it is clear that the model needs to be capable of runoff running at least in real time (real time factor = 1). Quite often the real-time factor should be higher, up to 10 times real time. This allows the operator to fast forward through periods of stabilization of the process, for example.

For an engineering study, the important factor is the total amount of time it takes to study an event. Ideally, that time would be 10 minutes or less. This means that 3-hour events should have a real-time factor of 18 or higher. If the event to be studied only lasts for one minute, then a real time factor of 0.1 is acceptable.

The model speed is mainly affected by the following factors.

- Time step of the integrator
- Complexity of the model

- Number of components used to represent the fluids and the complexity of the thermodynamic model

As the same factors will also affect the accuracy of the model, a balance must be found between speed and accuracy.

13.3.3 Equipment-Specific Considerations

The following sections recommend the information to consider when modeling various pieces of equipment. Depending on the available modeling tool, the recommended level of detail may differ. Recommendations only apply to the main aspects of the model.

13.3.3.1 Valves

The minimum requirement to model a valve properly is to use the correct Cv value and the type of valve characteristic. For some studies it is important to capture the dynamic behavior of the valve. For example, an emergency shutdown valve needs a certain time to close. For safety studies it is important to consider the time to close. Most plants will have one or more check valves. It is important to include these valves in the model, particularly when the model is used to run scenarios far away from normal operation.

13.3.3.2 Rotating Equipment

For pumps, compressors, and expanders, it is best to always use the performance curves of the equipment. If this information is not available, it is relatively simple to create a generic performance curve starting from the normal operating point of the equipment. This information is then complemented with either the speed or the absorbed power. For most motor-driven equipment, a speed specification will be the best option except when studying start-up and shutdown phenomena. For equipment driven by a gas turbine, a specification of the absorbed power is usually a better choice.

If the study concerns the start-up or shutdown of the equipment, it will be necessary to include details such as rotor inertia, friction losses, and dynamics of the driver (e.g., an electrical motor) in the model.

13.3.3.3 Piping Equipment

The level of detail for modeling the piping depends a lot on the application. For process piping, it is quite often sufficient to model the pressure loss. Most of the time the volume of the piping is negligible compared to the volume of the process equipment. A notable exception is the modeling of compressor antisurge systems. An accurate representation of the system volume is crucial in obtaining correct results. For transport pipelines the model should usually be more detailed, as the expected results may include information such as the time lag of a product in the pipeline, the evolution of the temperature, and the multiphase behavior of the fluid. The required information includes the pipeline elevation profile, the pipeline diameter, pipe schedule, insulation, and environment.

13.3.3.4 Columns

Distillation models should reflect the holdup volumes of both the liquid and the vapor phase properly. A significant difference between tray columns and packed columns is that the much smaller liquid holdup of packed columns will reduce the response time of the column compositional perturbations, for example.

It is customary to use a reduced number of theoretical stages to model distillation columns in steady-state simulations. If the same approach is used in a dynamic simulation, the tray or packing characteristics will be modified to use the correct holdup volumes for liquid and vapor. Another approach is to use tray efficiencies and keep the number of trays used equal to the number of real trays in the column. Keep in mind that the tray efficiency and the column overall efficiency are not the same.

13.3.3.5 Heat Exchangers

The level of detail for a heat exchanger will strongly depend on the role of the heat exchanger and the phenomena to be studied. For example, if the exchanger serves to cool down a condensate stream before it proceeds to storage and the focus of the study is on the equipment upstream of this exchanger, then it may be enough to use a model that simply assumes the exchanger is always cooling the fluid down to the required temperature. At the opposite side of the spectrum there would be a plate fin heat exchanger in a natural gas liquefaction plant. In this case the exchanger is the heart of the plant and the model needs to accurately represent the construction of the exchanger and to take into account elements such as the heat capacity of the metal and the dynamics of the metal temperature. The model will need to provide information such as the temperature and pressure profiles inside the exchanger.

13.3.3.6 Control Systems

Contrary to steady-state simulation, the modeling of the control equipment is crucial to the success of a dynamic simulation model. Quite often the control strategy and controller tuning is the final objective of the dynamic simulation, but without proper configuration of the control system the model will quickly end up in totally abnormal operating conditions.

For regular proportional, integral, and derivative (PID) controllers, the main points to consider are correct direction of the action (reverse or direct) taken, realistic tuning constants, and proper span of the instrumentation. Once the simulation model has reached relatively stable conditions, attention can focus on a high-fidelity representation of the control systems.

The high-fidelity representation can come in various forms. For an operator training system, most of the DCS vendors will be able to provide software that emulates the DCS system. The model itself is then only used to represent the noncontrol equipment. The DCS emulation will receive the plant measurements from the model like the real DCS would receive the measurements from the plant and the DCS will send signals to the valve positioners according to the control algorithms defined in the DCS.

The verification of a surge controller for a compressor is also an area where a high-fidelity representation of the particular controller is crucial. The representation can be built by combining blocks that are part of the dynamic simulator, by writing a custom model for the controller, by linking the model to an emulation program, or even linking the model to the controller hardware.

13.4 CONTROL OF EQUIPMENT AND PROCESS SYSTEMS

This section enumerates some typical applications of dynamic simulation in the various processes that are employed in gas treatment and transportation. The application is usually governed by the particular equipment used in the process.

13.4.1 Gas Gathering and Transportation

The key equipment in gas gathering and transportation are pipelines, valves, and compressors. The range of applications in this area is very wide.

- Assess the risk of condensate accumulation and associated slug sizes
- Line packing capacity studies
- Safety studies on pipeline shutdowns
- Pipeline depressuring studies
- Compressor station antisurge control studies

13.4.2 Gas Treating

The main equipment used for the absorption of CO_2 and H_2S are absorption and regeneration columns using amine solutions to absorb the acid gas components. A dynamic model will prove useful if the quality of the feed gas can fluctuate significantly. In such a case, the product gas quality will be affected directly by the amount of amine solution that is circulated and also by the quality of the lean amine solution, which is influenced indirectly by the amount of acid gas in the feed.

13.4.3 Sulfur Recovery

The performance of a sulfur recovery unit is mainly governed by the operation of the various reactors in the process. A key factor in the reactor performance is the correct air–gas ratio of the reactor feed. If the acid gas feed is unstable, dynamic modeling can be used to select the best control strategy to cope with these fluctuations and to improve the controller tuning (Young *et al*., 2001).

13.4.4 Gas Dehydration

The glycol gas dehydration process is very similar to the acid gas removal process. Fluctuations in water content of the feed gas and gas flow rate affect the product gas quality. The control strategy ensures quality of the product by selecting the appropriate glycol flow rate and by maintaining the quality of the lean glycol.

For dehydration processes using mole sieve beds the application of dynamic simulation is similar to an application for plant start-up. The model would include the logic that is driving the bed switching and regeneration cycles. Once implemented in the model, the logic can be tested by running it on the model and by tracking critical operating parameters of the mole sieve unit.

13.4.5 Liquids Recovery, Natural Gas Liquefaction

A cold box is a key piece of equipment in these processes. The flows that exchange heat in the cold box exchanger create a multitude of thermal loops in the process, which make it more difficult to control. A detailed dynamic model of the overall process, including a detailed model of the cold box, will help understand the severity of the interactions created by these thermal loops. A control strategy designed to cope with these interactions can be tested thoroughly (Valappil, 2005).

From the perspective of pressures and flows, the operation of turboexpanders or turboexpanders coupled to compressors is important. The efficiency of a turboexpander drops quickly as one deviates from the design conditions, and the impact of a temporary deviation of the operating conditions on the process dynamics is difficult to understand without help. A dynamic model will aid in understanding the behavior and in selecting the correct control structure and controller tuning to cope with a transient deviation from design conditions.

13.4.6 NGL Fractionation

NGL fractionation is composed of a series of distillation columns. As the purity specs on the columns are fairly severe on both the top and the bottom products, the control of the columns is not straightforward. A dynamic model will provide the capability to select the best control strategy given the particular column operation and specifications and given the expected disturbances in the feeds.

13.5 CASE STUDY I: ANALYSIS OF A FUEL GAS SYSTEM START-UP

This study analyzes the start-up philosophy of the fuel gas system of a Latin American offshore platform. The fuel gas system PID is shown

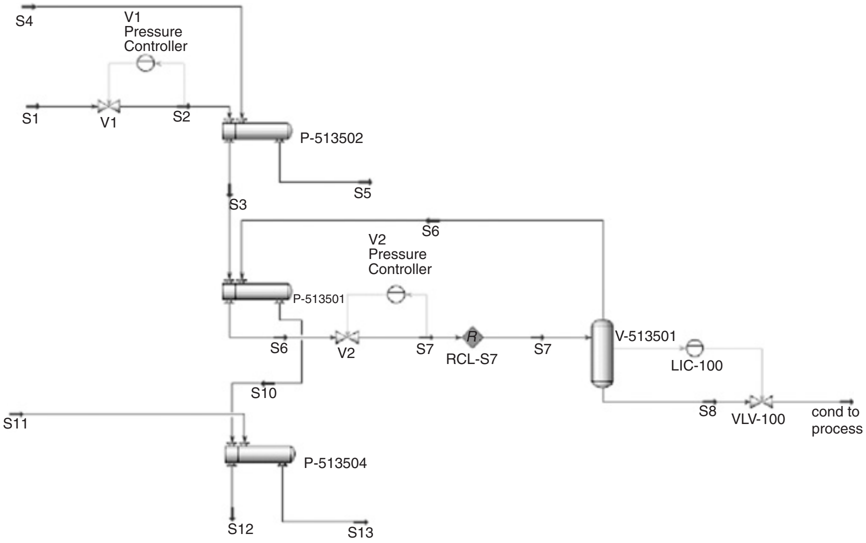


Figure 13-1. Fuel gas system process flow diagram.

in Figure 13-1. It is thought from steady-state analysis that the system can be started up using cold pipeline gas without preheating (Wassenhove, 2003).

13.5.1 Introduction

Under normal operation the fuel gas burnt in turbogenerators (TGs) comes from the platform main compression system at 180 kg/cm^2 and 38°C. At the system inlet, the pressure is reduced to 100 kg/cm^2 and the temperature goes down to 17°C. The fuel gas is then preheated in P-513502 to 60°C with hot water and another pressure reduction to 45 kg/cm^2 through a Joule–Thompson (JT) valve, resulting in a temperature of around 16°C. After this last pressure reduction, the mixture of condensate and gas is sent to a condensate vessel drum (V-513501), the condensate returns to the process, and the gas from the vessel proceeds through a gas–gas heater exchanger (P-513501). This fuel gas from P-513501 at an average temperature of 43°C is reheated with hot water in heat exchanger P-513504 to around 63°C and is sent to the turbo generators.

The objective of the system is to produce a fuel gas stream with a defined rate, a defined pressure, and a temperature at least 20°C above the dew point. This is a minimum value demanded by the turbogenerator vendor. The fuel gas exit temperature must also be maintained above 0°C to meet material temperature limitations.

During the start-up/restart of the platform there is no gas source on the platform but cold pipeline gas can be imported at 5°C. Also, there is no hot water available for the preheaters. One solution is to start the turbogenerators with diesel. This solution is undesirable, as diesel needs to be imported onto the platform, the turbogenerators need to be adapted to cope with diesel feed, and several other problems lead to production inefficiency and unnecessary cost.

The objective of the dynamic simulation is to study the exact behavior of the fuel gas system under the start-up condition of cold pipeline import gas and the circulation of the sea water at 25°C to heat the gas.

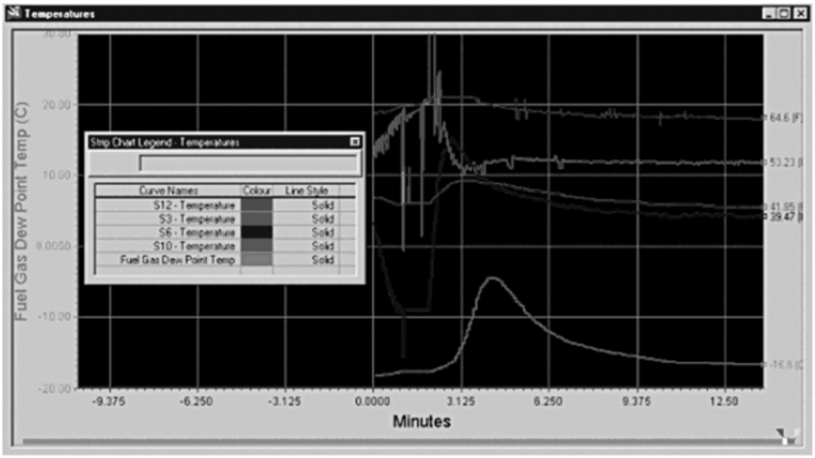
13.5.2 Steady-State Analysis

The steady-state results clearly show that thermodynamically and theoretically both normal operating and cold start scenarios result in gas feeds to the turbo generators that are well above the dew point approach limitation of 20°C. Hence it seems feasible to provide fuel gas with a temperature at 20°C above dew point. However, this does not account for the transients encountered during startup. The question can only be answered conclusively by a dynamic analysis.

13.5.3 Dynamic Analysis

In the first start-up scenario water is circulating at 25°C before any gas is fed. Once the water flow is stable, cold pipeline gas is brought on stream. The strip charts from a HYSYS dynamics model displaying the system temperatures for this case are shown in Figure 13-2.

As the cold import gas hits the warmer exchangers the gas is heated and some of the heavier components flash off in the condensate drum and pass back through the gas–gas chiller. The dew point of the fuel gas then rises for about 4 minutes until eventually the chill duty in the recycle and the JT effect in valve V2 stabilize. The temperature of the fuel gas outlet remains relatively constant as the heat exchange in P-513504 is established. Hence there is a closer approach between the fuel gas temperature and the dew point, reaching 24°C at 4.1 minutes. This is too close to the limit to accept without further study.



Curve Names	Colour	Line Style
S12 - Temperature	■	Solid
S3 - Temperature	■	Solid
S6 - Temperature	■	Solid
S10 - Temperature	■	Solid
Fuel Gas Dew Point Temp	■	Solid

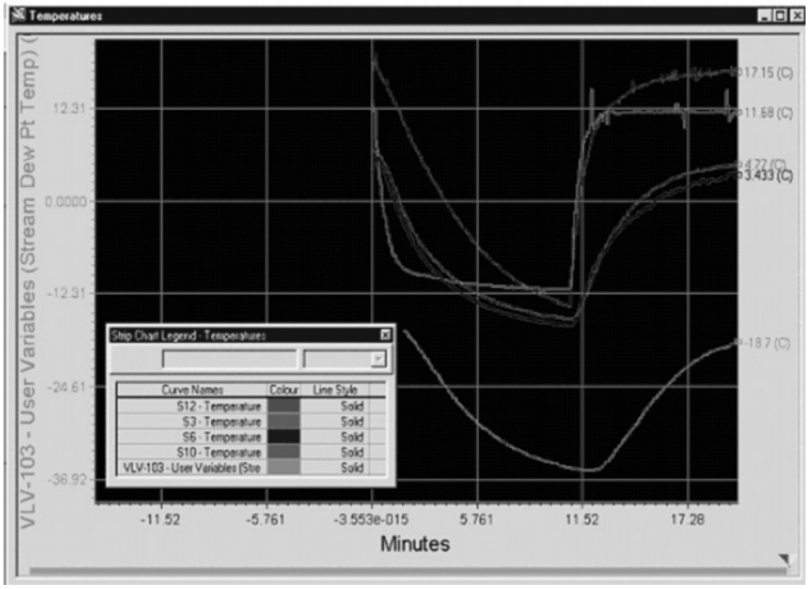
Figure 13-2. Cold water circulating, import pipeline gas: system temperatures.

The second start-up scenario assumes that the cold pipeline gas flow is established first and that the water system is brought on line afterward. The strip charts of the system temperatures in this case are shown in Figure 13-3.

Because the JT effect and cold duty on the gas-to-gas exchanger, the gas dew point does not initially increase but maintains a difference from stream 12 temperature. In this mode they would eventually equate. Also the fuel gas exit temperature is decreasing and within 4 minutes it will reach the material temperature limit of 0°C. On start up of the water system the temperature recovers after 10 minutes and the fuel gas always maintains more than a 40°C difference from the dew point.

13.5.4 Conclusion

A dynamic model clearly shows potential problems with all start-up modes of the platform fuel gas system with minimum approaches to dew point or minimum material temperatures. However, the dynamic model also demonstrates a combination of procedures, starting with the water off and then increasing flow quickly, that could maintain all proper flow rates and temperatures for fuel gas start up without diesel, thus saving millions of dollars.



Curve Names	Colour	Line Style
S12 - Temperature	■	Solid
S3 - Temperature	■	Solid
S6 - Temperature	■	Solid
S10 - Temperature	■	Solid
VLV-103 - User Variables (Stre	■	Solid

Figure 13-3. Import pipeline gas on, bring up water system: system temperatures.

13.6 CASE STUDY II: ONLINE DYNAMIC MODEL OF A TRUNK PIPELINE

The 235-mile NOGAT trunk line is located in the Dutch part of the North Sea, and it connects eight offshore platforms to onshore gas processing facilities near Den Helder, The Netherlands. Each platform delivers both gas and condensate to the pipeline so the line operates inside the two-phase region; two of the platforms sit on oil fields and the off gas from the oil stabilization units is compressed and delivered to the pipeline. The current total capacity of the gas transportation system is about 22 million m^3/day, associated with 750 m^3 of condensate/day (La Riviére and Rodriguez, 2005).

The onshore facilities include a 1000-m^3 slug catcher, condensate stabilization units that remove volatile components from the trunk line produced liquids, and a series of low-temperature separation (LTS) units to dry the sales gas prior to delivery to the distribution network. Figure 13-4 shows a representation of the system.

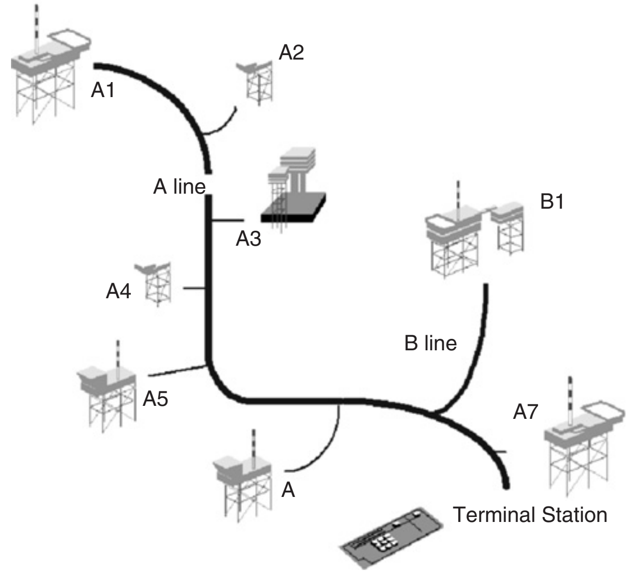


Figure 13-4. NOGAT trunk line system (La Riviére and Rodriguez, 2005).

There are two major challenges to operating the system.

1. Controlling the sales gas quality in terms of its Wobbe index.[1] The different platforms produce different gas qualities and quantities that are fed into the line at different locations. Therefore, the quality of the gas that travels through the trunk line varies along its length and also with time. Despite this, the sales gas quality must stay within the contractual limits at all times (i.e., Wobbe index values between 49 and 54).
2. Controlling the condensate inventory of the trunk line. The amount of condensate retained inside the line builds up during periods of low gas demand, particularly in trunk line depressions. The available slug catchers and condensate stabilization unit capacities limit the production ramp-up speed and force the scheduling of periodic cleanup cycles to keep the trunk line liquid holdup below certain critical limits.

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[1] The Wobbe index is a measure used to compare the equivalent thermal content characteristics of different gases. It is defined as a volumetric high heating value divided by the square root of relative density to air.

This chapter discusses the elements of automating today's gas processing plants, including considerations for instrumentation, controls, data collection, operator information, optimization, and management information. The advantages and disadvantages of various approaches are analyzed in this chapter. Also, strategies for identifying and quantifying the benefits of automation are discussed.

12.2 EARLY METHODS OF GAS PLANT AUTOMATION

The earliest gas processing plants were typically controlled manually by opening, pinching, and closing valves to meet their operating requirements. Pneumatic control systems were adopted quickly. These pneumatic control systems allowed the use of proportional, integral, and derivative (PID) controllers to send analog outputs to control valves to change their opening. As long as a sensor is available as a process variable for the controller, then a set point given by the operator can be targeted automatically. Discrete control could also be accomplished with pneumatic control systems by employing devices to fail with or without an air signal. In most cases, compressed air is used as the pneumatic conveyor; however, natural gas is used in some remote operations and hydraulic oils are also employed. A reliable source of clean, oil-free air is quite important in the operation of pneumatic air systems.

As electronic controls were developed, these systems became the standard, although some elements of the pneumatic control systems are still in use. Most control valves today are still pneumatically operated for new and existing plants even though electronic control valves have been on the market for decades. Electronic controllers were accepted due to their lower price and greater reliability. These controllers included fewer moving parts to maintain. For new installations, wire was now run instead of pneumatic tubing with electronic to pneumatic (I/P for current to pressure) transducers added at the control valves. Process sensors such as temperature, pressure, level, and flow indicators were converted to electronic types as available and justified. Electronics also brought widespread use of safety systems such as vibration sensors, burner management, and emergency shutdown systems. Other special controllers, such as dedicated surge controllers for rotating equipment and triple modular redundant shutdown devices, were developed.

cells than the online model (32) and performs its calculations 250 times faster then real time.

While the predictive model to some extent sacrifices accuracy, predictions are always found to be influenced more by the estimates and forecasts of future platform flows provided by the user as simulation inputs than by model inaccuracies.

Platform discharge pressures and gas delivery flow are computed by the simulation models and stored by executive application back into the data historian. Space discrete information, such as pressure, temperature, flow rates, velocities, and holdup profiles, is archived in a dedicated database. The historian-user interface for the system shows actual conditions vs model-calculated data. Predictive what-if or look-ahead studies are triggered from the system graphical interfaces. The predictive model retains the process and trunk line internal conditions to serve as an initial state for simulations. The user is given the option to edit individual platform flows and onshore landing pressure or flow. The time horizon of the prediction is also an input value.

Simulation results provided by the online and predictive parts of the tool are available to the operators by means of graphical user interfaces (GUIs) that were constructed using historian tools. Predictive results are shown as a series of tables and charts that track the evolution of line conditions (i.e., Wobbe index, compositions, condensate holdup, velocities, flows, and pressure) along the trunk line distance and as a function of time. Figure 13-5 shows a comparison between the observed and the simulated Wobbe index of the gas at the terminal station.

Connecting the process engineering models to online measurement data stored in historians has been successful for the support of operational decisions. The described computer program was installed in the control room of the onshore processing facilities in March 2003, and the operators successfully utilize the system for predicting and managing the gas quality. The calibrated models developed for the online system are now also used by process engineers for offline analysis of the clean-out cycles for managing the large amount of condensate, which develops inside the trunk line, for example. Potential slug catcher overflow was prevented by a correct adjustment of platform flows. The model centric approach of the system and the clear split between application components and the simulation model also facilitates model maintenance and upgrades. The speed of calculation is more than sufficient to keep the online model synchronized with real-time events. Similarly, the time response of the predictive case

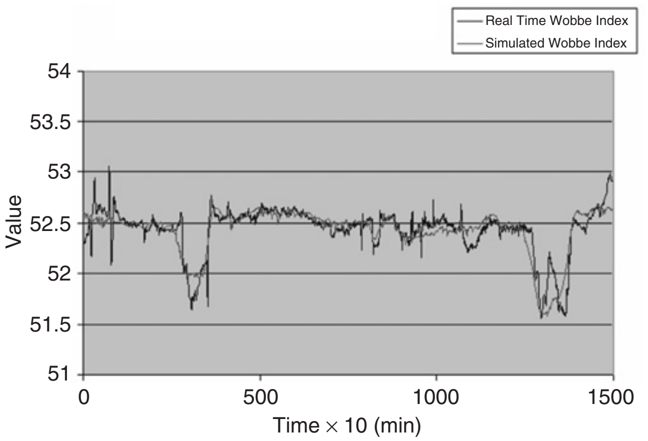

Figure 13-5. Comparison between the real and the simulated Wobbe index (La Riviére and Rodriguez, 2005).

is satisfactory for the end users. The accuracy of the calculated results for gas quality is more than sufficient to support the necessary operational decisions. Although the accuracy of the results regarding liquid inventories is not clear at present, the availability of these data is already a major improvement with respect to the previous situation when information on this aspect of the process was completely missing.

REFERENCES

Brown, C., and Hyde, A., "Dynamic Simulation on the Shell Malampaya Onshore Gas Plant Project." Paper presented at the GPA Europe Annual Conference, Rome, Italy (Sept. 25–27, 2001).

La Riviére, R., and Rodriguez, J-C., Program tracks wet-gas feedstocks through two-phase offshore trunkline." *Oil Gas J.* **103**(14), 54–59 (2005).

Valappil, J., Messersmith, D., and Mehrotra, V., LNG lifecycle simulation. *Hydrocarb. Eng.* **10**, 10 (2005).

Wassenhove, W.V., Analysis of fuel gas system startup with HYSYS dynamic process simulation. *Oil Asia J.* 18–20 (January–March 2003).

Young, B., Baker, J., Monnery, W., and Svrcek, W.Y., Dynamic simulation improves gas plant SRU control-scheme selection. *Oil Gas J.* **99**, 22 (2001).

SUGGESTED READING

James, G., and Reeves, J., “Dynamic Simulation Across Project and Facility Lifecycles.” Paper presented at the 6th World Congress of Chemical Engineering, Melbourne, Australia (Sept. 23–27, 2001).

Muravyev, A., Kelahan, R., Kowallis, P., and Torgesen, G., “Dynamic Modeling of the Plant Gas System: Process Control and Design Applications.” Paper presented at the ISA Expo 2003 Conference, Houston, TX (Oct. 21–23, 2003).

CHAPTER 14

ENVIRONMENTAL ASPECTS OF GAS PROCESSING AND USE

14.1 INTRODUCTION

Currently, natural gas represents approximately 24% of the energy consumed in the United States with increases in use projected for the next decade. These increases are expected because emissions of greenhouse gases are much lower with the consumption of natural gas relative to other fossil fuel consumption. For example, natural gas, when burned, emits lower quantities of greenhouse gases and criteria pollutants per unit of energy produced than other fossil fuels. This occurs in part because natural gas is fully combusted more easily and in part because natural gas contains fewer impurities than any other fossil fuel. However, the major constituent of natural gas, methane, also contributes directly to the greenhouse effect through venting or leaking of natural gas into the atmosphere.

Natural gas is seen by many as an important fuel in initiatives to address environmental concerns. Although natural gas is the most benign of the fossil fuels in terms of air pollution, it is less so than nonfossil-based energy sources such as renewable sources or nuclear power. However, because of its lower costs, greater resources, and existing infrastructure, natural gas is projected to increase its share of energy consumption relative to all other fuels, fossil and nonfossil, under current laws and regulations. Indeed, natural gas consumption and emissions are projected to increase more rapidly than other fossil fuels, at average annual rates of 1.7% through 2020. However, this represents reductions in totalcarbon emissions derived from the environmental advantages of natural gas. Concern about global warming and further deterioration of the environment caused by escalating industrial expansion and other development is being addressed by worldwide initiatives that seek a decrease in emissions of greenhouse gases and other pollutants. Natural gas is expected to play a key role in strategies to lower carbon emissions because it allows fuel users to consume the same Btu level of energy while less carbon is emitted. If carbon-reduction measures are implemented, natural gas demand will increase. However, emissions from natural gas consumption would also rise, but the natural gas share of total emissions would increase only slightly.

This chapter describes efforts on the part of the natural gas industry to lessen the environmental impact of natural gas processing and discusses the environmental effects of the use of natural gas, including comparing the emissions from natural gas to other fossil fuels.

14.2 ENVIRONMENTAL IMPACTS OF NATURAL GAS PROCESSING

The processing of natural gas poses low environmental risk, primarily because natural gas has a simple and comparatively pure composition. However, two important sources of emissions in the gas processing plants are discussed.

14.2.1 Air Pollutant Emissions

The atmosphere is a mixture primarily of the gases nitrogen and oxygen, totaling 99% with almost 1% water and very small amounts of other gases and substances, some of which are chemically reactive. With the exception of oxygen, nitrogen, water, and the inert gases, all constituents of air may be a source of concern due either to their potential health effects on humans, animals, and plants or to their influence on the climate. In the United States the Clean Air Act (CAA), which was last amended in 1990, mandates that the Environmental Protection Agency regulates *criteria pollutants* that are considered harmful to the environment and public health.

There are 188 substances that are identified as air toxics or hazardous air pollutants, with lead being the only one that is currently classified as a criteria pollutant and thus regulated. Air toxic pollutants are more acute biological hazards than most particulate or criteria pollutants but are much smaller in volume. Procedures are now underway to regulate other air toxics under the CAA.

In the natural gas and refining industries, as in other industries, air emissions include point and nonpoint sources. Point sources are emissions that exit stacks and flares and, thus, can be monitored and treated. Nonpoint sources are *fugitive emissions* that are difficult to locate and capture. Fugitive emissions occur throughout refineries and arise from, for example, the thousands of valves, pipe connections, seals in pumps and compressors, storage tanks, pressure relief valves, and flanged joints. While individual leaks are typically small, the sum of all fugitive leaks at a refinery can be one of its largest emission sources. These leaks can release methane and volatile organic compounds (VOCs) into the air. Companies can minimize "fugitive emissions" by designing facilities with the fewest possible components and connections and avoiding components known to cause significant fugitive emissions. When companies quantify fugitive emissions, this provides them with important information they can then use to design the most effective leak repair program for their company. Directed inspection and maintenance programs are designed to identify the source of these leaks and to prioritize and plan their repair in a timely fashion. A reliable and effective directed inspection and maintenance plan for an individual facility will be composed of a number of components, including a methods of leak detection, a definition of what constitutes a leak, set schedules and targeted devices for leak surveys, and allowable repair time. A directed inspection and maintenance program begins with a baseline survey to identify and quantify leaks. Quantification of the leaks is critical because this information is used to determine which leaks are serious enough to justify their repair costs. Repairs are then made only to the leaking components that are cost effective to fix. Subsequent surveys are then scheduled and designed based on information collected from previous surveys, permitting operators to concentrate on the components that are more likely to leak. Some natural gas companies have demonstrated that directed inspection and maintenance programs can profitably eliminate as much as 95% of gas losses from equipment leaks.

Most refinery process units and equipment are manifolded into a collection unit, called the blowdown system. Blowdown systems provide for the safe handling and disposal of liquid and gases that are either automatically vented from the process units through pressure relief valves or that are manually drawn from units. Recirculated process streams and cooling water streams are often purged manually to prevent the continued buildup of contaminants in the stream. Part or all of the contents of equipment can also be purged to the blowdown system prior to shutdown before normal or emergency shutdowns. Blowdown systems utilize a series of flash drums and condensers to separate the blowdown into its vapor and liquid components. The liquid is typically composed of mixtures of water and hydrocarbons containing sulfides, ammonia, and other contaminants, which are sent to the wastewater treatment plant. The gaseous component typically contains hydrocarbons, hydrogen sulfide, ammonia, mercaptans, solvents, and other constituents and is either discharged directly to the atmosphere or is combusted in a flare. The major air emissions from blowdown systems are hydrocarbons in the case of direct discharge to the atmosphere and sulfur oxides when flared.

Other potential hazardous air pollutant emission points are the tail gas streams from amine-treating processes and sulfur recovery units.Emissions from the sulfur recovery unit typically contain some hydrogen sulfide (H_2S), sulfur oxides, and nitrogen oxides. To address the risk associated with sour gas exposure and risks associated with explosions, sour facilities are generally required to have lower rates of fugitive emissions than gas plants that do not process sour gas (sweet gas facilities) and to have in place emergency response plans that address the hazard of hydrogen sulfide exposure.

Other emissions sources from refinery processes arise from the periodic regeneration of catalysts. These processes generate streams that may contain relatively high levels of carbon monoxide, particulates, and VOCs that are usually referred to as natural gas liquids or natural gasoline. Before being discharged to the atmosphere, such off-gas streams may be treated first through a carbon monoxide boiler to burn carbon monoxide and any volatile organic compounds and then through an electrostatic precipitator or cyclone separator to remove particulates. Facilities that dispose of acid gas by deep well injection generally have far lower emissions of sulfur dioxide than facilities that recover sulfur or that flare acid gas. Acid gas injection provides the additional benefit of "sequestering" the carbon dioxide part of the acid gas stream. Acid gas disposal is a proven technology that a growing number of new and existing oil and gas facilities are using. Companies must dispose of acid gas in an underground formation where there is no chance that the gas will escape and contaminate other formations. Workers must specifically design the acid gas injection well bore to handle highly corrosive wet acid gas.

Glycol dehydrators can also be significant sources of hydrocarbon emissions. Most gas processing plants use glycol dehydration to remove water from gas as it enters a pipeline to prevent hydrate formation and corrosion. In addition to extracting water from natural gas, glycol will extract some benzene, toluene, ethyl benzene, and xylene (collectively referred to as BTEX) molecules. When heat is used to regenerate the glycol, both water and BTEX molecules are driven off (Fitz *et al.*, 1987). Operators commonly vent these emissions to the atmosphere. There are a number of ways operators can reduce or eliminate emissions from glycol dehydrators. On existing dehydration units, operators can reduce emissions by optimizing the dehydration unit, for example, reducing the glycol circulation rate to the minimum required to ensure adequate freeze protection, or optimizing the temperature of the unit (Burns, 1999). Operators can also use a separator to remove water from gas before it enters the dehydration unit; this also reduces the amount of glycol they need to use in the dehydrator and reduces the quantity of emissions. If plant operators collect vapors from the regeneration column, the glycol pump, and any gas-operated instrumentation and then flare or incinerator these vapors, they can achieve near-zero emissions from glycol dehydrators. Other alternatives, which can be an improvement over glycol, include (1) line heater to heat the gas at the well site and raise the gas temperature above the freezing point and (2) molecular sieve dehydrator, which is typically a closed system that removes water by heating the crystals to above the boiling point of water. This releases the water and regenerates the crystals so that they can be reused. This process almost eliminates vapor and BTEX emissions. Because of its closed system process, the molecular sieve is suitable for dehydration of sour gas where the release of H_2S could be lethal.

The burning of large quantities of natural gas to drive engines, motors, and heaters is also a key source of air emissions (primarily carbon dioxide and nitrogen oxides) from gas processing facilities. If facilities use electrically powered units, this will avoid the creation of local combustion emissions. However, if using electricity requires power generation facilities to be expanded or new sources of power to be developed, this may contribute to the creation of emissions and impacts wherever the power is generated.

Because medium to large gas processing facilities contain many sources of air pollution, they can emit a large volume of pollutants into the atmosphere. For this reason, when regulatory authorities issue operating permits or approvals to companies running these facilities, they specify which processing units are allowed to emit pollutants to the air, and they sometimes set limits on how much of a pollutant the facility can release. Companies should place air quality monitoring equipment at their property line or close to the plant so that they can monitor air contaminants of concern. Companies usually place this equipment in the area where they expect there to be the highest concentrations of an air pollutant at ground level (also called the “maximum point of impingement”). Regulators can use the monitoring devices to make sure the facility is operating in compliance with the approval it was given.

14.2.2 Gas Flaring Emissions

Flaring is used to consume waste gases (including hydrogen sulfide-rich gases and gases burned during emergencies) in a safe and reliable manner through combustion in an open flame. It is used routinely to dispose of flammable gases that are either unusable or uneconomical to recover. Sometimes gas plant workers must do emergency flaring for safety purposes when they depressure equipment for maintenance.

In theory, the complete combustion of pure hydrocarbons produces only water and carbon dioxide. Low-efficiency flares do not completely combust all of the fuel gas, and unburned hydrocarbons and carbon monoxide are emitted from the flare with the carbon dioxide. If the waste fuel entering the flare contains impurities and/or liquid droplets, many other by-products can also be emitted from the flare stack. These products include particulate matter, VOCs such as benzene, toluene, and xylene, polycyclic aromatic hydrocarbons (PAH), and small quantities of sulfur compounds such as carbon disulfide (CS_2) and carbonyl sulfide (COS). Flaring is both a concern to the public and a government priority because of the potential health risks and environmental concerns associated with the activity and also because it wastes a valuable nonrenewable resource. Furthermore, the noise, odor, and smoke produced from flaring activities can interfere with nearby residents and their enjoyment of the outdoors. Flaring is an environmental concern regarding to global warming and acid deposition. Emissions of carbon dioxide and unburned natural gas from flares contribute to the greenhouse gas effect and global warming.

Flaring at sweet and sour gas plants and acid gas injection facilities can generate a large volume of air emissions that could negatively affect local air quality. Sometimes it will be less expensive for plant operators to flare gas for a long time while they try to fix a problem instead of shutting down the plant completely. Operators can reduce the impacts of flaring at gas processing/disposal facilities by minimizing the total volume of gas they flare at the plant and by reducing the frequency and duration of upset flaring events. Operators can prevent air impacts from ongoing flaring by shutting down all or part of a facility in phases, according to preset time increments. This will still give them some time to resolve any problems while the plant is running.

The efficiency of a flare is a measure of how effective that flare is in converting all of the carbon in the fuel to CO_2. Previous studies have indicated that flares have highly variable efficiencies, on the order of 62–99%. More recent research concludes that the yearly averaged combustion efficiency of gas flares could exceed 95% if certain design and operating conditions are met. Gas flares are operated in uncontrolled conditions. The tips of the flare can be exposed to wind, humidity, and temperature variations that reduce efficiency and increase variability. Other factors that can adversely affect the efficiency of gas flares are the composition of the waste gas stream entering the flare, which varies from site to site, and improper flaring practices that cause unsteady combustion conditions. Poor flare performance is often associated with incomplete combustion, leading to smoke and unburned hydrocarbon emissions. Under such conditions a review of the flare system will often bring benefit in operational cost, and ensuring minimum purge gas is demanded by the requirements of safe flare operation. An operation flare review will give operators the ability to discuss specific operational issues with a flare consultant and to examine the flexibility in the operation to enhance the flare system performance to achieve improved performance in terms of environmental, economic, and safety goals. Miles (2001a) examined how a flare system can be designed safely to meet the challenge of reducing emissions to the environment as well as reducing the cost of operating the overall flare system.

Incineration can potentially be a more efficient method to dispose of waste gases, although a more costly and sophisticated one. If operated properly, incinerators generally have more efficient combustion than flares because combustion occurs in an enclosed chamber, away from the effects of wind and weather, and the air-to-fuel ratio required for complete combustion can be controlled precisely. Although they can be highly efficient, incinerators are mainly used at sour gas processing plants and not for routine waste gas flaring. The reasons for this are that incinerators are more costly to install, require more frequent maintenance and monitoring, and are difficult to install and operate in remote locations. Other alternatives to flaring include (1) conserving the waste gas for processing at natural gas facilities, (2) reinjecting the waste gas underground to maintain reservoir pressure during production, (3) using the gas to power microturbine generators for electricity production, and (4) ensuring that flare systems are properly designed, constructed, and maintained through guidelines, codes of practice, or regulation. There are several proven economic and environmental benefit solutions now in service. Flare gas recovery (the zero flare option) is a major step forward in securing a brighter and cleaner future for the industry. Flare gas recovery, however, may not be possible in all cases. There is a basic precondition that the gas recovered has to be useful. Fuel gas-deficient facilities are ideal; however, future trends toward centralized power generation may make a significant future impact. Where flare gas recovery is simply not feasible, then the best flare solutions should be sort. Staged flares with the benefit of variable orifice solutions probably offer the best all round solution. The combination of these flare systems and flare gas recovery will give economic and environmental benefit (Miles, 2001b).

14.2.3 Methane Emissions

Methane is both the primary constituent of natural gas and a potent greenhouse gas when released to the atmosphere. Reducing methane emissions can yield substantial economic and environmental benefits. The implementation of available cost-effective methane emission reduction opportunities in the gas processing industry can lead to reduced product losses, lower methane emissions, and increased revenues.

Methane losses from natural gas processing plants account for 15% of total worldwide methane emissions. Emissions result primarily from normal operations, routine maintenance, and system disruptions. Emissions vary greatly from facility to facility and are largely a function of operation and maintenance procedures and equipment conditions. Some of the more significant sources of methane emissions and selected technologies and practices applicable to the gas processing section are presented (Fernandez *et al.*, 2005a).

14.2.3.1 Pneumatic Devices

At processing facilities without electrical power, workers can use pneumatic devices, which can run on natural gas from oil and gas formations, to drive pumps as well as power instrumentation and control equipment. High-bleed pneumatic devices can be major sources of methane emissions. Alternative technology is available that, while still using natural gas to drive pumps and instruments, does not vent to the atmosphere. Many companies in all natural gas sectors have achieved significant savings and methane emissions reduction by replacing, retrofitting, or improving maintenance of the high-bleed pneumatic devices. Field experience shows that up to 80% of all high-bleed devices can be replaced with low-bleed equipment or retrofitted. Although low-bleed devices cost more, most operators that install them (either initially or as a retrofit) end up making their money back on the investment. Another option available at facilities with available electric power is to replace their natural gas-powered pneumatic control systems with compressed instrument air systems, eliminating 100% of emissions from pneumatics. However, instrument air systems require electrical power to be available on site. In some cases, bottled nitrogen instead of instrument air or natural gas can be used for smaller services.

14.2.3.2 Dehydrator Systems

Many natural gas *dehydrator systems* in the processing sector use triethylene glycol to remove water from the natural gas stream in order to meet pipeline specifications. Often the glycol circulation rate is set much higher than it needs to be in order to achieve this objective. Overcirculation leads to increased methane emissions. Operators can adjust this circulation rate at no additional cost and decrease methane emissions from the dehydrator system. Another way to decrease methane emissions from triethylene glycol systems is to install flash tank separators. Flash tank separators capture approximately 90% of the methane entrained in the triethylene glycol systems, preventing the methane from being boiled off into the atmosphere when the triethylene glycol systems passes through the regenerator.

Desiccant dehydrators can be good alternatives to triethylene glycol systems under certain circumstances. Glycol dehydrators can be replaced altogether with desiccant dehydrators at a higher cost, nearly eliminating gas emissions and also saving both fuel gas used for the glycol reboiler (sometimes a gas heater) and pneumatic gas used for glycol unit controllers. With no regenerator, desiccant dehydrators produce methane emissions only when they are being refilled with desiccant, and even then the volumes are far below those from triethylene glycol systems.

14.2.3.3 Vapor Recovery Units

During storage of condensates, methane and other gases vaporize and collect in the space between the liquid and the fixed roof of the tank. As the liquid level in the tank fluctuates, these vapors are often vented to the atmosphere. One way production sector companies can prevent these emissions is to install *vapor recovery units* on condensate storage tanks. A vapor recovery unit draws over 95% of the hydrocarbon vapors out of a storage tank or set of tanks under low pressure. The vapors are then routed to a scrubber and then used as an on-site fuel supply or sold.

14.2.3.4 Compressors

Reciprocating compressors are the dominant type of compressors used in the gas processing industry. A Gas Research Institute (GRI) report (U.S. EPA and GRI, 1996) estimates that reciprocating compressors formed 85% of the total compressor population in the gas processing industry. Components associated with reciprocating compressors are subject to high thermal and vibrational stresses that make them prone to leaks and are therefore among the largest source of emissions in the gas plant. Clearstone Engineering Inc. (Clearstone) data estimate much higher emissions from compressor seals than the GRI study. Clearstone has also identified valves and connectors as big sources of emissions that are not explicitly listed in the GRI study (U.S. EPA and Clearstone, 2002). Thus, according to Clearstone, compressor seals, valves, and connectors are the top three emission sources in that order and together constitute 90% of the emissions from reciprocating compressors. This rate would justify the practice of cost-effective rod packing replacement and could potentially save another billion cubic feet of methane emissions annually. Replacement of wet seals with dry seals also leads to substantially reduced operating and maintenance expenses, improved reliability, and reduced contamination of the gas.

Centrifugal compressors are represented as a smaller source of emissions compared to reciprocating compressors. GRI shows an individual centrifugal compressor emitting more than an individual reciprocating compressor. However, due to the low numbers of centrifugal compressors used in the gas processing sector, they do not contribute as much as reciprocal compressors to the total emission volume (Fernandez *et al.*, 2005a).

14.2.3.5 Cryogenic Equipment

This category only contains turboexpanders and the equipment associated with them. This category of equipment was not identified in the GRI study and so is a “new” source of methane emissions. The total emissions from turboexpanders are small and contribute less than 3% of total methane emissions in a gas plant. The individual components contributing to emissions are valves, connectors, pressure-relief valves, compressor seals, and open-ended lines.

14.2.3.6 Flares

This category includes process flares and flare system piping. The GRI study does not identify flares as a source of methane emissions. Clearstone data estimate flares to be the largest single source of methane emissions, contributing more than 40% of the total methane emissions at a gas plant. It is unclear from Clearstone data whether the flares were lit or unlit when evaluated or what the source of gas in the flare stack was. However, Clearstone provides a total hydrocarbon emissions rate that was determined by engineering calculations rather than direct measurement. The methane content of this gas was assumed to be the same as the fugitive emissions from all other components.

14.2.3.7 Methane Emissions Reduction

In natural gas plants, one can reduce methane emissions by upgrading technologies or equipment and by improving management practices and operational procedures. Opportunities to reduce methane emissions generally fall into one of three categories.

- Technologies or equipment upgrades, such as low-emission regulator valves, that reduce or eliminate equipment venting or fugitive emissions.
- Improvements in management practices and operational procedures to reduce venting.
- Enhanced management practices, such as leak detection and measurement programs, that take advantage of improved measurement or emission reduction technology.

There are numerous ways natural gas companies can reduce their methane emissions, and many of these technologies and practices cost less to implement than the value of the gas they save (Fernandez *et al.*, 2005b). Cost-effective opportunities for reducing methane emissions in the gas processing plants vary greatly from country to country based on the levels of physical and institutional infrastructure. Many of the available cost-effective abatement options and technologies, however, can be applied universally throughout the gas processing industry.

14.2.4 Water Pollution

Refineries are also potential contributors to surface water and groundwater contamination. Wastewater in refineries may be highly contaminated and may arise from various processes (such as water from cooling, distillation, product fractionator reflux drum drains, and boiler blowdown). This water is recycled through many stages during the refining process and goes through several treatment processes, including a wastewater treatment plant, before being released into surface waters. The wastes discharged into surface waters are subject to state discharge regulations and are regulated under the Clean Water Act (CWA). These discharge guidelines limit the amounts of sulfides, ammonia, suspended solids, and other compounds that may be present in the wastewater. Although these guidelines are in place, contamination from past discharges may remain in surface-water bodies (Speight, 2005).

When managing surface water, companies must prevent contact between rainwater and contaminants at the processing facility. They must also design the sites properly to ensure that rainwater that lands on a gas plant is collected in a central location where it can be stored, treated, and reused or released into the environment in a controlled manner. Companies that use effective liners and secondary containment systems, and continue to monitor groundwater throughout the life of the project, can ensure that leaks and spills are prevented and any contamination that does happen can be detected and cleaned up quickly.

14.2.5 Soil Pollution

Contamination of soils from the refining processes is generally a less significant problem than that of contamination of air and water. Soil contamination, including some hazardous wastes, spent process catalysts, tank bottoms (this can be minimized by using recirculating pumps or mixers inside tanks to keep heavier parts suspended throughout the condensate rather than allowing them to collect on the bottom), sludge from the treatment processes, filter clay, and incinerator ash can occur from leaks as well as accidents or spills on- or off-site during the transport process. Treatment of these wastes includes incineration, land treating off-site, land filling on-site, land filling off-site, chemical fixation, neutralization, and other methods (Speight, 1996; Woodside, 1999).

14.2.6 Pollution Prevention

Pollution prevention is the reduction or elimination of pollutants to the environment. Pollutants such as hazardous and nonhazrdous wastes and regulated and unregulated chemicals from all sources may be discharged as air emissions, wastewater, or solid waste. All of these wastes are treated. However, air emissions are more difficult to capture than wastewater or solid waste. Thus, air emissions are the largest source of untreated wastes released to the environment.

The limits of pollutants emitted to the atmosphere, land, and water are defined by various pieces of legislation that have been put into place over the past four decades. Following passage of the Pollution Prevention Act (PPA) of 1990, the U.S. EPA developed a formal definition of pollution prevention and a strategy for making it a central guiding mission. Under Section 6602(b) of the PPA, Congress established a national policy that

- pollution should be prevented or reduced at the source whenever feasible;

- pollution that cannot be prevented should be recycled in an environmentally safe manner whenever feasible;
- pollution that cannot be prevented or recycled should be treated in an environmentally safe manner whenever feasible; and
- disposal or other release into the environment should be employed only as a last resort and should be conducted in an environmentally safe manner.

The particular suite of practices and measures adopted for a specific gas processing project should reflect local circumstances. Not all of the best practices or measures just listed are appropriate in all cases. However, this hierarchy of preferred options for dealing with environmental pollution officially places prevention at the top of the list.

Pollution prevention can be accomplished by reducing the generation of wastes at their source or by using, reusing, or reclaiming wastes once they are generated. However, as stated before, environmental analysis plays a major role in determining if emissions–effluents (air, liquid, or solid) fall within the parameters of the relevant legislation.

Pollution prevention is often touted as an economically advantageous, strategically wise way for companies to protect the environment while protecting themselves (from liability, legal infractions, and unforeseen or unnecessary costs). Setting up a pollution prevention program does not require exotic or expensive technologies. Some of the most effective techniques are simple and inexpensive. Others require significant capital expenditures, but many provide a return on that investment. Speight (2005) presents a description of various methods by which effluents and emissions are treated.

14.3 EMISSIONS FROM NATURAL GAS USE

14.3.1 Combustion Emissions

Composed primarily of methane, the main products of the combustion of natural gas are carbon dioxide and water vapour. The combustion of natural gas also produces significantly lower quantities of other undesirable compounds, particularly toxics, than those produced from the combustion of petroleum products or coal. The toxic compound benzene can be a component of natural gas, but the levels in natural gas are considered insignificant and are not generally monitored by gas processing plants and most pipeline companies. However, because the benzene content of the natural gas varies by source and can range from less than 0.4 to 100 ppm, depending on the efficiency of the combustion, some of the benzene will be oxidized to carbon dioxide and water, some will pass through unburned, and some will be converted to other toxic compounds.

Natural gas is the cleanest of all the fossil fuels. Coal and oil are composed of much more complex molecules, with a higher carbon ratio and higher nitrogen and sulfur contents. This means that when combusted, coal and oil release higher levels of harmful emissions, including a higher ratio of carbon emissions, nitrogen oxides (NO_x), and sulfur dioxide (SO_2). Coal and fuel oil also release ash particles into the environment, substances that do not burn but instead are carried into the atmosphere and contribute to pollution. The combustion of natural gas, however, releases very small amounts of sulfur dioxide and nitrogen oxides, virtually no ash or particulate matter, and lower levels of carbon dioxide, carbon monoxide, and other reactive hydrocarbons. Coal and oil plants beget masses of solid waste—up to 590 tons per day—while gas plants create none. Natural gas plants also release less waste heat due to their higher efficiency.

The most significant conventional pollutants released by gas combustion are oxides of nitrogen (NO_x) formed by heating air around the point of combustion. Harmful to human health itself, NO_x combines with airborne hydrocarbons to form ozone, a pervasive urban scourge. NO_x emissions are a precursor of airborne particulate pollution, which causes over 50,000 deaths per year in the United States (Dockery *et al.*, 1994). Because of their low NO_x emissions, some renewable energy technologies can make a greater immediate impact on environmental problems than natural gas plants. Advanced gas combustion technologies also reduce NO_x emissions significantly, although the majority of plants in service now use older technologies.

The environmental case for using natural gas as a bridge to a renewable energy future can be summarized as follows. First, natural gas pollutes more than renewable energy but less than oil or coal. Second, the supply of gas has limits, but for at least the next few decades, gas can generate far more electricity than all the renewable energy technologies combined. Third, due to its immediate availability, natural gas can displace many more tons of coal now and in the near future than renewable energy can.

14.3.2 Acid Rain Formation

Acid rain is another environmental problem that affects much of the eastern United States, damaging crops, forests, and wildlife populations and causing respiratory and other illnesses in humans. Acid rain is formed when sulfur dioxide (SO_2) and nitrogen oxides (NO_x) react with water vapor and oxidants in the presence of sunlight to produce various acidic compounds, such as sulfuric acid and nitric acid. Precipitation in the form of rain, snow, ice, and fog causes about half of these atmospheric acids to fall to the ground as *acid rain*, while about half fall as dry particles and gases. Winds can blow the particles and compounds hundreds of miles from their source before they are deposited, and they and their sulfate and nitrate derivatives contribute to atmospheric haze prior to eventual deposition as acid rain. The dry particles that land on surfaces are also washed off by rain, increasing the acidity of runoff.

The principal source of acid rain causing pollutants, sulfur dioxide and nitrogen oxides, are coal-fired power plants. Because natural gas emits virtually no sulfur dioxide and up to 80% less nitrogen oxides than the combustion of coal, increased use of natural gas could provide for fewer acid rain causing emissions.

14.3.3 Smog Formation

Smog is a pressing environmental problem, particularly for large metropolitan cities. The primary constituent of smog is ground-level ozone created by photochemical reactions in the near-surface atmosphere involving a combination of pollutants from many sources, including motor vehicle exhausts, volatile organic compounds such as paints and solvents, and smokestack emissions. Because the reaction to create smog requires heat, smog problems are the worst in the summertime.

Natural gas use is not much of a factor in smog formation. As opposed to petroleum products and coal, the combustion of natural gas results in relatively small production of smog-forming pollutants as it emits low levels of nitrogen oxides and virtually no particulate matter (Speight, 1993, 2005). For this reason, it can be used to help combat smog formation in those areas where ground-level air quality is poor. The smog-forming pollutants literally cook in the air as they mix together and are acted on by heat and sunlight. The wind can blow smog-forming pollutants away from their sources while the reaction takes place, explaining why smog can be more severe miles away from the source of pollutants than at the source itself.

14.3.4 Greenhouse Gas Emissions

The Earth's surface temperature is maintained at a habitable level through the action of certain atmospheric gases known as *greenhouse gases* that help trap the sun's heat close to the Earth's surface. The main greenhouse gases are water vapor, carbon dioxide, methane, nitrous oxide, and several man-made chemicals, such as chlorofluorocarbons (Fogg and Sangster, 2003). Most greenhouse gases occur naturally, but concentrations of carbon dioxide and other greenhouse gases in the Earth's atmosphere have been increasing since the industrial revolution of the 18th century with the increased combustion of fossil fuels and increased agricultural operations. Of late, there has been concern that if this increase continues unabated, the ultimate result could be that more heat would be trapped, adversely affecting Earth's climate. Consequently, governments worldwide are attempting to find some mechanisms for reducing emissions or increasing absorption of greenhouse gases.

One of the principal greenhouse gases is carbon dioxide. Although carbon dioxide does not trap heat as effectively as other greenhouse gases (making it a less potent greenhouse gas), the sheer volume of carbon dioxide emissions into the atmosphere is very high, particularly from the burning of fossil fuels. On a carbon-equivalent basis, 99% of anthropogenically sourced carbon dioxide emissions in the United States are due to the burning of fossil fuels, with 22% of this attributed to natural gas.

One issue that has arisen with respect to natural gas and the greenhouse effect is the fact that methane, the principal component of natural gas, is itself a very potent greenhouse gas. Methane contributes directly to the greenhouse effect because of its ability to trap heat in the atmosphere, which is estimated to be 21 times greater than the ability of carbon dioxide to trap heat. According to the Energy Information Administration (EIA), although methane emissions account for only 1.1% of total U.S. greenhouse gas emissions, they account for 8.5% of the greenhouse gas emissions based on the global warming potential.

Water vapor is the most common greenhouse gas, at about 1% of the atmosphere by weight, followed by carbon dioxide at 0.04% and then methane, nitrous oxide, and man-made compounds, such as chlorofluorocarbons. Each gas has a different residence time in the atmosphere, from about a decade for carbon dioxide to 120 years for nitrous oxide and up to 50,000 years for some of the chlorofluorocarbons. Water vapor is omnipresent and cycles into and out of the atmosphere continually. In estimating the effect of these greenhouse gases on climate, both the global warming potential (heat-trapping effectiveness relative to carbon dioxide) and the quantity of gas must be considered for each of the greenhouse gases.

Oil and gas companies should develop and carry out greenhouse gas management plans to minimize the cost of complying with the Kyoto protocol, ratified by Canada in December 2002, and with subsequent emission reduction requirements. Such plans should include ways to reduce emissions through internal energy efficiency, investments in offsets and "green power," and a commitment to limiting absolute volumes of emissions.

14.3.5 Industrial and Electric Generation Emissions

The use of natural gas to power both industrial boilers and processes and the generation of electricity can significantly improve the emissions profiles for these two sectors. Natural gas is becoming an increasingly important fuel in the generation of electricity. As well as providing an efficient, competitively priced fuel for the generation of electricity, the increased use of natural gas allows for improvement in the emissions profile of the electric generation industry. According to the National Environmental Trust (NET) in their 2002 publication entitled "Cleaning up Air Pollution from America's Power Plants," power plants in the United States account for 67% of sulfur dioxide emissions, 40% of carbon dioxide emissions, 25% of nitrogen oxide emissions, and 34% of mercury emissions. Coal-fired power plants are the greatest contributors to these types of emissions. In fact, only 3% of sulfur dioxide emissions, 5% of carbon dioxide emissions, 2% of nitrogen oxide emissions, and 1% of mercury emissions come from noncoal-fired power plants.

Natural gas-fired electric generation and natural gas-powered industrial applications offer a variety of environmental benefits and environmentally friendly uses, including fewer emissions, reduced sludge, reburning, cogeneration, combined cycle generation, and fuel cells. Essentially, electric generation and industrial applications that require energy, particularly for heating, use the combustion of fossil fuels for that energy. Because of its clean-burning nature, the use of natural gas wherever possible, either in conjunction with other fossil fuels or instead of them, can help reduce the emission of harmful pollutants.

14.4 PROTOCOLS AND ENVIRONMENTAL PROGRAMS

Natural gas is a collection of pollutant chemicals. Such gaseous emissions are often characterized by chemical species identification, e.g., *inorganic gases* such as sulfur dioxide (SO_2), nitrogen oxides (NO_x), and carbon monoxide (CO) or *organic gases* such as the hydrocarbon constituents of natural gas. The rate of release or concentrating in the exhaust air stream (in parts per million or comparable units) along with the type of gaseous emission greatly predetermines the applicable control technology. Recommended protocols must occur as a preclude to the cleanup of emissions and mitigating future releases. Necessary actions for the cleanup of natural gas emissions are (1) identification of the emissions, (2) identification of the emission sources, (3) estimation of emission rates, (4) atmospheric dispersion, transformation, and depletion mechanisms, (5) emission control methods, (6) air-quality evaluation methods, (7) effects on stratospheric ozone, and (8) regulations.

Members of the natural gas industry have a commitment to ensuring that their operations are environmentally sound and that every effort is being made to ensure that the environmental impacts of activities related to the processing of natural gas are as minimal as possible. Part of this commitment includes participating in voluntary industry programs aimed at maintaining the best possible environmental record for the natural gas industry. These programs include the following.

- The EPA natural gas STAR program. This program, sponsored by the EPA, is intended to reduce methane emissions—a potent greenhouse gas—from the oil and gas industry by implementing improved technologies, better work practices, and improved maintenance and inspection of distribution networks in order to minimize leaks and emissions.
- The API STEP program. The American Petroleum Institute runs a program called STEP (Strategies for Today's Environmental Partnerships). This program serves to encourage petroleum industry members to commit to environmental stewardship in their policies and principles and to develop programs to ensure safe, environmentally sound operating practices.

In addition to the aforementioned programs, the U.S. natural gas industry is actively involved in international programs that serve to share "best practices" with respect to environmental preservation. Usually, this is accomplished through affiliation with an association dealing with significant environmental matters.

In addition to industry participation in maintaining a healthy environment, the federal government has enacted a number of pieces of legislation to ensure that the natural environment is preserved and maintained into the future. The EPA is the primary federal government agency charged with protecting human health and safeguarding the natural environment. The EPA develops and enforces environmental regulations that exist under environmental laws in the United States, as well as leading a number of voluntary and educational programs intended to reduce pollution and protect the environment.

14.5 ENVIRONMENTAL MANAGEMENT SYSTEM

As stated earlier, organizations are increasingly concerned to achieve and demonstrate sound environmental performance by controlling the impact of their activities, products, or services on the environment, taking into account their environmental policy and objectives. For this purpose, the ISO 14001 standard has been developed to provide organizations with the elements of an effective environmental management system, which can be integrated with other management requirements, to assist organizations to achieve environmental and economic goals. This standard enables an organization to establish, and assess the effectiveness of, procedures to set an environmental policy and objectives, achieve conformance with them, and demonstrate such conformance to others. The overall aim is to support environmental protection and prevention of pollution in balance with socioeconomic needs.

The organization should review and continually improve its environment management system to achieve overall improvement in environmental performance. At regular intervals management should carry out a review of the environmental management system to ensure its continuing suitability and effectiveness. The scope of the review should be comprehensive, although not all elements of an environmental management system need to be reviewed at once and the review process may take place over a period of time.

REFERENCES

Burns, D.J., "Minimizing Benzene Emissions from Glycol Dehydrators." Paper presented at the Benzene Emissions Conference, Calgary, Alberta, Canada (January 27, 1999).

Dockery, D.W., et al., Acute respiratory effects of particulate air pollution. *Annu. Rev. Public Health* **15**, 107–32 (1994).

Fernandez, R., Robinson, D., and Aggarwal, V., "Methane Emissions and Reduction Opportunities in the Gas Processing Industry." Paper presented at the 84th Annual GPA Convention, San Antonio, TX (March 15, 2005a).

Fernandez, R., Petrusak, R., Robinson, D., and Zavadil, D., Cost-effective methane emissions reductions for small and midsize natural gas producers. *J. Petr. Technol.* 35–41 (June 2005b).

Fitz, C.W., Hubbard, R.A., and Campbell, J.M., Quick manual calculation estimates amount of benzene absorbed in glycol dehydrators. *Oil Gas J.* **72** (Nov. 23, 1987).

Fogg, P.G.T., and Sangster, J.M. (eds.), "Chemicals in the Atmosphere: Solubility, Sources, and Reactivity." Wiley, Hoboken, NJ (2003).

Miles, J.D., "Toward Zero Flaring: A Practical Approach to Cost Reduction and Environmental Benefit." Paper presented at the 80th GPA Annual Convention, San Antonio, TX (March 2001a).

Miles, J.D., "A Flare Gas Recovery System." Paper presented at NEL Flare Gas Metering Seminar, Aberdeen, UK (May 2001b).

Speight, J.G., "Gas Processing: Environmental Aspects and Methods." Butterworth Heinemann, Oxford, England (1993).

Speight, J.G., "Environmental Technology Handbook." Taylor & Francis Group, Philadelphia, PA (1996).

Speight, J.G., "Environmental Analysis and Technology for the Refining Industry." Wiley, Hoboken, NJ (2005).

U.S. EPA and GRI, "Methane Emissions from Natural Gas Industry," Vol. 8. NTIA, Springfield VA, Publication No. EPA-600/R-96-080h (1996).

U.S. EPA and Clearstone Engineering Ltd., "Identification and Evaluation of Opportunities to Reduce Methane Losses at Four Gas Processing Plants." GRI-02/0119 (2002).

Woodside, G., "Hazardous Materials and Hazardous Waste Management." Wiley Interscience, New York (1999).

CHAPTER 15

MAXIMIZING PROFITABILITY OF GAS PLANT ASSETS

15.1 INTRODUCTION

Maximizing the return on gas plant assets becomes increasingly difficult because of the rising cost of energy in some cases and the increased demand for operations agility in most cases. On top of these demands, there is the constant need to increase the availability and utilization of these plants. It is worthwhile to consider techniques and the profit improvement of some of the world's best gas plants. This analysis identifies some key manufacturing and business strategies that appear to be necessary and are feasible for many facilities to achieve and sustain satisfactory performance. The essential nature of operating a gas plant is dealing with change. Disturbances such as slugging, trips of gathering compressors, weather changes, and changes in market demands make it challenging to operate reliably let alone profitably. A stabilization strategy is necessary but insufficient for most gas plants; it is necessary to sustain a degree of flexibility. A useful objective can be tracking desired changes and withstanding undesired changes in a manner that is safe, environmentally sound, and profitable (DeVries *et al.*, 2001).

The main challenges to gas plant profitability are as follows.

- Continuous energy and yield inefficiency due to providing operating margin for upsets and plant swings
- Continuous energy and yield inefficiency due to not operating at optimum conditions

- Energy and yield inefficiency due to consumption during plant swings
- Labor costs to operate and support a gas plant — a poorer performing plant requires more personnel with a higher average salary
- Plant integrity
- Process and equipment reliability
- Poor yield due to low availability
- Maintenance
- Safety

Poorer performance compounds the cost challenges in most gas plants. To ensure that these problems are avoided, it is essential to develop a fundamental understanding of all the technical factors impacting the performance of the plant, how they interact, and how they manifest themselves in business performance. This can only be achieved through the application of an integrated approach utilizing the diverse skills of a multidisciplined team of engineers combined with the application of a robust performance-modeling tool. Technical assessments allow an understanding of technical risks to which the facility is exposed, ranging from deterioration of equipment due to exposure to corrosive environments to poor process efficiency as a result of poor design. Integration of expertise within a single team, which covers the broad range of technical fields involved in the operation of the gas processing plant, ensures that all the technical risks and interactions are identified.

Other business questions that asset managers are asking themselves (Howell, 2004):

- Are the assets performing to plan, and how do we know?
- Are we choosing the optimal plans for developing the assets over their lifetime?
- Are we achieving the targeted return on capital employed for the assets?
- Are we meeting all of the ever-growing health, safety, and environment guidelines?
- Can we forecast reliably, allocate with confidence, or optimize with knowledge?
- Do we derive enough value from the simulation and engineering model investments?
- How effective is the organization at capital avoidance?
- Are we drowning in data or are we knee deep in knowledge?

Another area of opportunity is optimization of the process to maximize capacity or yields and minimize energy consumption while maintaining product qualities. Many tools have become available in the past decade as powerful, inexpensive computing power has become available to run complex software quickly and reliably.

This chapter describes a vision of the integrated gas plant of the future and methods to identify solutions for attaining operational goals to maximize asset values.

15.2 THE PERFORMANCE STRATEGY: INTEGRATED GAS PLANT

Successful gas plants have found that a combination of techniques is necessary.

1. A strategy to influence organizational behavior
2. A strategy to integrate information
3. An operations strategy that uses remote operations and support of unmanned plants
4. Process performance monitoring
5. Asset management
6. Process optimization implementation

Merely adding technology without developing a new organizational behavior and operations strategy has often reduced gas plant performance instead of improving it.

The revolution in digital technologies could well transform the industry. Achieving the vision of the integrated gas plant of the future will require more than new technologies alone. It will require the alignment of strategy structure, culture, systems, business processes, and, perhaps most important, the behaviors of people. Visionary companies who truly want to capture the "digital value” will need to create a climate for change and then maintain strong leadership through the change and employ the skills and techniques from vendors to provide the technologies.

Gas plant operators are looking to integrate global operations and the energy supply chain into a cohesive picture. A global enterprise resource planning system gives companies the resources they need to better balance supply with demand, reduce inefficiencies and redundancies, and lower the total cost of information technology infrastructures. The challenge is to develop overlay solutions with more domain content that can improve the knowledge inside the enterprise systems that producers depend on most for operations, planning, project management, workflow, document management, executive information and decision support, scheduling, database management, data warehousing, and much more.

The industry must capitalize on the opportunities provided by ever more capable and cost-effective digital technology.

15.3 STRATEGIES FOR ORGANIZATIONAL BEHAVIOR AND INFORMATION

Understanding and managing organizational behavior are important elements of effective operations. The effective management and coupling of technology to people are crucial elements in ensuring that the plant's culture will change to support modern manufacturing strategies.

Four typical organizational cultures have been observed (Neumann, 1999). These cultures vary from cultures very resistant to change to cultures that change for the sake of change.

15.4 ORGANIZATIONAL BEHAVIOR MODEL

It is worthwhile to review recent modeling attempts for organizational behavior. Two models, each focusing on different aspects of the organization, deserve exploration. Both attempt to explain the influence of information and management activities on organizational behavior. The first model, as shown in Figure 15-1, attempts to identify and provide a structure for the various major influences on the organization.

This model supports the following observations.

1. Unless constrained by their situation, people will improve their behavior if they have better information on which to act.
2. Behavior is constrained by their capability to perform. This constraint is tied to the span of their control, their associated physical assets, and what the military calls "readiness."
3. The classical personal hierarchy of needs (Mazlow, 1968), which includes "survival," "hunger," and "need to belong," applies to organizations as well.

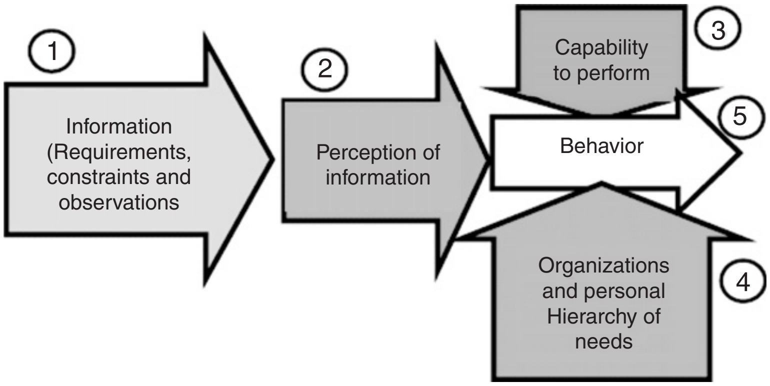

Figure 15-1. Major influences on the organization (DeVries *et al.,* 2001).

15.4.1 Information Quality

The natural gas processing industry is unusual in its high degree of dependence on information technology in order to meet its business goals. With the enormous quantities of data that it generates and processes, an edge is gained by ensuring the quality of data and using the information intelligently. In addition to this problem, the industry is addressing the changes required due to new and evolving working practices that the popularity of interdisciplinary asset teams has brought about.

This organizational model explains in part why people and organizations behave in a rebellious manner, despite receiving “better” information. For example, operators and their supervisors are penalized for “tripping” units and major equipment; however e-business decisions might require them to operate the plant in a manner that would increase the likelihood of trips. Another example is where a plant or unit’s performance is broadcast to a team of marketing and other plants, while the plant or unit is chronically “under performing” based on the other organizations’ expectations. This motivates the plant’s managers to hide information and resist plans for performance improvement.

This model also points to a key issue in managing technology. Poor-quality information will be discarded, and information that is harder to use will not be used. This points to the need to define the quality of information. Suggested attributes of information quality include the foloowing.

- Available (is the information chain broken?)
- Timely
- In the right context
- Accurate
- To the right place
- Easy to understand

It may be helpful to imagine a fuel gauge in the dashboard of a car. If the fuel gauge is working 75% of the time, the driver will not trust it — the gauge will be ignored. In process plants, information has to be much more reliable than that to be trusted.

Gas plants struggle to convert supply chain schedules into actual plans. Too often, the marketing department does not have appropriate feedback that equipment has failed or that a current swing in recovery mode is taking longer than expected to complete. Conversely, the operations department often lacks the tools to plan which resources and lineups to use, while avoiding poor utilization of equipment. This gap between strategic and tactical scheduling is sometimes referred to as the "operations wall."

Much attention is given to networking information and knowledge software. So much attention is given to making as many components as possible accessible via the Web to support virtual private networks and restructuring knowledge software using application service providers. This work is worthwhile, but the underlying information reliability issues are chronic and damage the credibility of major network and software installations.

The old expression "if garbage goes into the computer, then garbage will come out" is especially true for modern information technology strategies that end up with more than seven layers of components that process information. It is extremely important to note the following.

1. Much of the information used to enhance and change plant performance, as well as support supply chain management decisions, is based on sensors that degrade in accuracy and "fail" due to stresses or material that reduces their ability to produce a reliable measurement. A modern hydrocarbon processing plant uses hundreds of these to

support advanced control and advanced information software. Simply connecting digital networks to these sensors does not improve information reliability if the sensors do not have enough "intelligence" to detect and possibly compensate for failure or inability to measure.

2. Most components and software packages process online diagnostic information that gives some indication of the reliability of the information. However, most installations do not integrate these diagnostics with the calculations. If the information reliability of each of seven components from the sensor to the top e-business software is 95%, then on average the overall information reliability is no better than 69%. Raising the component reliability to 99% still only yields a maximum overall information reliability of 93%, which is unacceptable for supply chain management.

Therefore, it is important to use as much "intelligence" as possible at all levels and to maintain the integration of online quality statuses to maximize the quality and minimize the time to repair the failures.

15.4.2 Perception of Information

Tables of menus and numbers do not help when a downsized team is asked to open multiple applications and "mine" for information. Most information displays are essentially number tables or bar graphs superimposed on plant flowsheets. Several plant examples exist to show how this interferes with improved performance.

15.4.2.1 Two-Dimensional Curves and Plots

As an example, compressors can be the main constraining factor in major hydrocarbon processing facilities, and it is important to maintain the flow sufficiently high through each stage of compression to avoid surge. Too low a pressure increase is inefficient and limits production throughput. The most efficient is at higher pressure increases, but yields operation closer to surge. Operators are normally given a pair of numbers or bar graphs for pressure and flow, and they need to recall from experience where the "operating point" is compared to the "surge line." To make matters worse, the "surge line" changes position as the characteristics of the gas change. To solve this, plot the moving surge line and the operating

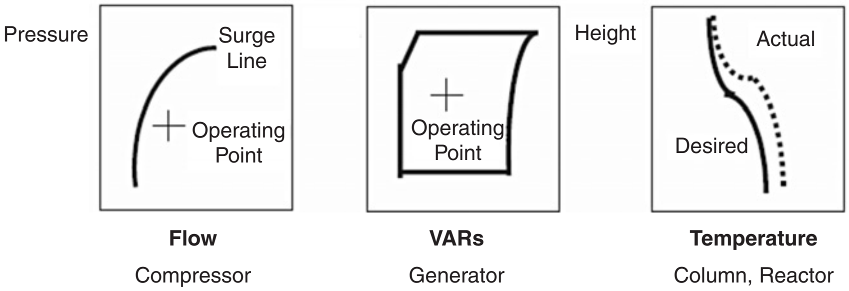

Figure 15-2. Examples of two-dimensional curves and plots (DeVries *et al.*, 2001).

point in real time. This is a suitable solution for distillation and reactor temperature profiles as well (Figure 15-2).

15.4.2.2 Prediction Trends

The same operators who avoid trips tend to shut off advanced control software when the software is driving the plant toward undesirable states. However, often the software is only using a spike to accelerate the transition to best performance, and the top of the spike still keeps the plant in the desirable state. Prediction trends allow operators to see the remainder of the curve so that if the curve looks safe, they will allow the software to continue with the spike.

15.4.2.3 Dynamic Performance Measures

Operators and their supervisors are given performance information and targets, but often the information is not "actionable," i.e. they cannot influence the information directly (Figure 15-3). Furthermore, as manufacturing strategies change, they receive e-mails with a set of numbers. Ideally, the senior operators and their chain of management are seeing the targets and acceptable bands along with each actionable performance measure. In this way, coordination of a change in a manufacturing tactic is faster and easier. A common example is changing from maximum throughput (at the expense of efficiency) to maximum efficiency (at the expense of throughout) to take advantage of market opportunities and minimize costs when the demand is reduced.

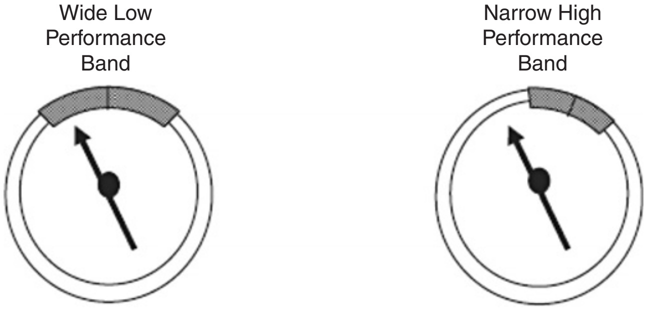

Figure 15-3. Dynamic performance dashboard examples (DeVries *et al.*, 2001).

15.4.2.4 Performance Messages

Operators who monitor long batches, especially batches that cover a shift change, tend to have lower productivity, especially during times that the main support staff is not available. Ingredients that exhibit quality problems can be used in applications such as in-line blending of fuels and batch blending with on-line analyzers to support a performance model with a set of messages and procedures that allow operators to modify the batch during the run and avoid a rework.

15.4.3 Capability to Perform

When the organization can perceive their performance and the distance away from targets, more accurately, they can develop a culture of learning how to improve. This is different than monthly accounting reports that show shadow costs and prices. This requires a condensed set of reliable information that supports factors that they can change. Examples include reliability, yield, quality, and throughput, and often it might be expressed in a different way. For example, reliability may be more "actionable" when expressed as "cycle time," "turnaround time," etc. Cutting the setup time from an overall 5 to 2.5% is actually cutting the time by 50%; this is far more vivid than showing availability changing from 95 to 97.5%.

Research has been conducted on complex operations where a unit is several steps away from the true customer, and attempts to establish internal costs and prices of feedstock, utilities, and internal products have had insufficient credibility or even ability to affect a change. Some attempts at activity-based costing have tried to increase the percentage of direct costs, with the result that the cost per unit of product is so weighted with costs outside of the unit that the unit manager cannot affect a useful improvement or is motivated to maximize throughput. Maximizing throughput is the only degree of freedom because the “costs” are relatively fixed. This is incompatible with modern manufacturing strategies that require precise, timely changes in prioritization of throughput, efficiency, and specific formulations to attract and retain key customers and more lucrative long-term contracts.

A key element of an organization’s capability to perform is its ability to handle both planned and unplanned disturbances. Modern manufacturing strategies become a nightmare if the organization and facilities cannot cope with the accelerated pace of change brought on by the tight coupling with suppliers and customers. The following model, as shown in Figure 15-4, helps describe different levels of readiness.

Containment is the lowest level of readiness, and at this level, unplanned disturbances stop production or cause product rework, but the damage is “contained” — no injury, equipment damage, or environmental release. Unplanned disturbances include operator errors, equipment failures, and deviations in feedstock quality as well as other factors. This level of readiness can be achieved with conventional information strategies.

- Loose or no integration between components, from sensors to supply chain software
- Minimal integration of diagnostics of information quality
- Minimal performance information

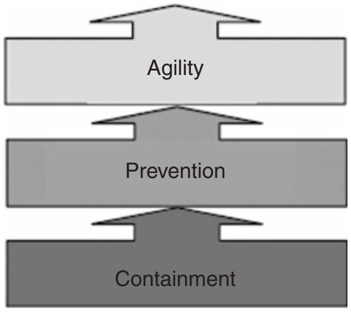

Figure 15-4. Levels of readiness (DeVries *et al.*, 2001).

- Minimal coordination between units or assets to withstand disturbances
- Reactive or scheduled maintenance strategies

Prevention is a level of readiness where unplanned disturbances rarely affect production availability or product quality. However, this operation cannot consistently support planned disturbances to production rates, feedstock quality, or changes in yield or product mix. It is usually a case of stabilizing the plant as much as possible in order to achieve this level of performance. This level of readiness requires more advanced information strategies.

- Moderate integration between components and a reliable connection to supply chain software
- Good management of information quality
- Good coordination between units and assets
- Scheduled maintenance at all levels
- Better performance information

Agility is a level of readiness where the plant can consistently support wider and faster changes in feedstock quality, production rates, and output mix without reducing production availability or quality. This operation consistently achieves "prevention" readiness so that it can outperform other plants because of its agility. This level of readiness requires the most advanced information strategies.

- The tightest integration of all levels of software and sensors
- Good models that drive performance information and online operations advice
- A reliability-centered maintenance strategy for information technology
- Good coordination with knowledge workers, who will likely come from key suppliers of catalysts, process licensors, information and automation technologies, and workers at other sites of the company

There is a tendency in many operations to try to jump from "containment" to an "agility" level of readiness by adding more information technology, but with the characteristics of "containment" (minimal integration, etc.). This is potentially disastrous. Supporting modern manufacturing strategies can mean that the quality of integration is as important or more important than the quality of the information components themselves.

Due to the complexity of the interactions between technical factors affecting asset performance and the difficulty in converting technical understanding into a business context, the traditional approach has been to try and simplify the problem. Application of benchmarking, debottlenecking studies, maintenance, and integrity criticality reviews have all been applied to enhance process plant performance.

Each of these approaches tends to focus on a particular element of an assets operation. As a result, enhancement decisions are made without a full understanding of the impact on the assets performance and the overall business impact. This can result in opportunities being missed, capital being poorly invested, and delivery of short-term benefits, which are unsustainable. In the worst cases, this can result in a net reduction in plant performance and increased life cycle costs.

15.4.4 Organizational Hierarchy of Needs

Attempts to evolve plant culture by transplanting methods and equipment have not only failed but actually eroded performance. Key observations include the following.

1. Personnel may not associate performance or knowledge with increased security, wealth, and sense of belonging or esteem. They observe promotion and compensation practices and evaluate the effectiveness of improving their situation against new expectations for better performance.
2. Risk-adverse cultures dread the concept of visual management, benchmarking, or any other strategy that is designed to help a broader teamwork together to continually improve performance. Information that shows performance can be used to support rivalries rather than inspire teams to improve.
3. Organizations and personnel worry about surviving — will the plant be shut down? Will the staffing be reduced? Efforts to change behavior using information have to be coupled to the hierarchy of needs and then the information can evolve as the organization evolves (and performance improves).

Research on dynamic performance measures has uncovered an effective model of organizational behavior that is especially appropriate for manufacturing. Two of the key issues that have been addressed are overcoming over 85 years of traditional organization structure that tries to isolate units and departments and the 20 years of cost and performance accounting.

The challenges with cost accounting are assigning costs in the correct manner. A plant manager or operations manager might face an insurmountable fixed cost. This motivates him/her to maximize production, which might be opposite to the current manufacturing requirements. The poor credibility of internal prices for utilities, feedstocks, and products also has motivated middle management to drive performance in the wrong way. These point to the need to redefine key performance indicators (KPIs). The group of research efforts to deal with this is called dynamic performance measures. The U.S. Department of Labor and many large corporations embrace dynamic performance measures.

Vollman *et al.* (1998) have developed a model that associates the manager's strategy and actions to their measures of performance. This combination is called the Vollman triangle, as shown in Figure 15-5.

The manager's subordinates will develop their own triangles to support their supervisor's measures, within the constraints of their capability to perform and so on as outlined in Figure 15-6. This occurs at all levels of the organization. The goal is to ensure that a small group of measures (no more than 4), which can be consistently affected by that level of operations, are maintained with suitable quality. If this structure is maintained, then it is easier for top management and modern manufacturing strategies to affect change.

The "actionable" measures tend to conflict with each other. Higher throughput or quality often comes at the expense of efficiency, or each has a different optimum. Each person learns how these interact. Discussions on performance improvement become more effective because personnel can now describe this behavior. The discussion is vastly different than reviewing monthly reports. As far as operations are concerned, something that occurred several days or weeks ago is ancient history.

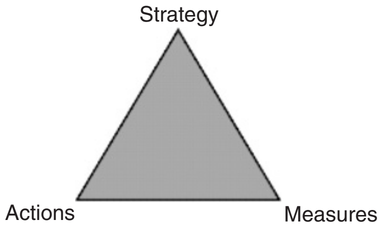

Figure 15-5. Vollman triangle.

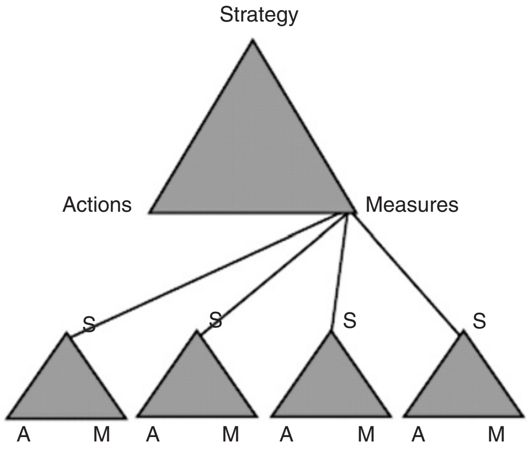

Figure 15-6. Application of Vollman triangle to subordinates (DeVries *et al.*, 2001).

15.4.5 Behavior

The most important variables for performance are work culture, job security, and career mobility. In many parts of the world, knowledge, experience, or skill is not sufficient to achieve promotion. Many cultures do not embrace an openness of sharing performance information, visual management techniques, etc. Therefore the information strategy needs to by adjusted to reflect the current human resources strategy and ideally both will evolve in phases to achieve world-class performance.

Another strong issue is traditions at the plant. It is often much easier to initiate new teamwork and performance initiatives to a new plant with a new organization. Nevertheless, the right information strategy becomes a catalyst of change. Everyone can see the performance against the current targets, and everyone can understand the faster change of priorities, such as changing from maximum production to minimum, where maximum efficiency is desired.

15.5 THE SUCCESSFUL INFORMATION STRATEGY

Technology makes it easier to measure variables such as pressure, speed, weight, and flow. However, many operations achieve profitability with very complex equipment, and the key characteristics are properties, not basic measurements. A notable example is found in oil refining. The basic manufacturing strategy is to improve yield by converting less valuable components from crude oil (molecules that are too short or too long) into valuable components (medium-length molecules). However, the essential measurement to determine the proportion of different molecules is chronically unreliable. New technology now exists to reliably indicate a property called "carbon aromaticity," which helps the operations team ensure that the process will be effective and that the feedstock will not degrade the expensive catalyst materials and erode production.

Reliable property measurement is the key. This also helps to evolve the plant culture from thinking about pressures and flows to thinking about properties. The information strategy becomes the following.

1. Assess the organization's readiness to support the desired manufacturing strategy and use of information to evolve the plant culture.
2. Depending on the organization's hierarchy of needs, develop a phased evolution of the information strategy.
3. Develop appropriate measures that will directly support the actions needed to implement the strategy at all levels.
4. Develop the appropriate information to maximize its quality, reliability, and perception.
5. Manage the quality and reliability of information.

Data integration and visualization will continue to be a key area of focus throughout the next few years in digitizing the asset. Many new applications will bring knowledge gleaning ability from the data collected.

15.6 THE IMPACT OF LIVING WITH INFORMATION TECHNOLOGY

If the organization invests in "intelligent" components that enable advanced maintenance strategies, such as reliability-centered maintenance or performance monitoring centers, then the organization is committing to the advanced strategy. Otherwise the flood of extra information will be disruptive. Intelligent sensors deliver up to 10 times as much information as conventional ones; a typical down-sized organization, with as little as one-fifth the staffing of 10 years ago, would have to deal with 50 times the volume of information — too much. The information is only useful if it is coupled with appropriate software and methods to use the information to optimize the maintenance and maximize overall information and plant availability.

If the organization invests in supply chain management technology, then the organization is committing to an "agility" level of operations readiness and a team-based performance culture. The organization is also committing to business measurement changes — cost per unit of product gets in the way of profitability measures of reducing procurement costs or increasing return on sales using modern manufacturing strategies.

The organization develops a new language and starts to use new terminology. Furthermore, managers start focusing on managing the higher levels of information. The following model, as shown in Figure 15-7, suggests a new terminology for dealing with information.

This terminology can be defined as follows.

- Data: raw information from sensors and personnel keyboard entry.
- Information: validated data, using diagnostics and techniques to enhance the quality and reliability of information.
- Knowledge: a comparison of information to targets and constraints. This answers the set of "how are we doing?" questions.
- Wisdom: guidance for best practices, customer satisfaction, supply chain management, and any other performance guidelines.

Many organizations have managers spending much of their time dealing with the "Information" function rather than focusing on "knowledge" and

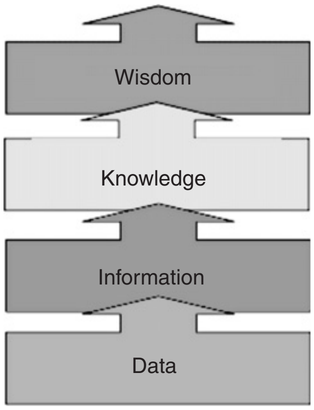

Figure 15-7. Hierarchy of information (DeVries *et al.,* 2001).

“wisdom.” It is extremely important to maximize the quality and type of information to ensure that management can support the “agility” demands of modern manufacturing strategies.

15.7 VISION OF THE MODERN PLANT OPERATION

There are four different groups of activities that require consistent information for most effective operations.

- Operational decisions on an hourly or more frequent basis by operators, shift supervisors, or engineers.
- Tactical decisions on a 1- to 30-day basis by shift supervisors, engineers, purchasing, trading, and accounting.
- Strategic decisions made on a 1-month to 5-year basis by plant management, purchasing and accounting.
- Monitoring on at least an hourly, weekly, and monthly basis by all parties involved.

Operational decisions include determination of set points for the equipment and switching equipment on or off. These decisions need to consider current pricing for commodities such as fuel and electricity, equipment availability, and environmental constraints. Furthermore, environmental constraints might be accumulative.

Tactical decisions include maintenance scheduling, demand forecasting, production planning, and emissions forecasting and trading. Strategic decisions require evaluation of future investment, budgeting, and long-term contract negotiations with suppliers and customers. Monitoring requirements include tracking plan versus target versus actual performance such as process energy use, cost accounting based on real costs, emissions accounting, and performance monitoring of utilities equipment.

Given that the quality and perception of information have to be improved in order to be used effectively to change an organization’s culture, there are five key issues highlighted by the aforementioned model that must be addressed.

1. Capability to perform
2. Operations readiness

3. Organizational hierarchy of needs
4. Establishing “actionable” information to act as a catalyst of change
5. Measuring the right things

Technology can be the most effective tool to evolve plant culture. The speed of distributing reliable, timely, and easy-to-use performance and supply information changes the level of discussion and the dynamic behavior of the organization. However, the organization needs to have the right set of performance measures, a persistent effort to enhance operations readiness, an appropriate human resources strategy, and, above all, appropriate quality of information. The impact of these on changing a plant culture can be dramatic — as soon as the team realizes the ability and need to improve, they learn quickly how to evolve their work methods to achieve it.

Modern manufacturing strategies are viable, and information technology strategies can transform the plant culture to achieve manufacturing success. The evidence of a change in the level of dialog and team dynamics confirms the success, but several key strategies must be maintained consistently in order to sustain this success.

15.8 OPERATIONS STRATEGY

Major installations have been able to consolidate up to 10 control rooms and improve plant output by up to 15%. The goal was not to reduce personnel — the goal was to improve flexibility because it is easier for a small team to drive change than a large one. This has been proven consistently in remote jungle sites in southeast Asia, remote mountain sites in the united states. and Canada, and remote sites in northern Africa. There are several techniques necessary to make this feasible.

- Remote, secure, online computer access to all control and monitoring equipment.
- Advanced alarm management that dynamically filters alarms during upsets so that “alarm showers” are avoided that blur the visibility to key cause-and-effect alarms.
- Stabilization techniques to help units withstand upstream and downstream disturbances.

15.9 MODEL BASED ASSET MANAGEMENT

Facility simulation has developed quickly over the last several decades. The integrated oil and gas field with production facilities is a complicated system with a high degree of interaction and dynamics that make it impossible for the human mind to control and optimize both technical and business parameters. Integrated discipline workflow and the development of model-based asset management are necessary to deal with these complexities effectively.

Model-based asset management techniques will begin to play a major role in the integrated gas plant of the future in bringing the predictive power of the production engineering tool set to real-time data platforms. Process and production simulation models will move from the domain of the engineering expert to be used by managers, operators, business development, contracts, and finance. The analysis and prediction of near real time and future asset performance become a reality in the world of model-based asset management. For this the complete integrated asset will be modeled dynamically in real time for both slow rigorous and fast proxy loop modes.

Facility simulation has come of age. The ability to predict very complex facility simulations in both steady state and dynamic mode has been achieved by continued integration of thermodynamic methods, hydraulic simulations, and unit operations. Facility simulation has progressed such that dynamic start-up and shutdown simulations have become routine and trusted among the engineering community. One can begin to think of dynamic simulation as a "virtual plant" for operational, advanced process control and business analysis by a number of different departments.

Facilities simulation, control, optimization, operator training, collaborative engineering and planning disciplines, have a rather fragmented application of a number of associated technologies. To overcome this fragmentation, the first step is the adoption of integrated asset models that use comprehensive existing applications that are integrated together by a common "glue" layer allowing a full model of the entire operation.

New concepts are appearing from the software community in the form of workflow solutions that enable assets to be modeled from a suite of selected software adaptors and can bring data from disperse and third-party tools into one common environment, creating the integrated asset model. One can then apply engineering and business applications to identify asset-wide improvement opportunities.

A portal environment can be built that will allow multiple asset models to be visualized, compared, and analyzed with distributed data sharing and events management via proven and accepted enterprise platform message bus technologies. Key visualization and workflow technologies required include look-forward analytics, production scorecard, workflow management, production reporting, capital planning and scheduling that can be performed on a uniform, company wide, basis and allow for rapid, informed decision making.

15.10 OPTIMIZATION

Maximizing the profit contribution of natural gas processing facilities is challenging given the fluctuating economics, changing ambient conditions, and feed variations that processors must address. New dynamic markets for gas components lead to a need for stronger analytical capabilities of decision support tools. Also the supply situation becomes more variable as the gas companies respond to market opportunities. The result is a more rapidly changing environment and, as a consequence, processing plants need to be reconfigured more often. There is a need for better understanding on how to make plant-wide production plans and implement these through process management.

Decision support tools have to combine both optimization and simulation capabilities in order to analyze the consequence of different scenarios. Advanced modeling and optimization are needed in order to address the new following challenges in plant control and optimization systems, which have an impact on the optimal design of gas processing plants.

- Many of the processing plants are short of capacity. It is more important than before to maximize flow through the plants or the profit from final products.
- New opportunities exist because of recent advances in modern control technology, e.g., model predictive control lifts the level of automation and gives better opportunities for process optimization systems at higher levels.
- Integration of tasks and systems in operation support centers. E-fields give a new perspective on operation of the oil and gas production, and this mode of operation must be prepared for in the gas processing plants.

- The man technology organization perspective, which provides relevant information for the personnel involved.
- The use of models for planning and process control plays a central role.
- Modeling the dynamics of the markets and incorporate it in contingent plans. The plans should reflect capacities and production possibilities of the plant. This again depends on the design and operation of the plant control system.
- Structuring of the information flow between layers and systems in the process in the process control hierarchy, real-time optimization level, and planning and scheduling levels.
- Self-regulating and robust systems, which optimize the other parameters.
- Limitations imposed by process design.

15.10.1 Tools for Optimization

Many of the tools mentioned in Chapter 12 could become integral elements of the operations strategy for maximizing profitability. A steady-state process model is necessary to determine the capability and current performance of the operation. Ideally, a dynamic model would be available to train operators and investigate control options. These models can then be used in a real-time mode to report the capability of the plant under different conditions.

Predictive equipment models can be used to determine when maintenance will yield long-term benefits that more than offset the short-term costs. These models can be a extension of existing steady-state models. These steady-state models can also be key elements of an online optimization strategy.

Real-time control models can be constructed from perturbations of a high-fidelity dynamic model. It is necessary to validate these models constructed from models against actual plant data. Often the control models require detuning to provide adequate and robust control.

15.10.2 Optimization Alternatives

Some optimization alternatives for natural gas processing plants include the following.

- Advanced regulatory control
- Multivariable predictive control

- Neural network controllers
- Off-line process simulators
- Online sequential simulation
- Online equation-based optimization
- Linear programs
- Web-based optimization

Advanced regulatory control and multivariable predictive control are discussed in Chapter 12.

Neural network-based controllers are similar to multivariable controls except that they gather plant data from the DCS and use the data to "learn" the process. Neural network controllers are said to handle nonlinearities better than multivariable controllers and are less expensive to commission and maintain.

Neural network-based models are only valid within the range of data in which they were trained. Changes inside the process, such as a leaking JT valve, would be outside of the range in which the model was trained and therefore the results may be suspect.

Refining and chemical companies have attempted using neural networks for control many times since the mid-1990s. The technology has not proven to be viable compared to the other approaches, such as advanced regulatory control and multivariable control.

Off-line process simulators are used to develop a rigorous steady-state or dynamic model of the process. They are used by process engineering personnel to design and troubleshoot processes. Off-line simulations allow for what-if case studies to evaluate process enhancements and expansion opportunities.

Off-line simulators are typically not used to support daily operational decisions. They must be updated and calibrated to actual plant conditions for every use. They are not as robust as equation-based optimizers and can have difficulty converging large problems reliably and quickly.

A few of the off-line simulation companies offer an inexpensive, sequential-based optimization system. The optimizer is based on a rigorous steady-state model of the process and is typically less expensive than equation-based systems. They leverage the work done to develop the off-line model for online purposes. The extended convergence times inherent in these systems bring into question the robustness of the technology. The sequential nature of the solving technology also can limit the scope of the system. These systems require hardware and software to be purchased, installed, commissioned, and maintained onsite and require specialized resources to support them.

Equation-based optimizers use a rigorous steady-state model of the process as the basis for optimization and include an automatic calibration of the model with each optimization run. The equation-based solving technology allows optimizers to execute quickly and robustly, making them viable for larger-scale problems, i.e., multiplant load optimization for plants on a common gathering system.

Equation-based optimizers require a hardware platform, a costly software component, and highly specialized engineering services to install, commission, and maintain the technology. Closed loop implementation requires a multivariable controller to be installed to effectively achieve the optimal targets.

On-line equation based optimizers, when coupled with a multivariable controller, represent the standard in optimization technology for refining and petrochemical industries. Most refiners and petrochemical companies are deploying this technology rapidly to improve the profit contribution of their larger scale processing facilities. Unfortunately for gas processors, this technology is justifiable only for very large gas processing facilities and is not scalable across their asset base.

Linear programs are used for evaluating feed and supply chain options. Linear programs are an off-line tool that allows for what-if case studies and evaluation of supply chain alternatives. They are relatively inexpensive. Linear programs provide a linear representation of the plant process and do not provide guidance for operators (Roop *et al.,* 2002).

Web-based optimization has been applied to cost effectively supply equation-based optimization to the natural gas processing industry. However, the time lag in collecting data, calculating the optimum, and providing advice to the operators to implement the advice may not be fast enough to keep up with the constantly changing conditions experienced in natural gas processing plants.

Another option is on-line performance monitoring tools that predict the optimum operating point under all conditions. These tools usually provide dashboards and graphical indicators to show the operator the gap between current plant performance and optimum performance. They may or may not provide advice on what parameters to change to reach optimum performance.

15.11 INDUSTRIAL RELEVANCE

In the upcoming decade there will be large investments in gas production and in facilities for transportation and processing. Optimal utilization of all these facilities is vital in order to maximize the value from produced natural gas. Advanced process control and operation have raised the level of automation in the process industry significantly in the last decades. For example, in the refinery industry, methods such as model-based predictive control (MPC) and real-time optimization (RTO) have become widely used (Qin and Badgwell, 2003). The focus on these technologies has given large benefits to the industry in the form of increased throughput and more robust operation. The improvements obtained by the use of better control and decision support tools ends up directly on the bottom line for the operating companies. The substance of these tools is in fact software realization of process knowledge, control, and optimization methods. MPC and RTO are now more or less off the shelf products, although for complex processes the adoption of this technology requires specialist competence. However, there are significant potentials for further improvements in this area. One challenge for the gas processing plants is being able to quickly adapt the plant operation to dynamic changes in the markets, thus the plant flexibility and ability to perform rapid production changes becomes more important. It is also required to know accurately the plant production capability, both on very short term (today-tomorrow) and on weekly, monthly, and even longer horizons. This requires the use of advanced optimization tools and efficient process calculations. It is important to consider these issues related to dynamic operation also in the design of new processing plants and for modification projects. Investments in this type of project have typically a short payback time (Moen, 2004).

A combination of individual units into an integrated plant gives a large-scale control problem that is more than just the sum of the units. Cross-connections, bypasses, and recycling of streams give more flexibility, but at the same time, the operation becomes significantly more complicated, and it is almost impossible to utilize the full potential of a complex plant without computer-based decision support tools. Thus there is a need for the development of new decision support tools that combines optimization technology, realize process calculation models at a suitable level of speed and accuracy, and structure the information flow, both from the process measurements and deduced variables and from the support tool down to the manipulated variables in the control system. It is also needed to develop further the methodology related to plan-twide control in this context (Skogestad, 2004).

15.12 THE TECHNOLOGY INTEGRATION CHALLENGE

There are three likely scenarios of the deployment of the integrated gas plant.

1. *Business as usual*. Digital technology, information, data, and models are used in an incremental way to reduce costs, increase recovery, and improve production, but no fundamental changes are made in business models, competitive strategies, or structural relationships.
2. *Visionary*. Those who can best adopt and apply digital technologies, and concepts, will use them to gain significant competitive advantage. This will involve significant investment in software and information technologies along with culture and management change. The industry will need to demand leaders in the highly technical processes and modeling software contribute new and innovative solutions.
3. *Symbiotic relationships*. Those who use the availability of turnkey solutions to optimize production and leverage into larger industry positions. This will involve the use of third-party technology consultants and solution partners in an unprecedented way.

For most companies, an evolution approach that blends all three scenarios may well be the chosen route.

The foundation of the integrated gas plant of the future is engineering simulation, with integrated asset models and portfolio views of the business built on top. At the heart of the digital revolution in the upstream energy industry is a shift from historic, calendar-based, serial processes to real-time, parallel processes for finding, developing, and producing oil and gas assets. Real-time data streams, combined with breakthrough software applications and ever-faster computers, are allowing the creation of dynamic, fast-feedback models. These dynamic models, running in conjunction with remote sensors, intelligent wells, and automated production and facility controls, will allow operators to visualize, like never before, what is happening in the facility and predict accurately what needs to happen next to maximize production and efficiently manage field development.

15.13 SCIENTIFIC APPROACH

Optimal operation of gas processing plants is a challenging multidisciplinary task. Large-scale process optimization is challenging in itself. Thus, when we also consider dynamic conditions in the market and on the supply side the operation will most certainly run into problems that must be solved. Some will arise from the size of the problem, some from complex process behavior and from requirements to solution of complex optimization problems.

The starting point in this project is the need for decision support as seen from the personnel in a plant operating company. This defines a set of tasks that require optimization calculations, process calculations, and measurement data handling. The personnel in question can be plant operators, production planners, sales personnel, maintenance planners, process engineers, managements, etc. Experience from other applications such as gas transportation will be utilized (Rømo *et al*., 2003).

The inclusion of market factors, capacity planning, and scheduling shall be focused, as this sets new requirements to vertical integration of process control and the optimization layers. We may, in fact, have several such layers, where, for example, the classical real-time optimization is just one element. In planning and capacity assessment, the RTO layer may be accessed by superior layers in order to compute the optimal process targets over a certain horizon.

The requirements to process calculations for each type of task shall be classified. This may result in a set of optimization problems with different properties and requirements to solution approach and to the underlying process calculations and data handling, e.g., one approach from the planning side is to start with empty or extremely simple process models and to refine the models based on the requirements to the planning.

Segmentation into suitable process sections and control hierarchies are central issues. Here we can apply methods from plant-wide control in order to structure the control of the plant units in a way so that the influence from unknown disturbances and model uncertainties are minimized (Skogestad, 2004). A very important output in the first phase is to define high-level targets for the process control. The next important issue is to develop methods to select the variables that should be exchanged between the optimization and the process control layers. This is a control structure design task where the focus is on selecting the variables that are best suited for set point control in order to fulfill the process optimization targets in the presence of unknown disturbances and model parameters and measurement errors. Segmentation of the control into suitable sections and layers is also a part of this task.

Recent advances in process control technology also give a new perspective. For example, with an active MPC controller, information about active and inactive process constraints is high-level information that can be exchanged with the optimization layer instead of representing the constraint equations at the optimization layer.

Efficient use of models is such a wide area that this issue can be subject to extended research in separate programs. For example, in process design it is industrial practice to use quite detailed process models, including rigorous thermodynamics and representation of detailed phenomena within each process unit. In operator training simulators, detailed dynamic models are used, but these are rarely the same models as used in design, and the built-in process knowledge in the form of model configurations and parameters is usually not interchangeable because of different modeling approaches and different model data representation. The models used in MPCs are normally captured from experiments on the process itself and are not connected to the other two types of models. For real-time optimization, steady-state models are normally used, and in some cases model tools with rigorous models are also used there. For capacity assessment, correct representation of potential bottlenecks is important.

15.14 OTHER MISCELLANEOUS INITIATIVES

Maintenance management, field information handling, work process optimization, compensation design, and procurement initiatives are several of the current gas processing management initiatives.

In field information projects, companies considering an upgrade should understand that technical support must also be upgraded and that care should be taken to select systems with an eye toward ensuring the ongoing availability of support over a reasonable period.

In an effort to determine how they are doing against the competition, as well as discovering new areas for potential, some processors have become involved in industry benchmarking activities. Benchmarking tools with a reasonable level of analytical content provide benchmarking against a select peer group as well as individual analysis of various cost components. Most organizations that persevere in the benchmarking process and are diligent following up on findings testify that benchmarking is a useful tool.

A company that fully utilizes second-wave technologies to streamline its back office and process support technologies could reap a reduction in selling, general, and administrative (SG&A) costs in the range of 8 to 10%. For a typical, large firm, SG&A costs represent an estimated 10% of the total enterprise costs. Thus, a 10% reduction in SG&A outlays would reduce overall corporate spending by 1% a major gain, given that these savings would drop to the bottom line.

15.15 CONCLUSION

The goal of an integrated operations environment is to enable direct translation of management strategies into manufacturing performance. The vision is that

- Utilization of raw materials is optimal
- Overall margin and yield of product(s) are maximized
- Planning, operational, and monitoring cycles are fully integrated
- Identification and correction of problems occur rapidly
- Operational (short and long term) factors are fully understood
- The work force is well informed and aligned for a common purpose

The system envisioned is an integrated platform for computing and information processing at the production level. It is built around the premise that information is not to be isolated and that better information, when made widely available, will help people operate the facility closer to the optimum. A key principle is to empower everyone to maximize the value of his activities. The production management system provides the tools to help personnel do their job better.

A fully functional system will enable the quality cycle of planning, measuring, analyzing, correcting, and then planning again. To make improvements the staff must be able to see and measure progress.

The integrated production/management system provides data and the means to analyze situations, define solutions, and track progress. It is an integrated platform of computers, networks, and applications. It brings together the many individual automated systems that exist today and fills any gaps to bring the overall system to a high level of performance. The production/management system spans the gulf between process control and corporate business systems to support the day-to-day operations.

To achieve these goals, the production management system must

- Be a single, comprehensive source of real-time and business data addressing all operations and available to all appropriate personnel. This means providing long-term storage of all data (e.g., historical process, laboratory, plan, production, and shipment data), merging of these data, and retrieving data.
- Provide information retrieval tools for the full range of users. This usually means highly graphical tools that provide ease of use and a consistent look and feel to minimize the burden of finding and accessing information.
- Integrate a wide range of computer systems and applications. No one system or set of tools will provide all the functionality needed. Instead, the production management system should allow for the use of the best products from different vendors.
- Provide standard screens and reports that focus attention on problems and opportunities. The system should report by exception, highlighting the unusual, the exceptions, and the opportunities. It should compare actual results with the established plans and economic KPIs.
- Maximize information content while minimizing data volume. This is achieved by the use of performance indices and other numerical, measurable indicators, and the presentation of this information in graphical form whenever possible.
- Present operational data in economic terms whenever possible. Opportunities, problems, and deviations from an operating plan should be prioritized based on their impact on overall profitability and, where possible, include an indication as to the action or follow-up activity that alleviates the deviation.
- Provide analytical tools that enable users to explore and pursue their own ideas. Much of the value of integrated operations comes not from presenting data about current operations, but from people looking for ways to improve current operations.
- Facilitate plant-wide communications and work flow. For instance, plans and economic KPIs set in the planning group should flow automatically to operations to help operate the plant. Beyond the

manufacturing issues, implementing a project with the scale and complexity of an integrated production/management system creates several management of technology issues.

These create the need to

- Balance the selection of individual applications with the need to integrate applications across departmental boundaries.
- Provide a single, accessible, look and feel. This is particularly important for users accessing data that originate in systems that belong to other departments ("single pane of glass").
- Employ the latest proven information system technology as it becomes available and at the pace the operator can assimilate and manage.

Integration is a true example of the total being greater than the sum of the parts. A gas plant can profoundly affect the nature, quality, and profitability of its operations throughout the life of the gas plant with a truly integrated production management system.

Several operations have adopted some or all of the discussed strategies to improve their performance. A couple of these plants are discussed in Kennedy *et al*. (2002). Other examples include gas processing operations in Tunisia, Norway, Nigeria, and Indonesia.

The future seems to belong to those who will be able to mix vision, intelligence, and understanding of human nature, technology, and the processing business into a formula for success in the new world of natural gas gathering and processing!

REFERENCES

DeVries, S., Lennox, K., and Poe, W., "Maximize The Profitability of Your Gas Plant Assets: The Integrated Gas Plant of The Future," Paper presented at the GPA Europe Meeting, Amsterdam, The Netherlands (Sept. 27, 2001).

Kennedy, J.M., Saunders, A., and Poe, W.A., "Reducing Human Intervention in Gas Processing," Paper presented at the 81st GPA Annual Convention, Dallas, TX (March 12, 2002).

Mazlow, Abraham H., "Toward a Psychology of Being," Van Nostrand Reinhold (1968).

Vollman, T. E., Berry, W. L., and Whybark, C., “Manufacturing Planning and Control Systems,” 3rd Ed. Richard D. lrwin, Home-wood, IL (1998).

Moen, Ø., “Machine Means for Design, Control and Economic Optimization of Polyolefin Production Processes and their Products,” Keynote at Escape-14, Lisbon, Portugal (May 16–19, 2004).

Qin, S.J., and Badgwell, T.A., A survey of industrial model predictive control technology. *Contr. Eng. Pract* **11**, 733–764 (2003).

Rømo, F., Tomasgard, A., Røvang, L.B., and Pedersen, B., “Optimal Routing of Natural Gas in Pipeline Networks,” SINTEF Report, SINTEF Industrial Management, Trondheim, Norway (2003).

Skogestad, S., Control structure design for complete chemical plants. *Comput. Chem. Eng.* **28**, 219–234 (2004).

Howell, A., “Obtaining Value from Oil and Gas Model Based Asset Management,” Paper presented at the GPA Europe Spring Meeting, Dublin, Ireland (May 19–21, 2004).

Roop, M., Leger, J., and Hendon, S., “Alternatives To Optimize Gas Processing Operations,” paper Presented at the 81st GPA Annual Convention, Dallas, TX (March 11–13, 2002).

Neumann, R.W., “Managing Your Gas Processing Plant Fundamentalist, Fashionable, Farsighted or Fantastic!” Paper presented at the 78th GPA Annual Convention, Nashville, TN (March 1–3, 1999).

SUGGESTED READING

“Improving Operational Performance by Integrating Technology, Information and Culture.” ARC Insights# 2001-013AP (March 14, 2001).

Bullin, K.A., and Chipps, J., “Optimization of Natural Gas Gathering Systems and Gas Plants,” paper presented at the 84th GPA Annual Convention, San Antonio, TX (March 13–16, 2005).

CHAPTER 16

GAS PLANT PROJECT MANAGEMENT

16.1 INTRODUCTION

Project management is the application of knowledge, skills, tools, and techniques to project activities in order to meet or exceed stakeholder needs and expectations of a project. The project manager, sometimes referred to as the project coordinator or leader, coordinates project activities on a day-to-day basis. This is an ongoing challenge that requires an understanding of the broader contextual environment of the project and the ability to balance conflicting demands between (1) available resources and expectations (especially with respect to quality, time and cost); (2) differing stakeholder priorities; (3) identified needs and project scope; and (4) quality and quantity of the project's deliverables. Project management for engineering and construction projects requires the application of principles and techniques of project management from the feasibility study through design and construction to completion. Good project management during the early stages of project development greatly influences the achievement of quality, cost, and schedule.

This chapter covers many aspects of managing capital projects in the gas processing business. For the most part, best-practice management for gas plant projects follows generic project management principles applicable to most industrial engineering and construction projects. This chapter reviews many of the standard and accepted practices that lead to successful installations, as well as some of the unique considerations for gas plant projects, which arise from relatively complex processes employed in typically remote locations.

16.2 PROJECT MANAGEMENT OVERVIEW

One or more parties can perform the design and/or completion of a project. Regardless of the method that is used to handle a project, the management of a project generally follows these steps.

- **Step 1**: Project definition. Determine the conceptual configurations and components to meet the intended use.
- **Step 2**: Project scope. Identify the tasks that must be performed to fulfill the project definition. Also clarifies what the project does not include.
- **Step 3**: Project budgeting. Define the permissible budget plus contingencies to match the project definition and scope.
- **Step 4**: Project planning. Determine the strategy and tasks to accomplish the work.
- **Step 5**: Project scheduling. Formalizing the product of planning.
- **Step 6**: Project execution and tracking. Complete project tasks and measure work, time, and costs that are expended to ensure that the project is progressing as planned.
- **Step 7**: Project close-out. Final testing, inspection, and payment upon owner satisfaction.

Successful projects require effective management, which means (1) clear objectives, (2) a good project plan, (3) excellent communication, (4) a controlled scope, and (5) stakeholder support. Project management in today’s organizations demands multiskilled persons who can handle and manage far more than their predecessors and requires competencies that span all of the critical management fields.

16.3 INDUSTRY PERSPECTIVE

Gas plant project management includes planning, design, engineering, construction, and commissioning of the plant. Key elements are covered under the general headings of engineering, procurement, and construction (EPC). Today in many large organizations there has been a trend away from the owner company performing the whole project management function toward the delegation of EPC in part or whole to engineering consultant organizations. At the same time, smaller companies have almost always subcontracted EPC activities. Most companies, however, specify and procure the major equipment themselves (or at least oversee those activities) to assure technical compatibility and adequate lead times for delivery and so on. For example, the compressors and drivers will be specified and selected by the owner company as a priority.

A successful project in the gas processing industry is not only one that is profitable, but one that leads to the safe, reliable, predictable, stable, and environmentally friendly operational characteristics.

Gas processing is a service to the oil and gas production business. Oil and gas operations desire to produce into a system that has high availability and can produce a salable product in a safe, quality-consistent, and environmentally friendly manner. Gas processors must, of course, provide this service in a profitable manner. Flexibility to operate in various modes to respond to the markets and provide various processing alternatives should provide competitive advantages for a plant.

In many cases, the gas processing facilities are owned by the oil and gas producers as first facilities to enable the production of oil and gas and then as a value-added operation. In some cases, particularly the present state of the business in the United States, independent gas processors compete for gas that can be produced into a number of gas gathering systems. Several types of contracts exist, but the prevalent contract type is a “percent of proceeds” arrangement. Processors, who can offer the greatest revenue to the oil and gas producers, have an advantage in any case. The key is recovery of the greatest percentage of feedstock at the highest availability and at the lowest cost.

16.4 THE PROJECT MANAGEMENT PROCESS

At the onset of a project, the owner company will undertake the required economic and business analysis regarding new or expanded facilities in order to receive board approval and budget allocation. Based on this approval and funding, the project definition will be refined. The owner company will initiate the project and set out design objectives, usually embodied in the design basis memorandum (DBM), which will lay down the operating parameters and any key design guidelines and specifications. The owner company will solicit bids from EPC contractors and from these will select the successful bidder. In some cases, owner companies will have a partnership with an EPC contractor for certain types and sizes of projects. Pricing will be prenegotiated according to a range of possible contract models (fixed price, cost-plus, risk-sharing, etc.) or performed on a reimbursable basis. Both parties, owner and contractor, will set up teams to do the work. The EPC company may be asked to be completely responsible for all aspects of engineering, procurement, and construction or may only be required to do engineering and some procurement with the construction carried out by another company. Estimates and schedules will be set up by the EPC company in consultation with the owner. EPC companies can be large, such as Bechtel, or small, such as Cimarron Engineering or Tartan in Calgary. Each company may have somewhat different expertise and experience but tend to operate in very similar ways. Consultants and contract personnel will fill areas where the EPC company lacks resources or expertise. The owner also embeds staff in the contractor to ensure oversight of the management process and also may include specialist engineering personnel to monitor progress against plan, including quality assurance, as well as training of the owner staff on major projects.

For more complex, large projects that involve elements of innovation or requirements to build a gas facility in a region or country where such projects have not been conducted previously it is not unusual for the process to include feasibility studies and front-end engineering and design (FEED) studies conducted prior to awarding the EPC contract. The feasibility and FEED studies usually conducted by competent engineering consulting companies, capable themselves of conducting the EPC work. The deliverable from a FEED study may form the basis of a competitive tender for the EPC contract in which a number of prequalified contractors may compete.

A successful project requires the owner and EPC company to work together very closely. Typically, the owner will bring in operations staff at an early stage in the project to ensure that these staff members contribute to the project specifications and review deliverables as they unfold. In this way commissioning and operation proceed without serious problems. Companies with experienced staff often prefer to have the operating people closely integrated with the construction work from the outset. Similarly, with controls people, it is essential that the owner's control philosophy is conveyed to the engineers of the contractor and that the machinery suppliers are brought into the loop so that engineering specifications and machine products reflect these perspectives. The owner company will undertake the required economic and business analysis regarding new or expanded facilities to receive board approval and budget allocation. From this the project will be defined.

16.4.1 Defining Business and Project Objectives

The first step in the project management process is to align the business and project objectives. A project can be installed on time and on budget, but if it does not meet the defined business objects, then the project cannot be deemed a success. Some of the questions that the business owners must be asked by the project team include the following.

- How much gas is available for processing (ultimate reserves and daily deliverable quantities)?
- What is the market demand for gas and gas products that can be met by this project?
- What is a realistic gas production schedule?
- What are the production pressure, temperature, and composition of the gas?
- How will the gas pressure, temperature, and composition change over time?
- What products can be sold and at what price?
- What are the product specifications?
- How will the products be delivered to market?
- What are the local environmental policies?
- What are the local safety policies?
- What infrastructure, such as roads, bridges, loading and unloading facilities, and personnel housing are required?
- What is the skill level of available operations and maintenance personnel?

Because most gas processing plants are services to the oil and gas producers, collaboration between reservoir and production engineers and gas marketers to obtain answers to these questions is imperative. In cases where processors compete, then collaboration with those responsible for obtaining the processing and sales contracts (e.g., economists, lawyers, and negotiators) can be critical.

16.4.1.1 The Project Charter

The project charter is a document that demonstrates management support for the project. In particular, it authorizes the project manager to lead the project and allocate resources as required. It simply states the name and purpose of the project, the project manager's name, and a statement of support by management. Senior managers of the responsible organization and the partner organizations sign it. The project charter should be distributed widely to anyone with an interest in the project. This will help build momentum, encourage questions and concerns early in the project's evolution, reinforce the project manager's authority, and possibly draw other interested and valuable stakeholders into the project.

The project owner may be a joint venture of oil and gas companies, with one designated as project operator. The project charter is then usually signed off by all joint venture partners together with an authority for expenditure (AFE) approving the project budget and/or initial stages of expenditure.

16.4.1.2 Project Team Roles and Responsibilities

Project team size and makeup are dependent on the complexity of the project; however, the basic composition of the project team and their responsibilities is recommended for all projects.

1. **Project manager.** The project manager is responsible for project development, developing schedule, budget, and deliverables definitions; evaluation of alternatives; determining return on investment; adherence to company policies; obtaining funding; acquiring internal and external project resources; contractor selection; maintaining project schedules and budgets; evaluating quality of project deliverables as they evolve; identifying and mitigating downside risks; identifying and exploiting upside opportunities; reporting to business owners; and creating project close-out reports.
2. **Business owner representative.** The business owner representative is responsible for assuring that the project adheres to business objectives as objectives may change or require alteration during the project.
3. **Plant manager.** A plant manager should be appointed as early as possible to address operability and maintainability issues.
4. **Project engineer/construction engineer/start-up engineer.** A project/construction/start-up engineer can be one role on smaller projects and multiple roles in larger projects. This engineer (or engineers) is responsible for technical specifications for contract bidding purposes, technical evaluation of contractor bids, owner's representative during construction, management of construction inspectors,

turnover of facility to operations, training of operators, determination of plant performance, and identification of any project deficiencies.

5. **Purchasing representative.** The purchasing representative is responsible for the commercial evaluation of contractor bids and negotiation of contract.
6. **Process engineer.** A process engineer is recommended for evaluating alternative processing schemes during project development, assistance with technical specifications and evaluation of contractor bids, and assistance with operator training and start-up issues.
7. **Environmental engineer.** An environmental engineer is recommended to review and provide advice on environmental issues encountered during the project, including technical specifications and obtaining environmental permits.
8. **Safety engineer.** A safety engineer is recommended to review and advise on safety issues encountered during the project, including technical specifications and participating on hazard analysis evaluations.
9. **Production or reservoir engineer.** A production or reservoir engineer is recommended to be available to evaluate any oil and gas production issues that may be encountered during the project.
10. **Facilities planner.** For larger projects, a facilities planner should be available to assist with project economics and to serve as a liaison for economic premises and marketing issues.

16.4.2 Contracting Strategy

There are several alternative contracting stages and strategies. The first stage of contracting may be a front-end engineering design. With this approach, a contract is entered based on the design objectives for an engineering contractor to evaluate process and construction alternatives, as well as develop technical specifications for the project. In some cases, the owner's engineers may accomplish the front-end engineering design tasks. The second stage of contracting is for engineering, procurement, and construction services. Either stage can be contracted as a lump sum, fixed fee price also known as a turnkey project, or on a reimbursable basis also known as a time and expense contract. In some cases more complex contracts, such as risk sharing or gain sharing, will divide risks and rewards more evenly between contractors and project owners.

16.4.3 Conceptual Estimates and Schedules

Most operating companies have developed estimating tools for budgeting of plants similar to those they are currently operating. Many operating companies have the capabilities and resources to evaluate alternative process and mechanical designs with budgetary or conceptual level estimates. These estimates typically have an accuracy of ±30–40%.

Under other circumstances, such as when proprietary processes are in use, unique locations are to be selected, or there is a lack of available resources, an engineering firm may be hired to evaluate alternatives and determine budgetary estimates. After evaluation and selection of a conceptual process and mechanical design, the operating or engineering company will undertake a front-end engineering design. The detailed specifications and request for proposals will be the deliverable from the front-end engineering design.

Conceptual estimates and schedules should take the following into consideration.

- Location
- Operators and operability
- Constructability
- Special materials

The availability of fresh water and electricity are considerations in determining location. Port facilities, roadways, and waterways are another consideration. A qualified and available work force is always a consideration when determining location. In some locations, qualified operating personnel are difficult to find so inexperienced and poorly educated operators may be hired. In order to overcome their lack of qualifications, intensive training is required. Generally, it is good to include in the project training using high-fidelity simulators, particularly where inexperienced operators are to be hired. In addition, plants with novel processes with which even experienced and highly educated operators are not familiar should include additional training provisions. Such training will impact the project's cost and schedule.

Regardless of schedule, the project team's capability to construct the facility must be addressed. For instance, vessels of large diameter and height will require shop facilities that have appropriate size capacity, as well as trucking and rail facilities that accommodate the finished products.

In some cases, the vessel may require field fabrication or multiple vessels will be needed if shop fabricated. Alternatives for prime mover drivers, such as electric motors, steam turbines, gas turbines, and gas engines, may be influenced by the availability of infrastructure to support these devices. If electrical service is not provided by a utility, then generation or cogeneration facilities may be required. These must be addressed in the project cost estimates and schedules.

Special materials are often required in gas processing plant construction due to components such as hydrogen sulfide, carbon dioxide, mercury, and water. The availability of these materials and their delivery should be considered. Sometimes cladding or linings may be alternatives to expensive and scarce alloys. In addition, approved welding procedures may not be available or the work force may not have the expertise to perform certain procedures. Addressing such obstacles must be part of the project plans.

During the proposal solicitation and award of the construction contract, the cost estimates and project schedule for the prime contractor will be focused on construction activities and therefore fairly detailed and inclusive. However, the overall project schedule from an operating company point of view must consider nonconstruction activities such as permits, licenses, and other government requirements, staffing, accounting, and other internal issues as well as contracts with suppliers and customers to name a few.

As project definition improves so the uncertainties associated with cost estimates should decrease to a funding level of accuracy of approximately +15%/−10% with a 10% contingency identified. A probabilistic approach to cost estimating identifying percentiles (P90, P50, and P10) is also widely used to illustrate cost uncertainties (McIntire, 2001).

16.4.3.1 Hazards and Operability (HAZOP) Analysis

A HAZOP analysis or equivalent is good practice even when not a statutory requirement. Such an analysis will most likely recommend the addition or deletion of valves, lines, instrumentation, and equipment needed for safe and reliable operation.

Figure 16-1 illustrates a stages and gates approach to oil and gas facilities project management that emphasizes the importance of the planning stages (feasibility, FEED) leading into EPC contracting, construction, and fabrication activity onto HAZOP and ultimately plant commissioning. To move from one stage to another requires a gate to be passed where

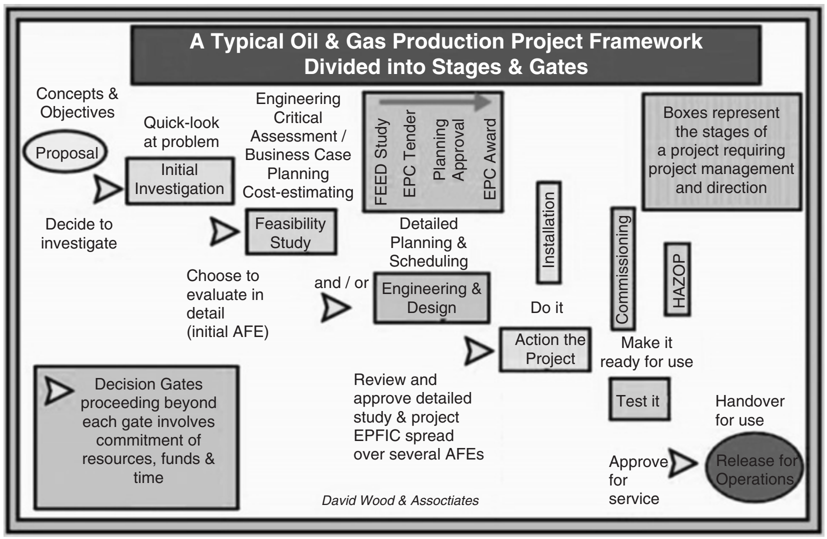

Figure 16-1. Stages and gates approach to project management (Wood, 2006).

decisions and approvals have to be made associated with funding, technical design, and project priority. Such approvals are usually structured in the form of AFEs to be signed off by the project owners (and often other stakeholders, e.g., government authorities) as positive approval to proceed under stages of a project budget. Although the diagram for simplicity suggests a linear process proceeding from one stage to another, in practice there are often loops and feedbacks to the work of earlier stages that require adjustments to design and so on.

16.4.3.2 Scheduling and Cost Estimating Software

Software exists to assist with both conceptual cost and time estimates, as well as detailed estimates and complex project networks involving the optimization of project networks with critical path analysis. Most major engineering, procurement, and construction contractors have their own custom tools. Smaller contractors and operating companies may use products supplied by a vendor specializing in these tools. For larger projects, it is increasingly common for Monte Carlo simulation analysis to be used in conjunction with critical path identification to yield probabilistic estimates of cost and time associated with each project activity and for the project as a whole (Wood, 2001).

16.4.4 Project Execution Planning

A project must be planned and tracked against the plan to assure successful execution. A project plan sets the ground rules and states them in a clear fashion. The project plan helps control and measure progress and helps deal with any changes that may occur. Previous experience is the best guide for determining the necessary tasks and the time to complete them. Many engineering and operating companies maintain databases that include previous project plans with actual time to completion and costs. To be able to benefit from such an approach requires good quality record keeping both during a project and following its completion.

Although no two projects are identical no matter how similar they appear, these databases of past experience contain valuable information on which to plan. It is necessary to understand any unique requirements that previous projects met and how the current project compares. Some dissimilarities may include the following.

- location, which impacts the government regulatory bodies, remoteness, cost of labor, etc.,
- make-up of the project team, including expertise, time available, geographic dispersion of team members, cultural differences, organizational affiliations/loyalties,
- project scope, and
- current economic conditions, which affect inflation and employee availability.

The project plan should be relevant, understandable, and complete and reflect the size and complexity of the unique project. The project plan should include the following elements (Hauge and Cramer, 2002).

- A project charter
- A project time line
- A responsibility matrix
- A project plan budget
- Major milestones with target dates
- A risk management strategy

16.4.5 Pre Project Planning Measurement

The project objectives, or the measure for project success or failure, are often defined in terms of cost, schedule, and technical performance. In order to serve as a baseline for project execution, measurements should be in place to identify target completion dates, budgets, and expected technical performance. These measures should be included in a system that allows tracking of actual, target, and projected dates and costs with variances highlighted. The technical expectations should be tracked as well and checked for compliance as the project proceeds.

16.4.6 The Responsibility Matrix

Projects are a collaborative effort between a number of individuals and organizations working together toward a common goal. Managing a diverse team, often spread over several locations (and countries), can present some special challenges. A responsibility matrix is a valuable project management tool to help meet these challenges. The matrix ensures that someone accepts responsibility for each major project activity. It also encourages accountability. The responsibility matrix should correspond with the project time line. An example is shown in Table 16-1.

The left-hand column in Table 16-1 lists all the required tasks for the project while the team members (e.g., project manager, project engineer, safety engineer, plant manager, purchasing agent) are listed across the

Table 16-1
Typical Responsibility Matrix[a]

Task	Project team members				
	Contractor	Owner/operator	1	2	3
1	A	S	P	—	—
2	—	A	S	P	I
3	—	A	S	P	—
4	A	S	P	—	—
5	A	S	P	I	I
6	A	S	P	—	I
7	—	A	S	—	P

[a] S, sign off; A, accountable; P, primary responsible; I, input.

top. A code is entered in each cell that represents that team member's involvement in the task in that row. For example, choose codes appropriate to the project; the key is to clearly identify who has a role in every activity, who is accountable, and who must sign off. Make sure the matrix is included in the project plan so that every participant is clearly aware of his or her responsibilities.

16.5 PROJECT CONTROLS

The two main elements of a project plan are the time line and cost control. A time line assures that the project is scheduled properly to meet anticipated and promised dates. For projects that involve several parallel or overlapping activities, the time line becomes a network. Cost control assures that the project meets its budget.

16.5.1 Project Time Line

By dividing a project into the individual tasks required for completion, the project time line (Greer, 2002):

- Provides a detailed view of the project's scope
- Allows monitoring of what has been completed and what remains to be done
- Allows tracking of labor, time, and costs for each task
- Allows assigning of responsibility for specific tasks to team members
- Allows team members to understand how they fit into the "big picture"

This time line can take on a variety of formats and philosophies. The most prevalent philosophy is to determine the date for project completion and work backward to identify key dates when certain milestones require completion. In practice, the project completion date is first determined through iterations of what is possible going forward with contingencies and identifying an end date. This usually involves establishing a critical path of activities, which must be completed on time for a project not to fall behind its scheduled end date. If this end date is not acceptable, then acceleration plans should be explored. Most methods of acceleration require additional expense to accomplish objectives. These methods may include parallel tasking, overtime, and contractor bonuses for meeting an aggressive date to name a few.

Once the project milestones are set, then subtasks and assignments are identified. On large projects, certain tasks and subtasks will be assigned to an assistant project manager. The overall project manager will become responsible for coordinating the activities of the assistant project managers. The milestones, major tasks, subtasks are commonly shown in a Gantt chart. This type of presentation presents dependencies in a graphic display such as predecessors (tasks that must be completed prior to commencing the next task) and successors (tasks that cannot be started until a previous task is completed). Milestones are significant events in a project, usually the completion of a major deliverable or tied to vendor or contractor payments. A very good method to analyze subtasks is to consult with an expert on accomplishing the subject task, and the very best way is to receive an estimate with commitment from the person responsible for executing the task. This is where contractors and their bids contribute to the best task analysis possible.

Note that as the project progresses, there may be tasks that were not foreseeable in the original plan or tasks that are added to enhance the overall project outcome. The impact of additional tasks, delays, or even acceleration requires consideration of the impact on both the time schedule and resources. When changes to the schedule are warranted and feasible, the project manager should get a written agreement for the revised plan from all the key stakeholders in the project. A regular update of the time schedule is recommended as part of a routine project status report. The time period for such reporting may be weekly or monthly depending on the project or the stage of the project.

16.5.2 Risk Management

Risk is inherent in all projects. In project management terms, “risk” refers to an uncertain event or condition that has a cause and, that if it occurs, has a positive or negative effect on a project’s objectives and a consequence on project cost, schedule, or quality. As discussed in previous sections, the measure for project success or failure is defined in terms of cost, schedule, and technical performance. Project risk management is intended to increase the likelihood of attaining these project objectives by providing a systematic approach for analyzing, controlling, and documenting identified threats and opportunities during both the planning and the execution of a project. The application of project risk management will vary from the operator (owner) or the contractor side. The term risk management is used to lump together different activities. These activities may be divided into (Hauge and Cramer, 2002) the following.

1. Activities related to the day-to-day identification, assessment, and control of uncertainties, i.e., risk management activities related to understanding and controlling the most important risks threatening the achievement of well-defined project objectives. This type of risk management may be based on a qualitative approach where each risk is assessed separately.
2. Activities related to the periodic assessment of achieving project objectives, i.e., assessing the probability of achieving well-defined project objectives with respect to schedule, budget, or performance. The periodic assessment must use a quantitative approach based on the aggregation of most critical uncertainties.
3. Activities related to the ranking of a set of alternative decision options/system solutions, i.e., ranking different alternatives with respect to their desirability measured in terms of the corresponding project objectives. Such ranking is typically performed at major decision gates during the conceptual design stage.

Risk management is a key to success for project execution, but is often constrained by inadequate work processes and tools. An overall understanding of the different risk factors and how these affect the defined project performance goals is critical for successful project management and decision making. Project risk management is a systematic approach for analyzing and managing threats and opportunities associated with a specific project and will increase the likelihood of attaining project objectives. The usage of project risk management will also enhance the understanding of major risk drivers and how these affect the project objectives. Through this insight, the decision makers can develop suitable risk strategies and action plans to manage and mitigate potential project threats and exploit potential project opportunities. Project risk management is based on a number of different analysis techniques. The choice between these techniques is dependent on the quality of the information available and what kind of decisions project risk management should support. Day-to-day usage of project risk management is typically based on using risk matrices, accounting for both threats and opportunities. With sufficient uncertainty information, the project risk management analysis can be extended to provide more direct decision support through probabilistic cost–benefit analyses.

Gas processing projects are often characterized by large investments, tight time schedules, and the introduction of technology or construction practices under unproven conditions. These challenges can result in a high-risk exposure but also opportunities that bring great rewards.

16.5.2.1 Project Risk Management Methodology

It is important to perform risk management in a structured manner. Indeed transparent risk management frameworks are becoming statutory requirements for many companies, obliging them to demonstrate how they are managing risks throughout their organizations [enterprise risk management (ERM)] or on an enterprise-wide basis (e.g., the COSO framework in the United States). It is important to ensure that the project risk management methodology is consistent with ERM frameworks (Wood and Randall, 2005). The project risk management process is often broken down into the following five general steps (Hauge and Cramer, 2002).

1. Initiation and focusing: initiate risk management process, including identify project objectives. The initiation should also assign personnel to the main risk management roles such as risk manager.
2. Uncertainty Identification: identify risks affecting the project objectives. Assign responsibility for assessing and mitigating each risk.
3. Risk analysis: assess for each risk the probability of occurring and the corresponding objective consequences given that the risk occurs. Based on the risk assessment, classify each risk in terms of criticality.
4. Action planning: identify risk-mitigating actions so that the most critical risks are mitigated. Assign responsibility and due dates for each action.
5. Monitoring and control: review and, if necessary, update risk assessments and corresponding action plans once new and relevant information becomes available.

The "initiation and focusing" step is normally performed once at the start of the project, whereas the four other steps are performed in an iterative manner. The initiation of project risk management in projects has the following set of goals (Hauge and Cramer, 2002).

- Identify, assess, and control risks that threaten the achievement of the defined project objectives, such as schedule, cost targets, and performance of project delivery. These risk management activities should support the day-to-day management of the project, as well as contribute to efficient decision making at important decision points.
- Develop and implement a framework, processes, and procedures that ensure the initiation and execution of risk management activities throughout the project.
- Adapt the framework, processes, and procedures so that the interaction with other project processes flows in a seamless and logical manner.

The project risk management process should be assisted by a set of tools that supports these processes.

16.5.2.2 Risk Response Planning

A risk response plan can help maximize the probability and consequences of positive events and minimize the probability and consequences of events adverse to the project objectives. It identifies the risks that might affect the project, determines their effect on the project, and includes responses to each risk. The first step in creating a risk response plan is to identify risks that might affect the project. The project team members should collaborate referring to the project charter, project time line, and budget to identify potential risks. Those involved in the project can often identify risks on the basis of experience. Common sources of risk include the following.

- Technical risks, such as unproven technology
- Project management risks, such as a poor allocation of time or resources
- Organizational risks, such as resource conflicts with other activities
- External risks, such as changing priorities in partner or contractor organizations
- Construction risks, such as labor shortages or stoppages and weather

16.5.2.3 Developing Risk Response Strategies

There is no preparation for mitigating all possible risks, but risks with high probability and high impact are likely to merit immediate action. The effectiveness of planning determines whether risk increases or decreases for the project's objectives. Several risk response strategies are available (HRDC, 2003).

- Avoidance — changing the project plan to eliminate the risk or protect the objectives from its impact. An example of avoidance is using a familiar technology instead of an innovative one.
- Transference — shifting the management and consequence of the risk to a third party. Risk transfer almost always involves payment of a premium to the party taking on the risk. An example of transference is using a fixed price contract.
- Mitigation — reducing the probability and/or consequences of an adverse risk event to an acceptable threshold. Taking early action is more effective than trying to repair the consequences after it has occurred. An example of mitigation is seeking additional project partners to increase the financial resources of the project.
- Acceptance — deciding not to change the project plan to deal with a risk. Passive acceptance requires no action. Active acceptance may include developing contingency plans for action should the risk occur. An example of active acceptance is creating a list of alternative vendors that can be supply materials with little notice.

Because not all risks will be evident at the outset of the project, periodic risk reviews should be scheduled at project team meetings. It is also important not to view risk with a negative mind set. In addition to the downside consequences associated with many risks lie opportunities, which should be identified and strategies developed to exploit them where possible. Risks that do occur should be documented, along with their response strategies in a risk register that assigns responsibilities for specific risks.

16.5.2.4 Qualitative Project Risk Management

The routine day-to-day identification, assessment, and control of project risks are similar to hazard and operability identification techniques.

The identification of risk consists of collecting and examining information on potential events that may influence the achievement of the project objectives. Each such event is categorized as a risk or an opportunity. The identification of these events should involve expertise from all main project competencies to reduce the possibility of important risks being overlooked. These risks will normally be prioritized so that only the most likely and consequential risks will be entered into a formal risk management process. The prioritization of risk should only be performed after thorough assessments and discussions among the project team. New information could mean that risks that have been determined previously as a lower likelihood must be inserted into the risk management process.

The assessment of each risk or opportunity is made in terms of scores for probability of occurrence and for consequence, given that it occurs, for each project objective. Based on the probability and consequence scores, the criticality of each risk with respect to achieving the project objectives can be assessed. Typically, a classification consisting of several possible probability scores and several possible consequence scores leads to several different classes of criticality, e.g., "critical," "significant," and "negligible." Figure 16-2 is an example of how risk can be classified in relation to probability, project cost (economy), project effort or duration (time), and project performance.

The control of each risk event is normally based on its risk classification. A risk that is classified as "critical" will normally result in actions being identified in order to reduce the risk classification to either "significant" or "negligible." The risk reduction can be caused by either preventive measures (reducing the probability that the event will occur) or corrective measures (lessening the consequences of the event) or both.

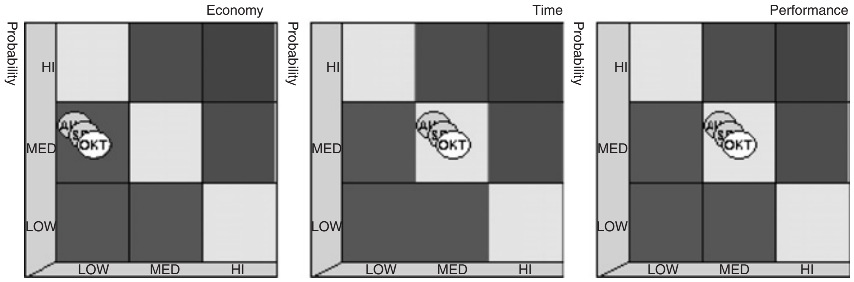

Figure 16-2. Risk matrices (Hauge and Cramer, 2002).

16.5.2.5 Quantitative Project Risk Management Assessment

A periodic assessment of the probability of meeting project objectives must be based on quantitative calculation of the aggregated effect of the most important risks on the project objectives. The aggregation must also take into account sequences of scenarios and risk events, as well as the project structures given by the budget, schedule, and operability. Several methodologies exist with supporting tools that can be used for risk assessment of the total budgets and schedules. The challenge is to apply a methodology and to find a tool that supports this methodology so that the integration of risks in the different domains can take place, their mutual independence can be represented, and their aggregate effect on the project objectives can be assessed. Fortunately, the usage of influence diagrams enables such a methodology and several influence diagram tools exist.

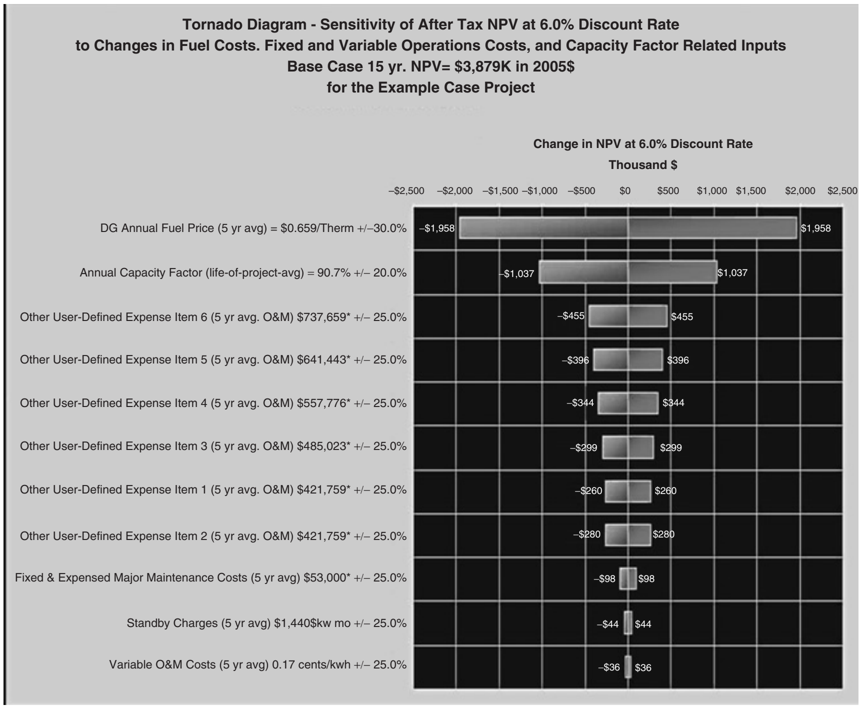

Figure 16-3. Typical tornado diagram (generated with Palisade risk software).

In an influence diagram, each risk is represented as a symbol (or "node") in a graphic diagram. The diagram represents the structural relationship between the different risks and their aggregate effect on the project objectives. In this manner, influence diagrams are well suited to represent risk scenarios. The mathematics used for assessing the aggregate effect is hidden away "behind" the diagram. In this way the influence diagram also represents a methodology to split a risk management model into two: (1) the structural relationship between the various risks and (2) the mathematics of the risk model, such as probabilistic distribution functions for the risks. Other quantification methods to identify criticality and rank risks include Monte Carlo simulations and presentation through "tornado diagrams." Tornado diagrams (an example shown in Figure 16-3) help identify which input parameters, if they were to change, would have the biggest consequential impacts on the analysis. They help to establish materiality of potential outcomes and to illustrate how the project will be impacted by changes, in order or significance, of those selected inputs. This provides valuable insight into which parameters might warrant further investigation to determine how changes would impact the objectives.

16.5.2.6 Risk Process Modeling

The model of the general risk management process is by no means complete. The two most important omissions are (Hauge and Cramer, 2002) as follow.

- No direct representation of the interaction with external organizations and their processes.
- No direct representation of the interaction with other internal processes.

Interaction with other processes should be designed so that the (new) risk management process is integrated as seamlessly as possible into the existing organizations and its already existing other processes (e.g., ERM framework). Any required modifications should be as small as possible. Hence, this risk management process is integrated into the organization by having the already existing weekly management meeting also assessing the weekly risk report. Hence, no new forum for management review of risk is established, only an additional item is added to an existing agenda.

Because different parties in the risk management process have different interest, these interests should correspond to different views into the risk management process. Risk mitigation (and/or exploitation) strategies themselves involve costs to instigate and may lead to secondary risks. It is important to reconcile such costs and secondary risks with the risk management objectives, materiality of the risks being addressed, the project budget, and resources available.

16.5.2.7 Project Risk Management in Interaction with other Management Processes

It is important to note that risk management has major similarities with other common management processes. Examples of such processes are (HRDC, 2003) as follow.

- Management of project changes
- Management of public permits
- Management of health, safety, and environmental issues
- Management of decision gates

Similarities exist in the identification of items, the assessment of their criticality, the identification of corresponding mitigating actions, and the follow-up of criticality assessment and corresponding mitigation plans. These similarities should be exploited when setting up the risk management processes and establishing the tools for the management of these processes.

The risk management process will remain constant over the different project phases throughout the life cycle of a construction project. The different risk management techniques for assessing day-to-day risks, calculating the ability to meet defined project objectives, and ranking different decision alternatives will also remain the same. In the ranking of different investment opportunities for a gas processing project, a number of issues will in general need to be looked into.

- Revenue
- Costs (costs of different activities: capital expenditures, operational costs)
- Schedule (of project tasks and completion of milestones)
- Taxes and depreciation

- Health, safety, and environment (meeting regulations and company requirements)
- Structural reliability (design that meets requirements)
- Onstream factor (design that meets availability requirements)

What will vary, however, over the life cycle is the quality of the available risk related information, the kind of competence that is needed to compile and prioritize this information, and the kind of decisions that are supported by the risk management activities. A believable risk management process must be conducted by personnel with domain knowledge of the project phase in which decisions are to be made. Because the required competence will vary with the project phase, it is unlikely that the same person can fill the risk manager role throughout the project. Whether one or more risk managers or a team is involved in managing risks it is crucial that analysis, actions, and outcomes are documented in a risk register and widely communicated both within the project team and to project stakeholders.

16.5.2.8 Other Risk Mitigation Concepts

Other methods of risk mitigation include (Hauge and Cramer, 2002) the following.

- Cost overrun protection. Cost overrun insurance can be purchased. This insurance is used most often for infrastructure such as bridges and islands that may be required to access or locate the facility.
- Regulatory risk. A wide-ranging political or regulatory risks form of insurance is available. This is designed to provide an indemnity in the event that any changes occur to the regulatory requirements or political stance perceived at conception of the project during construction and into commercial operations.
- Revenue stream stabilization. In addition to the historic insurance market places, capital markets are available that enable industry to transfer risks to financial vehicles and through structures different from traditional insurance policies. This market convergence has produced a wide range of creatively devised financial products, which can be applied to uninsurable risks. Hedging the sale prices for specific volumes of products is now widely used, particularly in the initial years of plant production prior to project payback to help reduce project finance risks.

- Blended risk solution. These are integrated risk solutions, which result in a more comprehensive package of protection than the traditional set of policies providing limited but specific coverage addressing different areas of risk.

16.6 QUALITY ASSURANCE

An important part of defining the result and performance of the project is the specification of its quality-related features, which the project must then aim to deliver. Quality assurance has been an issue at the forefront of organizational concerns for decades. The development of quality-conscious construction practices has been identified as being of the utmost importance in gaining and retaining a competitive edge. In the context of a project that aims to deliver a complex result, the quality aspects of that result will need to be planned, designed, aspired, and monitored. Quality assurance is a term used to incorporate the quality policy, quality management, and quality control functions, which combine to assure that the end result will be consistently achieved to the required condition. Its aim is to attain and assure quality through the adoption of a cost-effective quality control system and through external inspections and audits. Quality planning is an integral part of the planning activity. It manifests itself in the descriptions and in the scheduling of quality-related activities. Results of the quality planning activities are reflected in the resource and technical plans at each level of the project. Quality control is concerned with ensuring that the required qualities are built into all of the tasks throughout their development life cycles. Quality control utilizes measurable quality criteria and is exercised via quality reviews, by project reviews, and by the testing of products. Quality assurance requires agreement on the level of quality controls to be adopted, both specifically relating to the project and to the overall organizational policy. It is important that all three interests represented by the project owner are taken into account when deciding the mechanisms to be adopted.

The task descriptions should describe the purpose, form, and components of a task. It should also list, or refer to, the quality criteria applicable to that task. Task descriptions should be created as part of the planning process to shadow the identification of the tasks that are required by the project. Each task description may either apply to a specific item or to all the tasks of a given type. The component tasks of a complex task may be described in separate descriptions, giving rise to a hierarchy of task descriptions for that task.

Quality criteria should be used to define the characteristics of a task in terms that are quantifiable, and therefore allow it to be measured at various points in its development life cycle, if required. The criteria effectively define quality in the context of the product and are used as a benchmark against which to measure the finished task. Quality criteria should be established by considering what the important characteristics of a result or task are in satisfying the need that it addresses, and they should always be stated objectively. Subjective or descriptive criteria such as “quick response” or “maintainable” are unsatisfactory as they do not permit meaningful measurement.

Quality planning should ensure that all quality-related activities are planned and incorporated into the project schedule. The tasks required to ensure the quality of the delivered result are often overlooked, with the result that the project schedule fails to represent quality-related work. This can have serious consequences for the quality levels achieved, the overall budget, or both.

Quality control is concerned with ensuring that the required qualities are built into all of the tasks throughout the construction cycle. It defines the method of inspection, in-process inspection, and final inspection to determine if the result has met its quality specification. Quality control utilizes measurable quality criteria and is exercised via change control, quality reviews, project reviews, and by the testing of products.

Reviews should be scheduled prior to key decision dates and important milestones such as shipment of rotating equipment and major equipment. For instance, modifications to a gas turbine are made more easily in the shop prior to shipment rather than in the field after transit. Modifications required that could have been found with shop inspections can cause significant delays.

Many of the top engineering consulting companies operate integrated quality, health, safety, and environmental (QHSE) management systems, which they apply generically across their operations. This is appropriate as it makes clear to their staff that all four of its components are important to ensure successful outcomes and that QHSE all influence each other as well as the budgetary, schedule, and risk issues that drive project decisions. The adage that one should expect what you inspect applies to gas processing project management. Management and technical peer reviews are good practice. Inspection reports monitor not only quality but progress and should be used liberally for best results.

16.7 COMMISSIONING AND START-UP

Commissioning and start-up of a new facility or unit are very important phases of any processing project. These activities can be very expensive from a standpoint of project costs as well as deferral of operating revenues if delays are encountered. For this reason, it is good practice to assign a start-up engineer to plan and coordinate these activities. Operators and maintenance personnel should be hired well before the start-up date and trained on the equipment and process basics. Familiarity with the particular control system can be accomplished through the use of operator training simulators. These simulators can pay off easily by reducing start-up time. More detailed training on particular aspects of the equipment and process can be given to select individuals. Indeed, many now argue that it is important to have operations and maintenance personnel involved in the design and engineering phase of a project as it is much easier to sort out operational control logistics at the planning stage than later on.

Thorough checkout of the equipment should be conducted by the owner and operator of the facility. Punch lists of all deficiencies should be prepared, reviewed with the constructor, and updated on a daily basis. Purging of the equipment is quite important for safe start-up. Process gas or an inert gas, such as nitrogen, can be introduced to remove pockets of oxygen, which could lead to explosions. All high point vents should be opened and checked with a portable oxygen analyzer. These vents should only be closed when oxygen is no longer detected. Excess water and other liquids should also be removed from the system by opening low point drains until purge gas escapes through the drains.

Start-up should be a joint effort among the constructor, operator, and process designers. The constructor should be advised of any deficiencies in the equipment and instrumentation as they are identified by operations and have personnel readily available to resolve the issues. It is not uncommon for construction contracts to include a retention portion of the plant cost that the owner will withhold for a period (e.g., 6 months or 1 year) following start-up to motivate the contractor to deal promptly with any teething problems.

16.8 OPERATE AND EVALUATE

The final phase of a processing plant project is continuous operation and evaluation of the project results, including plant capability over a finite period of time. Before the plant is turned over to operation, the performance of the plant should be measured. Sometimes this is required by the contractor or process licensor to meet performance guarantees. The capability of each process and equipment should be calculated and become the baseline for operation. At this point you may discard the design capability as this was used only as a basis for sizing equipment. Many times equipment is sized with a contingency giving better than design capability. Other times, the equipment may not be capable of design. For example, a difference in the actual inlet gas composition from the design basis is a common culprit.

After detailed measurement of the plant's capabilities and deficiencies, these items should be well documented. The capability information is valuable should the plant require expansion in the future, and deficiencies become the basis for possible debottleneck and retrofit projects.

16.9 PROJECT CLOSEOUT

In addition to the evaluation activities mentioned in the previous section, project activities should be reviewed. This review should comprise:

- What worked well?
- What did not work well?
- What were the actual costs?
- What was the actual schedule?
- What assumptions need revision?
- What risks materialized and required attention (as documented in the project risk register)?
- Did risk mitigation strategies employed achieve their objectives?
- What are the project economics as constructed and operated?

As a result of this review or project post mortem, the project manager should write a project closeout report that includes the results of the project closeout review and recommendations for future similar projects.

16.10 CONCLUSION

Planning is critical to project success. Detailed, systematic, team-involved plans are the foundation for such success. When events cause a change to the plan, project managers must make a new one to reflect the changes so continuous planning is a requirement of project management. Project managers must focus on three dimensions of project success. Project success means completing all project deliverables on time, within budget, and to a level of quality that is acceptable to sponsors and stakeholders. The project manager must keep the team's attention focused on achieving these broad goals and the stakeholders aligned to the project objectives. It is essential that the project team be composed of all key disciplines who create or use the deliverables. The responsibilities of all team members should be clearly defined. Project managers must feel, and transmit to their team members, a sense of urgency. Because projects are endeavours with limited time, money, and other resources available, they must be kept moving toward completion. Because most team members have many other priorities, it is up to the project manager to keep their attention on project deliverables and deadlines. Regular status checks, meetings, and reminders are essential. All project deliverables and all project activities must be visualized and communicated in vivid detail. The project manager and project team must create a picture of the finished deliverables in the minds of everyone involved so that all effort is focused in the same direction. Avoid vague descriptions and make sure everyone understands what the final product will be.

Projects require clear approvals and sign-off by sponsors. Clear approval points, accompanied by formal sign-off by sponsors and key stakeholders, should be demarcation points in the evolution of project deliverables. Anyone who has the power to reject or to demand revision of deliverables after they are complete must be required to examine and approve them as they are being built.

Risk management is an essential responsibility of project management. All risks should be identified and a contingency plan should accompany all critical risks.

A gas processing project is not complete until the plant, unit, or equipment is placed in service. An evaluation of the operability should be a deliverable upon completion of the project.

REFERENCES

Greer, M., "The Project Manager's Partner: A Step-By-Step Guide to Project Management," 2nd Ed. Human Resource Development (HRD) Press, Amherst, MA (2002).

Hauge, L.H., and Cramer, E., "Project Risk Management in Deepwater Field Developments." Paper presented at the Deep Offshore Technology (DOT) Conference, New Orleans, LA (2002).

HRDC, "Introduction to Project Management Principles." Learning Module, Office of Learning Technologies, Human Resources Development Canada (HRDC), Canada (2003).

McIntire, P., Cost estimating challenges face frontier projects. *Oil Gas J.* **99**, 33 (2001).

Wood, D.A., Probabilistic methods with simulation help predict timing, costs of projects. *Oil Gas J.* **99**(46), 79–83, (2001).

Wood, D.A., and Randall, S., "Implementing ERM." *Oil Gas J.* Part 1: 8–23, 103, 11 (March 21) and Part 2: 20–26, 103, 12 (March 28) (2005).

SUGGESTED READING

Cleland, D., and Kerzner, H., "A Project Management Dictionary of Terms." Van Nostrand, New York (1985).

Clough, R.H., Sears, G.A, and Sears, S.K., "Construction Project Management," 4th Ed. Wiley, New York (2000).

http://www.asapm.org/l_compmodel.asp

http://www.coso.org

http://www.michaelgreer.com

http://www.pmforum.org/library/glossary/index.htm

http://www.pmi.org/prod/groups/public/documents/info/pp_pmbok2k_conf.asp

Pinto, J., and Taylor, J., "Essentials of Project Control." Project Management Institute, PA (Sept., 1999).

Wysocki, R., and McGary, R., "Effective Project Management: Traditional, Adaptive, Extreme." 3rd Ed. Wiley, Indianapolis, IN (2003).

APPENDIX 1

THREE-PHASE FLASH CALCULATION FOR HYDROCARBON SYSTEMS CONTAINING WATER

One of the most important engineering problems encountered in modeling chemical and petroleum processes is the multiphase flash problem. The present flash algorithm can be used to calculate phase equilibria for multicomponent systems with three coexisting phases (water, oil, and gas) at equilibrium with both simplicity and accuracy (Mokhatab, 2003).

This three-hase flash algorithm is based on the thermodynamic condition of equal fugacities for each component in each phase. The resulting phase compositions then provide better values for updating the distribution coefficients using an equation of state (EOS). Equations of state have been used successfully to describe the phase behavior of reservoir crude and gas condensates, but the phase behavior of water/reservoir crude oil has not yet been predicted successfully by an EOS. In this study, it is demonstrated that using the present algorithm with the Shinta and Firoozabadi (1995) association model provides reliable and better results in comparison with experimental data for three-phase flash calculation in the vapor–liquid–liquid region.

A generalized model of a three-phase equilibrium system is shown in Figure A1-1. The phases are assumed to be in thermodynamic equilibrium with each other and any component can appear in all three phases.

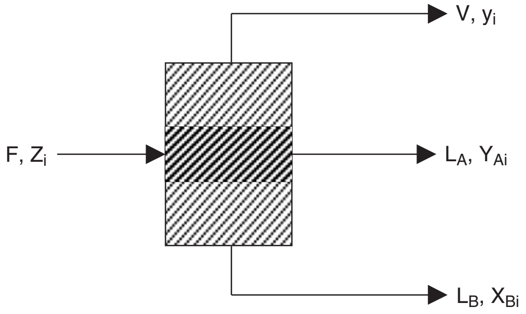

Figure A1-1. Model of a system exhibiting three-phase behavior.

The overall and component material balance around the flash tank gives

$$F = L_A + L_B + V \tag{A1-1}$$

$$FZ_i = L_A X_{Ai} + L_B X_{Bi} + V_{yi} \tag{A1-2}$$

where F is total moles of feed, lbmole; L_A is total moles of hydrocarbon-rich liquid, lbmole; L_B is total moles of water-rich liquid, lbmole; V is total moles of vapor, lbmole; Z_i is mole fraction of component i in the feed; X_{Ai} is mole fraction of component i in the hydrocarbon-rich liquid phase; X_{Bi} is mole fraction of component i in the water-rich liquid phase; y_i is mole fraction of component i in the vapor phase.

The relations describing compositions in each phase that must be satisfied are

$$\sum_{i=1}^{n} X_{Ai} = \sum_{i=1}^{n} X_{Bi} = \sum_{i=1}^{n} y_i = 1 \tag{A1-3}$$

where i is indicates each component and n is the number of components.

Also the equilibrium relations between the compositions of each phase are defined by the following expressions (Peng and Robinson, 1976):

$$K_{Ai} = \frac{y_i}{X_{Ai}} = \frac{\hat{\phi}_i^A}{\hat{\phi}_i^V} \tag{A1-4}$$

$$K_{Bi} = \frac{y_i}{X_{Bi}} = \frac{\hat{\phi}_i^B}{\hat{\phi}_i^V} \tag{A1-5}$$

where K_{Ai} is equilibrium ratio of component i in the hydrocarbon-rich liquid phase, K_{Bi} is equilibrium ratio of component i in the water-rich liquid phase, $\hat{\phi}_{\mathrm{i}}^{\mathrm{A}}$ is fugacity coefficient of component i in the hydrocarbon-rich liquid phase, and $\hat{\phi}_{\mathrm{i}}^{\mathrm{B}}$ is fugacity coefficient of component i in the water-rich liquid phase.

Combining Equations (A1-1) through (A1-5) gives the following equations:

$$\sum_{i=1}^{n} x_{\mathrm{Ai}} = \sum_{i=1}^{n} \left[\frac{z_{\mathrm{i}}(1 - K_{\mathrm{Ai}})}{\frac{L_{\mathrm{A}}}{F}(1 - K_{\mathrm{Ai}}) + \frac{L_{\mathrm{B}}}{F}\left(\frac{K_{\mathrm{Ai}}}{K_{\mathrm{Bi}}} - K_{\mathrm{Ai}}\right) - K_{\mathrm{Ai}}} \right] \tag{A1-6}$$

$$\sum_{i=1}^{n} x_{\mathrm{Bi}} = \sum_{i=1}^{n} \left[\frac{z_{\mathrm{i}} K_{\mathrm{Ai}}}{\frac{L_{\mathrm{A}}}{F}(1 - K_{\mathrm{Ai}}) + \frac{L_{\mathrm{B}}}{F}\left(\frac{K_{\mathrm{Ai}}}{K_{\mathrm{Bi}}} - K_{\mathrm{Ai}}\right) - K_{\mathrm{Ai}}} \right] \tag{A1-7}$$

$$\sum_{i=1}^{n} y_{\mathrm{i}} = \sum_{i=1}^{n} \left[\frac{z_{\mathrm{i}}}{\frac{L_{\mathrm{A}}}{F}(1 - K_{\mathrm{Ai}}) + \left(\frac{L_{\mathrm{B}}}{F}\right)\left(\frac{K_{\mathrm{Ai}}}{K_{\mathrm{Bi}}} - K_{\mathrm{Ai}}\right) - K_{\mathrm{Ai}}} \right] \tag{A1-8}$$

The combining of the these equations can then be used to determine the phase and volumetric properties of the three-phase systems.

Peng and Robinson (1980) proposed the following equations for three-phase flash calculations:

$$\sum_{i=1}^{n} x_{\mathrm{Ai}} - \sum_{i=1}^{n} y_{\mathrm{i}} = 0, \left[\sum_{i=1}^{n} x_{\mathrm{Bi}} \right] - 1 = 0 \tag{A1-9}$$

Provided that the equilibrium ratios and the overall composition are known, the aforementioned equations can be solved simultaneously by using the modified Rachford and Rice (1952) iterative method. In the course of making phase equilibrium calculations, it is always desirable to provide initial values for the equilibrium ratios so the iterative procedure can proceed as reliably and rapidly as possible. Peng and Robinson (1980) adopted Wilson's (1969) equilibrium ratio correlation to provide initial

K_A values, as follow:

$$K_{Ai} = \frac{P_{ci}}{P} \exp\left[5.3727(1 + \omega_i)\left(1 - \frac{T_{Ci}}{T}\right)\right] \tag{A1-10}$$

where P is system pressure, psia; T is system temperature, °F; P_{Ci} is critical pressure of component i, psia; T_{Ci} is critical temperature of component i, °F; and ω_i is acentric factor of component i.

While for determination of initial K_B values, Peng and Robison (1980) proposed the following expression:

$$K_{Bi} = 10^6 \left[\frac{P_{ci}.T}{P.T_{ci}}\right] \tag{A1-11}$$

A logic diagram for three-phase flash calculations is shown in Figure A1-2.

In this method, the fugacity coefficients can be obtained from Peng and Robinson (1976) EOS for nonpolar compounds. Note that in the association equation of state (AEOS), the compressibility factor of the association component such as water is subdivied into two physical and chemical parts as follow (Prausnitz *et al.*, 1986):

$$Z = Z^{ph} + Z^{ch} - 1 \tag{A1-12}$$

The physical compressibility factor, Z^{ph}, can be obtained from modified Peng and Robinson (1980) EOS. To evalute Z^{ch} in Equation (A1-12), the Shinta and Firoozabadi association model is used. Also, the fugacity coefficient of an associating component in each phase is the sum of both chemical and physical contributions as follows (Anderko, 1991):

$$\ln(\phi_i Z) = \ln\left(\phi_i^{ph} Z^{ph}\right) + \ln\left(\phi_i^{ch} Z^{ch}\right) \tag{A1-13}$$

where chemical fugacity coefficients are obtained by the Shinta and Firoozabadi (1995) association model. Note that in using the AEOS, only binary interaction coefficients between water and hydrocarbon, and nonhydrocarbon components and hydrocarbon cuts are required because calculations are often very sensitive to those. Therefore, in this method, binary interaction coefficients suggested by Nishiumi and Arai (1988), Peng and Robinson (1980), and Shinta and Firoozabadi (1995) are used.

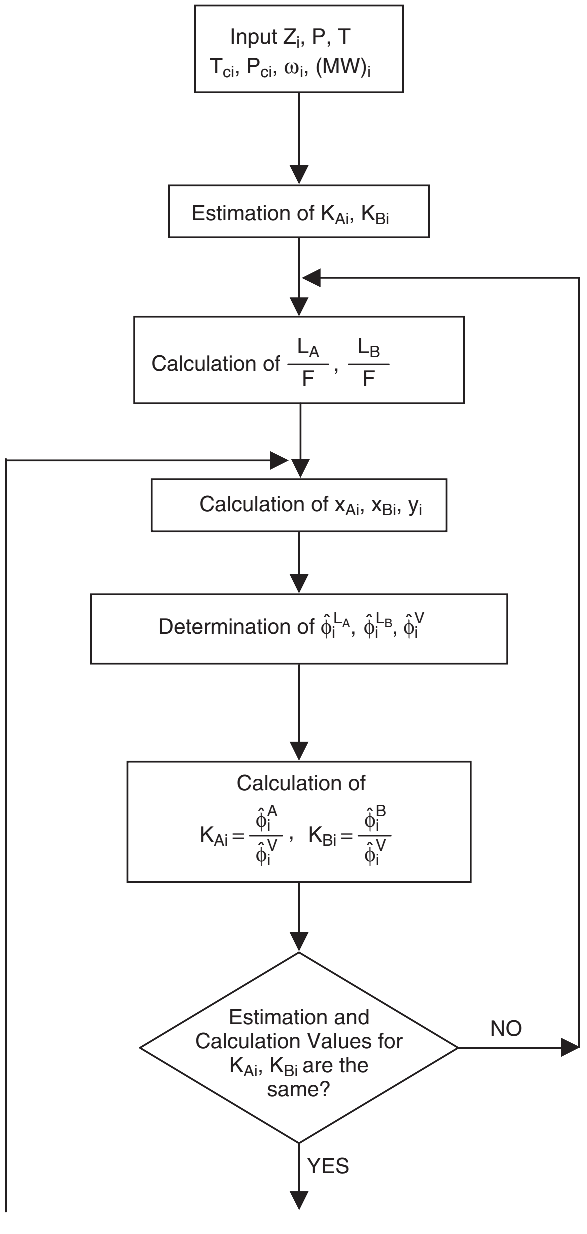

Figure A1-2. Logic diagram for three-phase flash calculations.

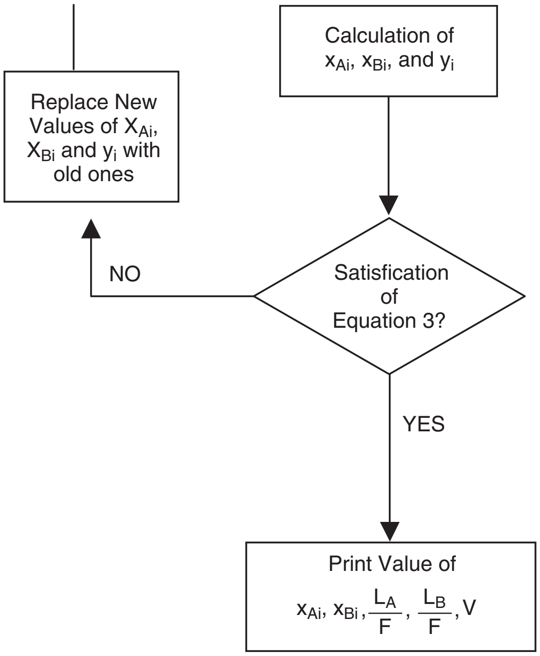

Figure A1-2. Continued.

REFERENCES

Anderko, A., Phase equilibria in aqueous systems from an equation of state based on the chemical approach. *Fliud Phase Equilib*. **65**, 89–110 (1991).

Mokhatab, S., Three-phase flash calculation for hydrocarbon systems containing water. *J. Theor. Found. Chem. Eng*. **37**(3), 291–294 (2003).

Nishiumi, H., and Arai, T., Generalization of the binary interaction parameter of the peng-Robinson EOS by component family. *Fluid Phase Equilib.* **42**, 43–62 (1988).

Peng, D.Y., and Robinson, D.B., Two and three-phase equilibrium calculations for systems containing water. *Can. J. Chem. Eng*. **54**, 595–599 (1976).

Peng, D.Y., and Robinson, D.B., "Two and Three-Phase Equilibrium Calculations for Coal Gasification and Related Process." ACS

Symposium Series, Thermodynamics of Aqueous Systems with Industrial Applications, 133, 193 (1980).

Prausnitz, J.M., Lichtenthaler, R.N., and Azevedo, E.G., "Molecular Thermodynamics of Fluid-Phase Equilibria," 2nd Ed. Prentice-Hall, Englewood Cliffs, NJ (1986).

Rachford, H.H., and Rice, J.D., Procedure for use of electronic digital computers in calculating flash vaporization hydrocarbon equilibrium. *Trans. AIME* **195**, 327 (1952).

Shinta, A.A., and Firoozabadi, A., Equation of state representation of aqueous mixtures using an association model. *Can. J. Chem. Eng.* **73**, 367–379 (1995).

Wilson, G., "A Modified Redich-Kwong Equation of State, Application to General Physical Data Calculation." Paper No. 15C Presented at AIChE 65th National Meeting, Cleveland, OH (4–7 May, 1969).

APPENDIX 2

CONVERSION FACTORS

Length

1 m = 39.37 inch = 10^6 μm
= 10^{10} A°
1 inch = 2.54 cm
1 ft = 30.48 cm = 0.3048 m
1 mile = 5280 ft = 1760 yds
= 1609.344 m

Mass

1 lbm = 453.6 gr = 0.4536 kg
= 7000 grain
1 kg = 1000 gr = 2.2046 lbm
1 slug = 1 lbf s^2/ft = 32.174 lbm
1 US ton = 2000 lbm (also called short ton)
1 long ton = 2240 lbm (also called British ton)
1 tonne = 1000 kg (also called metric ton)

Force

1 lbf = 4.448 N = 4.448 × 10^5 dynes = 32.174 poundals
= 32.174 lbm.ft/s^2
= 1 lbw

Energy

1 J = 1 W.s = 1 kg m^2/s
= 1 N.m^3
= 10^7 dyne.cm = 10^7 erg
1 Btu = 778 ft.lbf = 252 cal
= 1055 J = 10.41 lit.atm
1 HP.hr = 2545 Btu
1 kW.hr = 3412 Btu
= 1.341 HP.hr

Power

1 HP = 550 ft.lbf/s
= 33000 ft. lbf/min
= 746 W = 0.746 kW

Gas Constant

R = 1.9859 Btu/lbmole °R
= 1.9859cal/grmole.°K
= 0.73024 atm ft^3/lbmole.°R
= 1545.3 ft.lbf/lbmole.°R
= 10.732 psia.ft^3/lbmole.°R
= 0.082057 lit. atm/grmole.°K
= 82.057 atm. Cm^3/ grmole.°K
= 8314.5 Pa.m^3/kgmole.°K or J/kg mole.°K
= 8.3145 kJ/kg mole.°K

Pressure

1 atm = 14.696 psia
= 2116 lbf/ft^2
= 29.92 in Hg = 760 mm Hg = 760 Torr.
= 1.013 bar
= 33.9 ft H_2O = 1.013×10^5 Pa
= 101.3 kPa
Pa = 1 N/m^2 = 10^{-5} bars

Volume

1 ft^3 = 7.4805 U.S. gal
= 6.23 Imperial gal
= 28.317 lit
1 m^3 = 1000 lit = 264.2 U.S. gal
= 35.31 ft^3
1 bbl = 42 U.S. gal (oil)
= 5.615 ft^3
1 lit = 1000 cc

Density

Water = 62.43 lbm/ft^3
= 1000 kg/m^3
= 1 g/cm^3
= 8.346 lbm/US gal
mercury = 13.6 g/cm^3

Temperature

°F = 1.8 (°C) + 32
°R = °F + 459.67
= 1.8 (°K)

Dynamic Viscosity

1 cp = 0.01 Poise = 0.01 g/cm s
= 0.01 $(dyne.s)/cm^2$
= 0.001 kg/m.s = 0.001 Pa.s
= 0.001 $N.s/m^2$
= 2.42 lbm/ft.hr = 0.0752 slug/ft.hr
= 6.72×10^{-4} lbm/ft.s
= 2.09×10^{-5} $lbf.s/ft^2$
1 Pa.s = 0.0209 $lbf.s/ft^2$
= 0.672 lbm/ft.s

Kinematic Viscosity

1 St = 1 cm^2/s = 0.0001 m^2/s
1 ft^2/s = 929 St = 0.0929 m^2/s

Force-mass conversion factor

g_c = 1 $kg.m/s^2.N$
= 1 g $cm/s^2.dyne$
= 32.174 $lbm.ft/s^2.lbf$
= 1 $slug.ft/s^2.lbf$

Acceleration due to gravity

g = 32.2 ft/s^2
= 9.81 m/s^2 = 981 cm/s^2
(varies very slightly with longitude and elevation)

APPENDIX 3

PHYSICAL PROPERTIES OF FLUIDS

Table A3-1
Properties of Hydrocarbons and Common Gases (GPSA, 1987)

Component	Formula	Molecular weight	Boiling point (°F) (1 atm)	Vapor pressure at 100°F, (psia)	Critical		Liquid specific gravity (60/60°F)	Volume ratio, Scf gas per gal liquid
					Pressure (psia)	Temperature (°F)		
Methane	CH_4	16.043	−258.73	(5000)[a]	666.4	−116.67	0.3[a]	59.135[a]
Ethane	C_2H_6	30.070	−127.49	(800)[a]	706.5	89.92	0.35619[c]	37.476[c]
Propane	C_3H_8	44.097	−43.75	188.64	616.0	206.06	0.50699[c]	36.375[c]
iso–Butane	C_4H_{10}	58.123	10.78	72.581	527.9	274.46	0.56287[c]	30.639[c]
n-Butane	C_4H_{10}	58.123	31.08	51.706	550.6	305.62	0.58401[c]	31.790[c]
iso–Pentane	C_5H_{12}	72.150	82.12	20.445	490.4	369.10	0.62470	27.393
n-Pentane	C_5H_{12}	72.150	96.92	15.574	488.6	385.8	0.63112	27.674
n-Hexane	C_6H_{14}	86.177	155.72	4.960	436.9	453.6	0.66383	24.371
n-Heptane	C_7H_{16}	100.204	209.16	1.620	396.8	512.7	0.68820	21.729
n-Octane	C_8H_{18}	114.231	258.21	0.537	360.7	564.2	0.70696	19.580
n-Decane	$C_{10}H_{22}$	142.285	345.48	0.061	305.2	652.0	0.73421	16.326
Nitrogen	N_2	28.013	−320.45	—	493.1	−232.51	0.80940[d]	91.413[c]
Oxygen	O_2	31.999	−297.33	—	731.4	−181.43	1.1421[d]	112.93[c]
Carbon dioxide	CO_2	44.010	−109.26[b]	—	1071	87.91	0.81802[c]	58.807[c]
Hydrogen sulfide	H_2S	34.08	−76.50	394.59	1300	212.45	0.80144[c]	74.401[c]
Water	H_2O	18.0115	212.00	0.950	3198	705.16	1.00000[d]	175.62[c]
Air	Mixture	28.9625	−317.8	—	546.9	−221.31	0.87476[d]	95.557[c]

continued

Table A3-1
Properties of Hydrocarbons and Common Gases (GPSA, 1987)—cont'd

Component	Acentric factor, (ω)	Flammability limits (vol% in air mixture)		Heating value at 60°F (1 atm, Btu/scf)		Freezing point at 1 atm, °F	Heat of vaporization at 1 atm, Btu/lb
		Lower	Upper	Net	Gross		
Methane	0.0104	5.0	15.0	909.4	1010.0	−296.44[f]	219.45
Ethane	0.0979	2.9	13.0	1618.7	1769.6	−297.04[f]	211.14
Propane	0.1522	2.0	9.5	2314.9	2516.1	−305.73[f]	183.01
iso–Butane	0.1852	1.8	8.5	3000.4	3251.9	−255.82	157.23
n-Butane	0.1995	1.5	9.0	3010.8	3262.3	−217.05	165.93
iso–Pentane	0.2280	1.3	8.0	3699.0	4000.9	−255.82	147.12
n-Pentane	0.2514	1.4	8.3	3706.9	4008.9	−217.05	153.57
n-Hexane	0.2994	1.1	7.7	4403.8	4755.9	−139.58	143.94
n-Heptane	0.3494	1.0	7.0	5100.8	5502.5	−131.05	136.00
n-Octane	0.3977	0.8	6.5	5796.1	6248.9	−70.18	129.52
n-Decane	0.4898	0.7	5.4	7189.6	7742.9	−21.36	119.65
Nitrogen	0.0372	—	—	—	—	−346.00[f]	85.59
Oxygen	0.0216	—	—	—	—	−361.82[f]	91.59
Carbon dioxide	0.2667	—	—	—	—	−69.83	246.47
Hydrogen sulfide	0.0948	4.3	45.5	586.8	637.1	−121.88[f]	235.63
Water	0.3442	—	—	—	—	32.00	970.18
Air	—	—	—	—	—	—	88.20

[a] Above critical point, extrapolated or estimated.
[b] Sublimation point.
[c] At saturation pressure, 60°F.
[d] At normal boiling point.
[e] Gas at 60°F, liquid at normal boiling point.
[f] At the triple point pressure.

Table A3-2
Physical Data for Common Refrigerants (Ludwig, 1983)

Refrigerant	Boiling point (1 atm) °F	Critical temp, ° F	Critical pressure (psia)	Freezing point, °F
R-12, CCl_2F_2	−21.6	233.6	596.9	−252.0
R-11, CCl_3F	74.7	388.4	635.0	−168.0
R-21, $CHCl_2F$	48.0	353.3	750.0	−221.0
R-22, $CHClF_2$	−41.4	204.8	716.0	−256.0
R-113, $CClF_2CCl_2F$	117.6	417.4	495.0	−31.0
R-114, $CClF_2CClF_2$	38.4	294.3	474.0	−137.0
Carbon dioxide, CO_2	−108.4	37.8	1071	−69.9
Ammonia, NH_3	−28.0	271.2	1651	−108
Methyl chloride, CH_3Cl	−10.76	289.6	969	−144
Sulfur dioxide, SO_2	14.0	314.8	1142	−99
Methylene chloride, CH_2Cl_2	103.7	421	640	−143

Table A3-3
Approximate Ratio of Specific Heats ("*k*" values) for Various Gases[a]

Gas	Symbol	Molecular Weight	*k* at 14.7 psia		Density at 14.7 psi and 60°F (lbs./cu. ft.)
			60°F	150°F	
Monatomic	He, Kr, Ne, Hg	—	1.67	—	—
Most diatomic	O_2, N_2, H_2, etc.	—	1.4	—	—
Acetylene	C_2H_2	26.03	1.3	1.22	0.0688
Air	—	28.97	1.406	1.40	0.0765
Ammonia	NH_3	17.03	1.317	1.29	0.451
Argon	A	—	1.667	—	0.1056
Benzene	C_6H_6	78.0	1.08	1.09	0.2064
Butane	C_4H_{10}	58.1	1.11	1.08	0.1535
Isobutane	C_4H_{10}	58.1	1.11	1.08	0.1578
Butylene	C_4H_8	56.1	1.1	1.09	0.1483
iso-Butene	C_4H_8	56.1	1.1	1.09	0.1483
Carbon dioxide	CO_2	44.0	1.3	1.27	0.1164
Carbon monoxide	CO	28.0	1.4	1.4	0.0741
Carbon tetrachloride	CCl_4	153.8	1.18	—	0.406
Chlorine	Cl_2	70.9	1.33	—	0.1875
Dichlorodifiluoromethane	CCl_2F_2	120.9	1.13	—	—
Dichloromethane	CH_2Cl_2	84.9	1.13	—	0.2215
Ethane	C_2H_6	30.0	1.22	1.17	0.0794
Ethylene	C_2H_4	28.1	1.25	1.21	0.0741
Ethyl chloride	C_2H_5Cl	64.5	1.13	—	0.1705
Flue gas	—	—	1.4	—	—
Helium	He	4.0	1.667	—	0.01058
Hexane	C_6H_{14}	88.1	1.08	1.05	0.2276
Heptane	C_7H_{16}	100.2	—	1.04	0.264
Hydrogen	H_2	2.01	1.41	1.40	0.0053
Hydrogen chloride	HCl	36.5	1.48	—	0.09650
Hydrogen sulfide	H_2S	34.1	1.30	1.31	0.0901
Methane	CH_4	16.03	1.316	1.28	0.0423
Methyl chloride	CH_3Cl	50.5	1.20	—	0.1336
Natural gas (approximate)	—	19.5	1.27	—	0.0514
Nitric oxide	NO	30.0	1.40	—	0.0793
Nitrogen	N_2	28.0	1.41	1.40	0.0743
Nitrous oxide	N_2O	44.0	1.311	—	0.1163
Oxygen	O_2	32.0	1.4	1.39	0.0846
Pentane	C_5H_{12}	72.1	1.06	1.06	0.1905
Propane	C_3H_8	44.1	1.15	1.11	0.1164
Propylene	C_3H_6	42.0	1.16	—	0.1112
Sulfur dioxide	SO_2	64.1	1.256	—	0.1694
Water vaper (steam)	H_2O	18.0	1.33*	1.32	0.04761

*At 212°F

[a] Compiled from "Plain Talks on Air and Gas Compression, Fourth of Series," Worthington Corp. and Reciprocating Compressor Calculation Data Cooper-Bessemer Corp. (1956), by permission.

Table A3-4
Physical Properties of Gas Treating Chemicals (GPSA, 1998)

Compound	Monoethanol-amine	Diethanol-amine	Triethanol-amine	Diglycol-amine	Diisopropanol-amine
Formula	$HOC_2H_4NH_2$	$(HOC_2H_4)_2$ NH	$(HOC_2H_4)_3N$	$H(OC_2H_4)_2NH_2$	$(HOC_3H_6)_2NH$
Molecular weight	61.08	105.14	148.19	105.14	133.19
Boiling point at 760 mm Hg, °C	170.5	269	360(decomposes)	221	248.7
Freezing point, °C	10.5	28.0	22.4	−12.5	42
Critical constants					
Pressure, kPa (abs)	5985	3273	2448	3772	3770
Temperature, °C	350	442.1	514.3	402.6	399.2
Density at 20°C, g/cc.	1.018	1.095	1.124	1.058 at 15.6°C	0.999 at 30°C
Wieght, kg/m^3	1016 at 15.6°C	1089 at 15.6°C	1123 at 15.6°C	1057 at 15.6°C	
Relative density 20°C/20°C	1.0179	1.0919 (30°/20°C)	1.1258	1.0572	0.989 at 45°C/20°C
Specific heat at 15.6°C, kJ/(kg. °C)	2.55 at 20°C	2.51	2.93	2.39	2.89 at 30°C
Thermal conductivity at 20°C, W/(m. °C)	0.256	0.220	—	0.209	—
Latent heat of vaporization, kJ/kg	826 at 760 mm Hg	670 at 73 mm Hg	535 at 760 mm Hg	510 at 760 mm Hg	430 at 760 mm Hg
Heat of reaction, kJ/kg of Acid gas					
H_2S			−930	−1568	—
CO_2			−1465	−1977	—
Viscosity, mPa.sec	24.1 at 20°C	350 at 20°C (at 90% wt. solution)	1013 at 20°C (at 95% wt. solution)	40 at 16°C	870 at 30°C 198 at 45°C 86 at 54°C
Refractive index, N_d 20°C	1.4539	1.4776	1.4852	1.4598	1.4542 at 45°C
Flash point, COC, °C	93	138	185	127	124

Table A3-5
Physical Properties of Selected Glycols and Methanol (GPSA, 1998)[a]

Compound	Ethylene glycol	Diethylene glycol	Triethylene glycol	Tetraethylene glycol	Methanol
Formula	$C_2H_6O_2$	$C_4H_{10}O_3$	$C_6H_{14}O_4$	$C_6H_{18}O_5$	CH_3OH
Molecular mass	62.1	106.1	150.2	194.2	32.04
Boiling point[b] at 760 mm Hg, °C	197.3	244.8	285.5	314	64.5
Vapor pressure at 77°F (25°C), mm Hg	0.12	<0.01	<0.01	<0.01	120
Density at 77°F (25°C), kg/m^3	1110	1113	1119	1120	790
Freezing point, °C	−13	−8	−7	−5.5	−97.8
Pour point, °C	—	−54	−58	−41	
Viscosity in centipoise					
77°F (25°C)	16.5	28.2	37.3	44.6	0.52
140°F (60°C)	4.68	6.99	8.77	10.2	
Specific heat					
77°F (25°C), kJ/(kg.K)	2.43	2.30	2.22	2.18	2.52
Flash point, °C (PMCC)	116	124	177	204	12

[a] These properties are laboratory results on pure compounds or typical of the products, but should not be confused with, or regarded as, specifications.

[b] Glycols decompose at temperatures below their atmospheric boiling point. Approximate decomposition temperatures are ethylene glycol, 165°C; triethylene glycol, 207°C; diethylene glycol, 164°C; and tetraethylene glycol, 238°C.

Table A3-6
Specific Gravities of Aqueous Triethylene Glycol Solutions[a]

	Specific gravity at T/60°F = A + Bx+ Cx^2, where *x* is TEG wt%		
T°F	A	B	C
−50	1.0502	1.8268E-3	−5.2009E-6
0	1.0319	1.7466E-3	−4.8304E-6
50	1.0121	1.5247E-3	−2.8794E-6
100	0.9920	1.7518E-3	−5.4955E-6
150	0.9804	1.5410E-3	−4.3884E-6
200	0.9627	1.4068E-3	−3.5089E-6
250	0.9413	1.3205E-3	−2.7991E-6
300	0.9177	1.2511E-3	−2.0848E-6

[a] Union Carbide Chemicals & Plastics Company Inc., "Triethylene Glycol," 39 old Ridgebury Rd., Danbury, CT 06817 (1989).

Table A3-7
Specific Heat for Aqueous Triethylene Glycol Solutions[a]

	Specific heat = $A + BT + CT^2$, where T is temperature, °C		
TEG wt%	A	B	C
0	1.00540	−2.7286E-4	2.9143E-6
10	0.96705	−2.7144E-5	2.4952E-6
20	0.92490	2.0429E-4	2.4524E-6
30	0.88012	4.3000E-4	1.6952E-6
40	0.83229	6.2286E-4	1.3714E-6
50	0.78229	7.9286E-4	1.0857E-6
60	0.72200	9.4000E-4	8.0000E-7
70	0.66688	1.0871E-3	4.7620E-7
80	0.60393	1.2043E-3	2.8571E-7
90	0.53888	1.2800E-3	1.9048E-7
100	0.48614	1.3929E-3	−5.7140E-8

[a] Union Carbide Chemicals & Plastics Company Inc., “Triethylene Glycol,” 39 old Ridgebury Rd., Danbury, CT 06817 (1989).

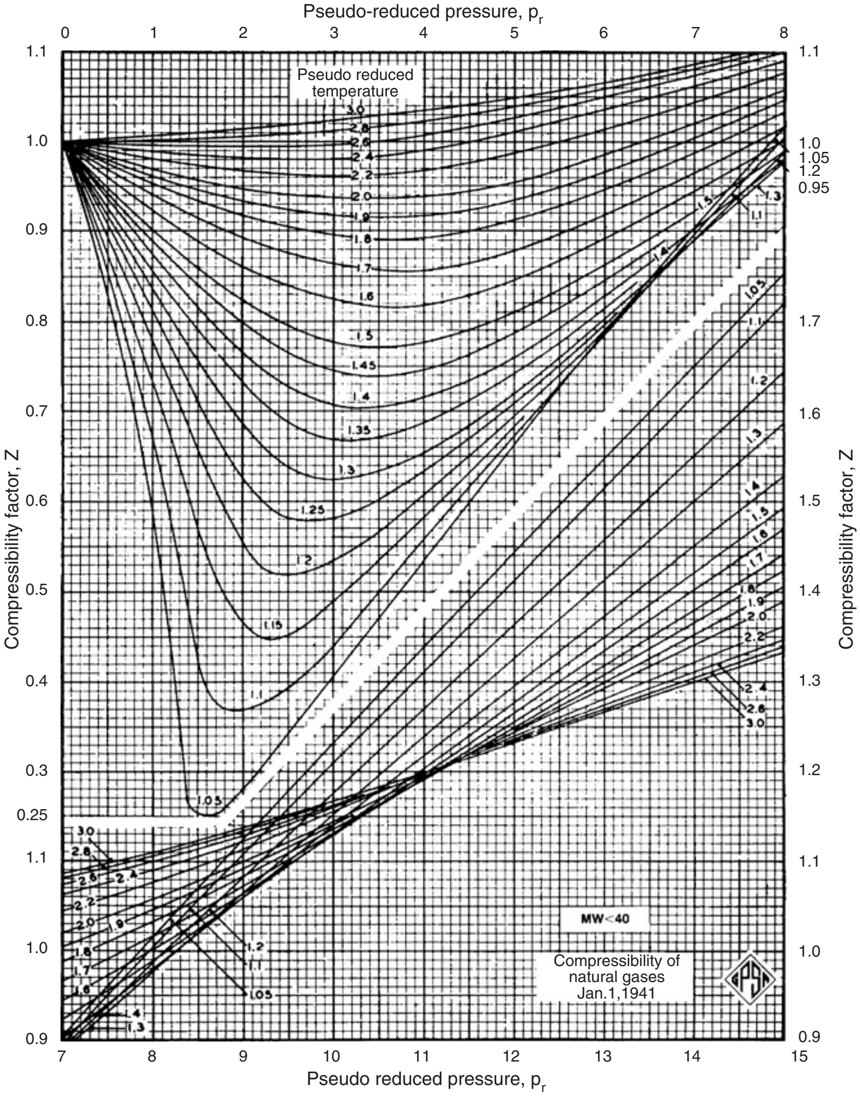

Figure A3-1. Compressibility of natural gases (GPSA, 1998).

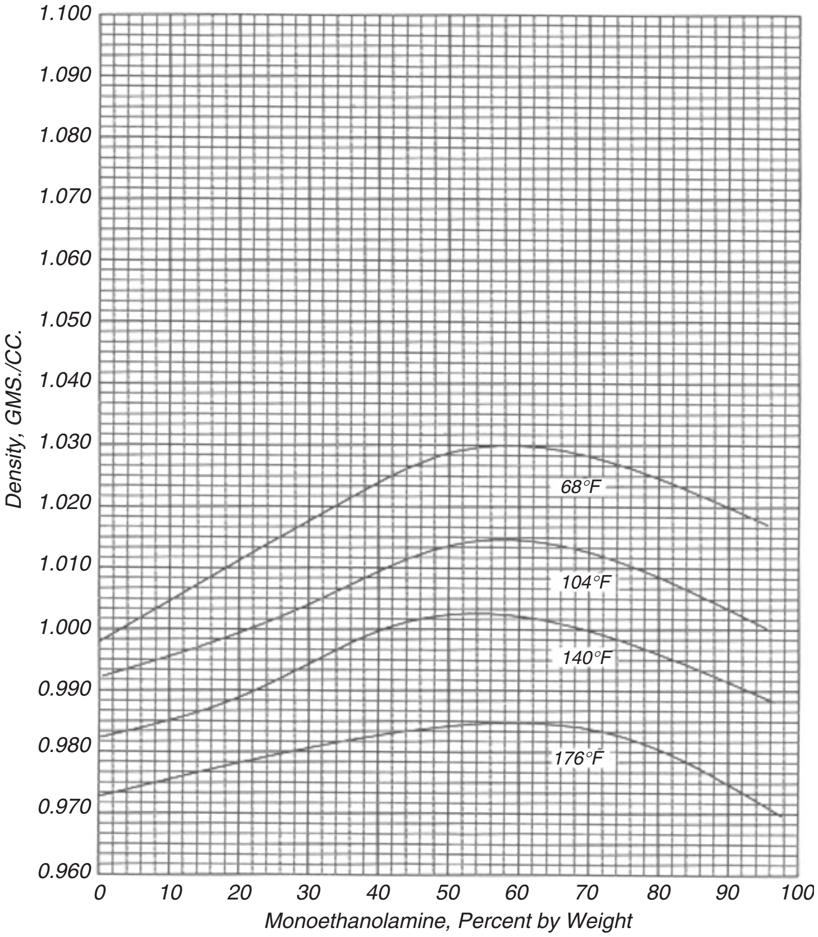

Figure A3-2. Density of MEA solutions (Gas Conditioning Fact Book, Dow Chemical Co., Houston, Texas, 1969).

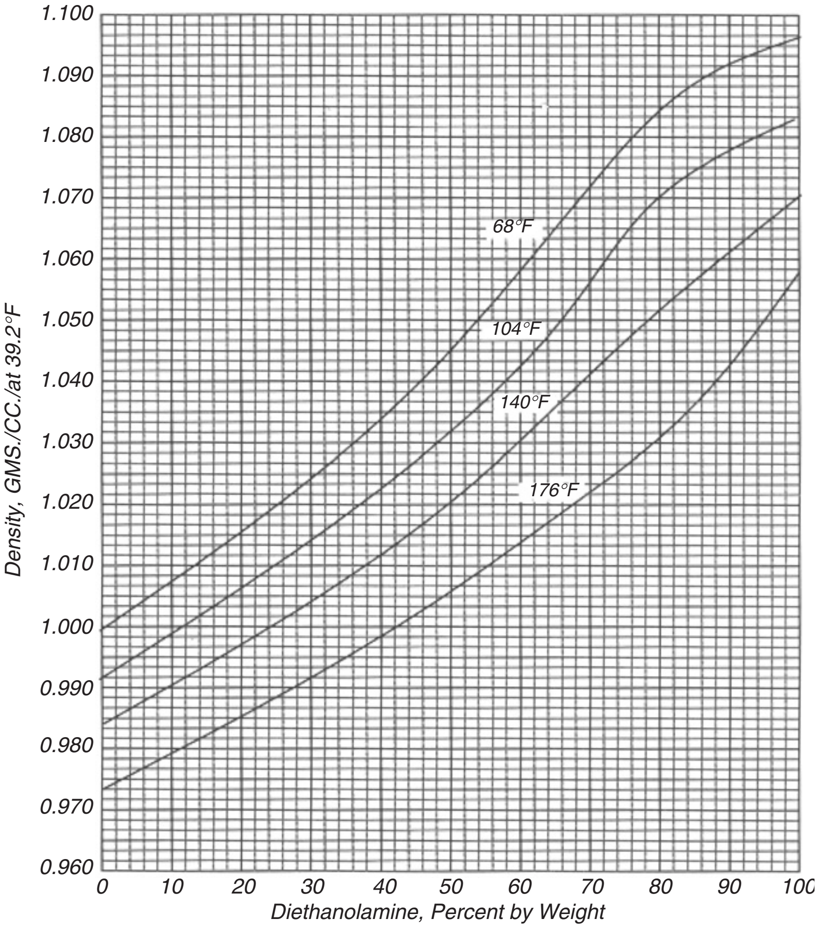

Figure A3-3. Density of DEA solutions (Gas Conditioning Fact Book, Dow Chemical Co., Houston, Texas, 1969).

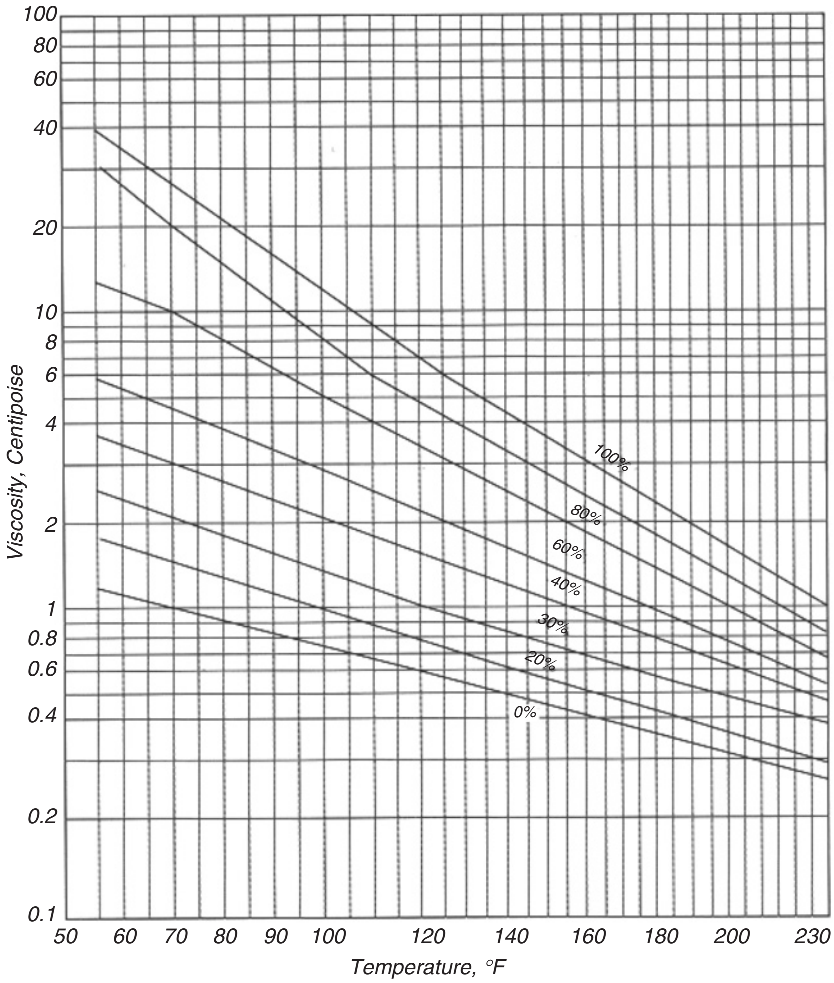

Figure A3-4. Viscosity of aqueous MEA solutions as function of weight percent MEA (Gas Conditioning Fact Book, Dow Chemical Co., Houston, Texas, 1969).

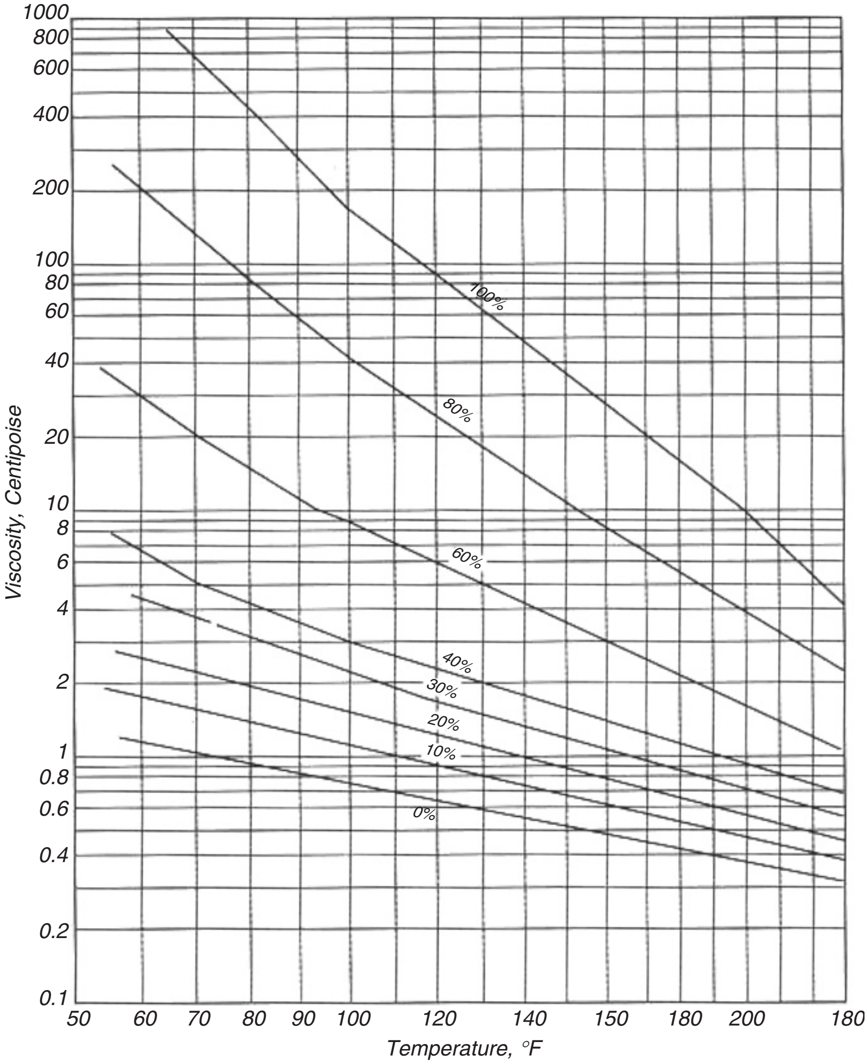

Figure A3-5. Viscosity of aqueous DEA solutions as function of weight percent DEA (Gas Conditioning Fact Book, Dow Chemical Co., Houston, Texas, 1969).

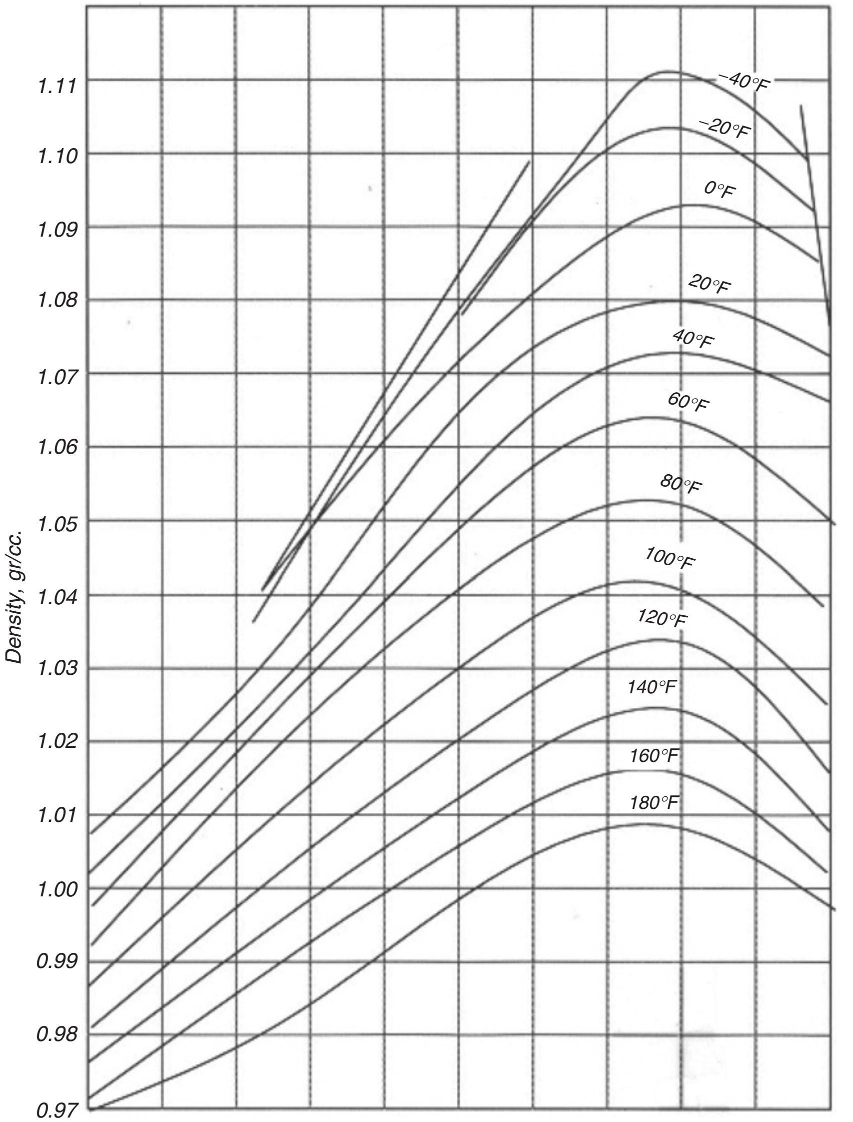

Figure A3-6. Density of aqueous DGA solutions (Gas Treating Data Book, Jefferson Chemical Co., Houston, Texas, 1969).

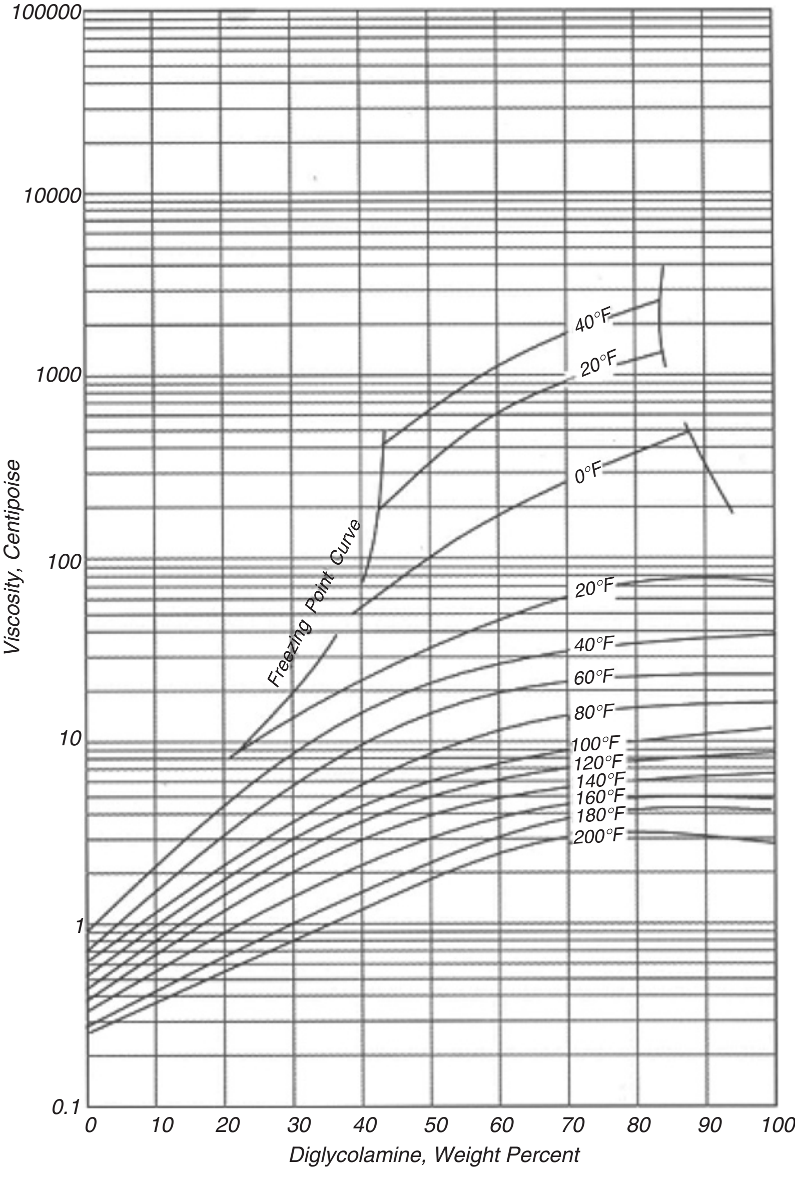

Figure A3-7. Viscosity of aqueous DGA solutions (Gas Treating Data Book, Jefferson Chemical Co., Houston, Texas, 1969).

GLOSSARY AND ACRONYMS

Absorbent: Liquid used to remove specific components from a gas stream. Common examples are TEG to remove water; absorption oil to recover heavy hydrocarbons from natural gas; and amines to remove carbon dioxide or hydrogen sulfide.

Absorption: A separation process involving the transfer of a substance from a gaseous phase to a liquid phase through the phase boundary.

Accumulator: A storage vessel for liquid refrigerant; also known as surge drum.

Acid gas loading: The amount of acid gas, on a molar or volumetric basis, that will be picked up by a solvent.

Acid gases: Impurities in a gas stream usually consisting of CO_2, H_2S, COS, RSH, and SO_2. Most common in natural gas are CO_2, H_2S, and COS.

Adsorbent: Solid pellets used to remove specific components from natural gas.

Adsorption: The process by which gaseous components are adsorbed on solids because of their molecular attraction to the solid surface.

Air cooler: Heat exchanger in which air is blown over a bundle of tubes to cool the fluid in the tubes.

Antifoam: A substance, usually a silicone or a long-chain alcohol, added to the treating system to reduce the tendency to foam.

Associated gas: Gas produced with crude oil and separated in a gas–oil separator.

Blowdown: Water discharged from the system to control concentrations of salt or other impurities in the circulating water (percentage of circulating water rate, or gpm).

Boiling point: Temperature at which the vapor pressure of a liquid equals the external pressure. When external pressure is 1 atm, temperature is called the normal boiling point.

British thermal unit (Btu): The heat required to raise the temperature of 1 pound of water by 1°F at or near 39.2°F.

Bubble point: The temperature at which the vapor pressure of the liquid refrigerant equals the absolute external pressure of the liquid–vapor interface.

Bypass: Pipe connection around a valve (or other equipment) to permit flow to continue while the valve (or equipment) is repaired.

By-product: A material produced in addition to the main product or result.

Capacity, refrigerating system: The cooling effect produced by the total enthalpy change between the refrigerant entering the evaporator and the refrigerant leaving the evaporator.

Capacity: The volume of gas that can be moved through the pipeline or stored in a storage facility over a period of time. The amount pipeline capacity reserved for a customer is determined by contract.

Carbonyl sulfide: A compound containing a carbonyl group and sulfur group (COS). It is a contaminant in natural gas and NGL and is usually removed to meet sulfur specifications.

Chiller, evaporator: A heat exchanger in which the liquid refrigerant is vaporized by a process stream, which is in turn cooled.

Claus: Process name for the type of sulfur recovery system.

Cleaning: Removing liquids and solids from natural gas.

Compressibility factor (Z): Dimensionless ratio of the actual volume occupied by a gas to that predicted using the ideal gas law.

Compression ratio: The ratio of absolute discharge pressure from a compressor to the absolute suction or intake pressure.

Condensate: A light hydrocarbon separated from natural gas and crude oil after production.

Condenser: A heat exchanger in which the refrigerant, compressed to a suitable pressure, is condensed by rejection of heat to a cooling medium.

Conditioning: Operations used to make produced gas and/or oil suitable for storage and transport.

Counterflow: Air flow direction through the fill is countercurrent to that of falling water.

Cracking: The refinery process in which large, heavy, complex hydrocarbon molecules are broken down into simpler and lighter molecules in order to produce a variety of fuel products.

Cross flow: Air flow direction through the fill is essentially perpendicular to that of falling water.

Cryogenic plant: Gas processing plant capable of recovering NGL, including ethane, using a turboexpander to generate very low operating temperatures.

Cycling: Process in which gas produced from a gas reservoir is processed or separated, and the remaining or residue gas is returned to the reservoir. Also called "recycling."

Deepwater: Oil and gas developments in water depths of 1000 feet or more.

Degradation products: Impurities in a treating solution that are formed from both reversible and irreversible side reactions.

Dehydration: The act or process of removing water from gases or liquids.

Desiccant: An adsorbent used to dry a fluid.

Desulfurization: Removal of sulfur and sulfur compounds from gases or liquid hydrocarbons.

Dew point: The temperature and pressure at which liquid initially condenses from a gas or vapor. Specifically applied to the temperature at which water vapor (water dew point) or hydrocarbons (hydrocarbon dew point) condense.

Doctor sweet: A hydrocarbon stream that has had mercaptans removed to a level that passes the Doctor test (GPA-1138).

Downstream pipeline: A pipeline closer to the market as opposed to an upstream pipeline, which is closer to the production area.

Dry natural gas: (1) Natural gas that contains very little natural gas liquids. The gas is primarily methane. It is usually not economical or necessary to process dry gas for the removal of associated NGLs. The heating value is approximately 1000 Btu per cubic feet. (2) Gas with water content reduced by dehydration.

Evaporation loss: Water evaporated from the circulating water into the air stream in the cooling process (percentage of circulating water rate or gpm).

Expansion valve: A valve for controlling the flow of refrigerant to an evaporator or chiller.

Feedstocks: Crude oil, natural gas liquids, natural gas, or other material used as raw materials for the production of gasoline, other refined products, or chemicals.

Flared: The volume of gas burned in flares on the base site or at gas processing plants.

Flash gas: Gas resulting from the instantaneous evaporation of refrigerant by a pressure-reducing device, such as a control valve.

Flash point: Minimum temperature at which a liquid gives off vapor in sufficient concentration to form an ignitable mixture with air near the surface of the liquid.

Flash tank: A vessel used to separate the gas evolved from liquid flashed from a higher pressure to a lower pressure.

Flashing (liquid): When a liquid at its bubble point travels across a restricting control valve, its pressure and temperature decrease. Some of the liquid turns into vapor in the process, for which the word “flash” is used.

Gas processing: All the field operations used to separate the constituents of and remove the contaminants in a natural gas. Purpose is making salable products and/or meeting sales specifications.

Gas–condensate field: Petroleum reservoir in which hydrocarbons exist at high pressure. Lowering the pressure causes condensation of the heavier hydrocarbons, which are then not produced.

Gasoline: A mixture of hydrocarbons used in a spark-ignited internal combustion engine and having an octane rating of at least 60. Aviation gasolines have an octane rating of 100 or more. Automotive gasolines typically have an octane rating from 87 to 94.

Gas-well gas: Natural gas produced from a formation containing little or no hydrocarbon liquid.

Gathering system: Pipelines and other equipment installed to collect, process, and deliver natural gas from the field, where it is produced, to the trunk or main transmissions lines of pipeline systems.

Handling: See “Conditioning.”

HC dew point: The temperature at which a hydrocarbon gas being cooled will begin to condense its components. The dew point depends on gas composition and pressure.

Heavy ends: Relative term used to compare those hydrocarbons that have more carbon atoms than others. For instance, in a mixture containing methane, ethane, propane, butanes, and pentanes, the pentanes (and even the butane) could be referred to as the heavy ends.

Heptane plus (C7+): All hydrocarbon components in a fluid mixture with seven or more carbon atoms in the molecule.

Hydrate: A solid (ice-like) material that is the result of the combination of hydrocarbons with water, usually under high pressure and at low temperature.

Hydrocarbons: Organic chemical compounds of hydrogen and carbon atoms that form the basis of all petroleum products, natural gas, and coals. Hydrocarbons may be liquid, gaseous, or solid.

Lean gas: (1) Residue gas after recovery of NGL. (2) Unprocessed gas containing little or no recoverable NGL.

Lean oil: Absorption oil from which all absorbed NGL have been stripped.

Light ends: The low-boiling, easily evaporated components of a hydrocarbon mixture.

Liquefied petroleum gas (LPG): A compressed or liquefied gas produced as a by-product from refining petroleum. Includes propylene, propane, butylene, butane, isobutane, and other elements.

Makeup: Water added to the circulating water system to replace water lost by evaporation, drift, windage, blowdown, and leakage (percentage of circulating water rate or gpm).

Manifold: A pipe with one or more inlets and two or more outlets, or vice versa.

Mercaptan: Any of a homologous series of compounds of the general formula RSH. All of the materials possess a foul odor, e.g., methyl mercaptan.

Midstream: The intermediate segments of the energy industry downstream of oil and gas production and upstream of end user consumption. These segments provide services to producers and consumers of energy. These services generally include but are not limited to natural gas gathering, processing and wholesale marketing and NGL fractionation, transportation, and storage.

Molecular sieve: Synthetic zeolite (essentially silica-alumina) used in adsorption processes.

Natural gas: Gaseous hydrocarbon mixture found in nature. Principal hydrocarbon components are C1, C2, C3, C4's, and C5+. May also contain N_2, CO_2, H_2O, H_2S.

Natural gasoline: A gasoline obtained by extracting small concentrations of pentane and heavier hydrocarbons present in natural gas.

Pentane plus (C5+): All hydrocarbon components in a fluid mixture with five or more carbon atoms in the molecule.

Pig: Device forced through a pipeline to clean out debris or inspect line.

Pigging: Procedure of forcing a solid object through a pipeline for cleaning or other purposes.

Pipeline gas: Gas that meets a transmission company's minimum specifications.

Plate-fin exchangers: Heat exchangers using thin sheets of metal to separate the hot and cold fluids tube.

Producer: A natural gas producer is generally involved in exploration, drilling, and refinement of natural gas. There are independent producers, as well as integrated producers, which are generally larger companies that produce, transport, and distribute natural gas.

Production: The process of removing natural gas supplies from a gas well and introducing it into a pipeline or gathering system.

Purge: Flow of air or inert gas through furnace, gas passages, and associated flues and ducts that effectively removes any gaseous or suspended combustibles and replaces them with air or inert gas.

Raw gas: Unprocessed gas or the inlet gas to a plant.

Recirculation: Describes a condition in which a portion of the tower's discharge air reenters the air inlets along with the fresh air. Its effect is an elevation of the average entering wet-bulb temperatures compared to the ambient.

Recompressor: A compressor used for some particular service, such as compressing residue gas; implies restoring of pressure after pressure reduction.

Refrigerant: Fluid used for heat transfer in a refrigeration system, which absorbs heat at a low temperature and low pressure and rejects heat at a higher temperature and a higher pressure.

Reid vapor pressure (RVP): Vapor pressure of a liquid determined by laboratory testing in a standard manner in the Reid apparatus at a standard temperature of 37.8°C/100°F (ASTM). The RVP is reported as pounds per square inch at 100°F. The RVP is always less than the true vapor pressure at 100°F.

Residence time: The time period for which a fluid will be contained within a specified volume.

Residue gas: Gas remaining after the recovery of liquid products.

Rich natural gas: Raw, unprocessed natural gas that contains large amounts of assoicated natural gas liquids.

Rich oil: The oil leaving the bottom of an absorber. Lean oil plus all the absorbed constituents.

Rich oil demethanizer: Fractionating column used to remove methane from rich oil.

Sales gas: Same as pipeline gas.

Scavenging: Automatic purge and cleaning of the lines that may contain liquid fuel after a shutdown.

Scrubber or knockout: A vessel designed to handle streams with high gas-to-liquid ratios. The liquid is generally entrained as mist in the gas or is free flowing along the pipe wall. These vessels usually have a small liquid collection section. The terms are often used interchangeably.

Selective treating: Preferential removal of one acid gas component, leaving at least some of the other acid gas components in the treated stream.

Separator: A vessel used to separate a mixed-phase stream into gas and liquid phases that are "relatively" free of each other. Other terms used are scrubbers, knockouts, and decanters.

Shutdown: Control actions undertaken to stop operation of equipment or a process. Shutdown can be triggered automatically or initiated by voluntary action.

Slug catcher: A particular separator design able to absorb sustained inflow of large liquid volumes at irregular intervals. Usually found on gas gathering systems or other two-phase pipeline systems. A slug catcher may be a single large vessel or a manifolded system of pipes.

Sour gas: Gas containing undesirable quantities of hydrogen sulfide, mercaptans, and/or carbon dioxide.

Specific gravity: Dimensionless ratio of the mass of a given volume of a substance to that of an equal volume of another standard substance. Usually air is used as the standard for gases and water for liquids, and volumes are measured at 60°F and atmospheric pressure (or bubble point pressure).

Stabilized condensate: Liquid mixture of hydrocarbons containing iC4, nC4, iC5, nC5, C6, and C7+.

Supervisory control and data acquisition: Remote-controlled equipment used by pipelines and local distribution companies to operate their gas systems. These computerized networks can acquire immediate data concerning flow, pressure, or volumes of gas, as well as control different aspects of gas transmission throughout a pipeline system.

Sweet gas: Gas that has no more than the maximum sulfur content defined by (1) the specifications for the sales gas from a plant and (2) the definition by a legal body, such as the Railroad Commission of Texas.

Sweet: The near or absolute absence of objectionable sulfur compounds in either gas or liquid as defined by a given specification standard.
Tail gas: The exit gas from a plant.
Tail gas cleanup unit: A process unit designed to take tail gas from a Claus sulfur recovery plant and remove additional sulfur with the goal of meeting environmental sulfur emission standards.
Three-phase separator: A vessel used to separate gas and two immiscible liquids of different densities (e.g., gas, water, and oil).
Throughput: The sum of volumes of gas sold and volumes of gas transported for customers.
Turboexpander: A device that converts the energy of a gas or vapor stream into mechanical work by expanding the gas or vapor through a turbine.
True vapor pressure: The pressure at which the gas and liquid in a closed container are in equilibrium at a given temperature.
Vapor pressure: Pressure exerted by a liquid when in equilibrium with a vapor.
Volumetric efficiency: Actual fluid volume taken in by a compressor divided by volume displaced by its piston or pistons.
Water dew point: The temperature at which water begins to condense from a gas as it is cooled.
Wet natural gas: Raw, unprocessed natural gas that contains large amounts of associated natural gas liquids.
Working pressure: Maximum pressure at which equipment can be used.

Acronyms

acf: Actual cubic feet
acfm: Actual cubic feet per minute
AG: Acid gas
AGA: American Gas Association
AISI: American Iron & Steel Institute
ANSI: American National Standards Institute
API: American Petroleum Institute – National trade association of U.S. petroleum industry, a private standardizing organization.
ASME: American Society of Mechanical Engineers
ASTM: American Society for Testing and Materials
Atmosphere

ATM:	Atmosphere
bara:	Unit of pressure, bar absolute
barg:	Unit of pressure, bar gauge
bbl:	Barrel (42 U.S. gallons). The oil industry standard for volumes of oil and its products. Always reduced to 60°F and vapor pressure of the liquid.
BHP:	Brake horsepower
BS & W:	Basic sediment and water; water and other contaminants present in crude oil
Btu:	British thermal unit
C1:	Methane
C2:	Ethane
C3:	Propane
C4's:	Butanes
C5's:	Pentanes
C6:	Hexane
C6+:	Hexane and heavier
C7:	Heptane
C7+:	Heptane and heavier
C8:	Octane
Cfm:	Cubic feet per minute
CO_2:	Carbon dioxide
DEA:	Diethanolamine
DEG:	Diethylene glycol
DGA:	Diglycolamine
DIPA:	Diisopropanolamine
EG:	Ethylene glycol
ft/sec:	Feet per second
FWKO:	Free-water knockout-separator that removes water from the crude oil or natural gas
Gal:	U.S. gallon
GLR:	Gas liquid ratio, expressed as scf/bbl
GOR:	Gas–oil ratio, combined gas released from stage separation of oil, expressed as scf/Bsto
gpm/GPM:	(1) gpm (gallons per minute): flow rate in gallons per minute (2) GPM – Preferably gal/Mcf: natural gas liquids content in gallons of liquid products per thousand cubic feet.
H_2O:	Water
H_2S:	Hydrogen sulfide

HC:	Hydrocarbon
HP:	High pressure
iC4:	Isobutane
iC5:	Isopentane
ID:	Internal diameter
J-T:	Joule–Thompson (constant enthalpy) expansion
KW:	Kilowatts
lb:	Pounds
LNG:	Liquefied natural gas
LP:	Low pressure
LPG:	Liquefied petroleum gas
Max:	Maximum
Mcf:	One million cubic feet
Mcfd:	Thousand cubic feet per calendar day
MDEA:	Methyldiethanolamine
MEA:	Monoethanolamine
Min:	Minimum
MMscfd:	Million standard cubic feet per calendar day
MTA:	Mass transfer agent, e.g., absorber oil
MTZ:	Mass transfer zone
MW:	Molecular weight
NACE:	National Association of Corrosion Engineers
nC4:	Normal butane
nC5:	Normal pentane
NGL:	Natural gas liquids, includes ethane, propane, butanes, pentanes, or mixtures of these
OD:	Outside diameter
ppm:	Parts per million
ppmv:	Parts per million by volume
ppmw:	Parts per million by weight
psi:	Pounds per square inch
psia:	Pounds per square inch absolute
psig:	Pounds per square inch gauge
PVC:	Polyvinyl chloride
ROD:	Rich oil demethanizer
RP:	Recommended practices, e.g., API RP I4E
rpm:	Revolutions per minute
RS:	Relative saturation

RVP:	Reid vapor pressure
scfm:	Standard cubic feet per minute
SG:	Specific gravity
SI:	Abbreviation (1) shut in, (2) System International (French for International System of Units)
SPE:	Society of Petroleum Engineers
SRU:	Sulfur recovery unit
TEG:	Trictylene glycol
TEMA:	Tubular Exchanger Manufacturers Association
VLE:	Vapor–liquid equilibrium
VLLE:	Vapor–liquid–liquid equilibrium
VOC:	Volatile organic compounds
Δ:	Increment or difference
°API:	Degrees API gravity
°F:	Degrees Fahrenheit

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