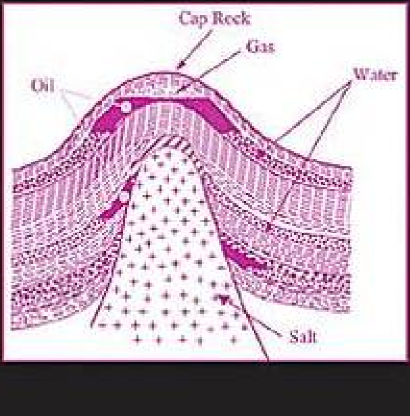
PETROLEUM AND GAS FIELD PROCESSING



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MARCEL DEKKER
MARCEL DEKKER, INC. NEW YORK • BASEL

Library of Congress Caraloging-in-Publication Data
A catalog record for this book is available from the Library of Congress.

ISBN: 0-8247-0962-4
This book is printed on acid-free paper.

Headquarters
Marcel Dekker, Inc., 270 Madison Avenue, New York, NY 10016, U.S.A.
tel: 212-696-9000; fax: 212-685-4540

Distribution and Customer Service
Marcel Dekker, Inc., Cimarron Road, Monticello, New York 12701, U.S.A.
tel: 800-228-1160; fax: 845-796-1772

Eastern Hemisphere Distribution
Marcel Dekker AG, Hutgasse 4, Postfach 812, CH-4001 Basel, Switzerland
tel: 41-61-260-6300; fax: 41-61-260-6333

World Wide Web
http://www.dekker.com

The publisher offers discounts on this book when ordered in bulk quantities. For more information, write to Special Sales/Professional Marketing at the headquarters address above.

Current printing (last digit):

10 9 8 7 6 5 4 3 2 1

PRINTED IN THE UNITED STATES OF AMERICA

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Preface

This book is written mainly for two groups of readers: engineering students studying petroleum and chemical engineering and graduate engineers whose major interest is gas and petroleum production and processing operations. The book is arranged so that it can be used as both a text and a reference. As a text, the organization of materials permits flexibility in designing courses in this field for both undergraduate and postgraduate students.

The subject matter is based on lectures given and notes prepared by the authors and materials taken from the literature. It provides well-organized theoretical and technical background knowledge on oil and gas field operations. Emphasis is given to the separation of the produced reservoir fluids, oil, gas, and water, and their subsequent treatments at various stages in the oil field in order to produce marketable quantities of oil and gas.

The preparation of the book was a joint effort of Dr. Abdel-Aal and the coauthors, who wrote chapters for which they were especially qualified and possess valuable experience. Dr. Aggour prepared Chapters 1, 4, and 9 and coauthored Chapters 3 and 5 with Dr. Abdel-Aal. Dr. Fahim prepared Chapters 6, 11, and 12. Dr. Abdel-Aal wrote Chapters 2, 7, 8, 10, and 13. Significant contributions to the editing of the final manuscript and some major additions to other chapters are credited to Dr. Aggour, on top of the five chapters for which he had primary responsibility.

Following the Introduction, the book contains 13 chapters grouped into four parts. Part I offers general background on the production of oil and gas and crude oil composition. Part II is concerned with the separation of the produced fluids: both two- and three-phase separation. Treatment of the produced fluids—namely, crude oil and water—is considered in Part III, which includes dehydration, desalting, stablization/sweetening of crude oil, and water treatment. Finally, Part IV deals with field processing and treatment of natural gas, which covers dehydration and hydrate inhibition, sweetening (acidic gas removal), and separation and fractionation of natural gas constituents.

Illustrative examples and sample problems are used extensively in the text to demonstrate the application of the principles to practical cases. Students majoring in petroleum engineering, chemical engineering, and allied areas will have the opportunity to learn and benefit from many of these examples and problems.

The preparation of a book that covers such a broad subject requires many sources of information. Although we have made every effort to extend proper acknowledgement, it is possible that some credits have been omitted inadvertently. We gratefully acknowledge our indebtedness to many individuals and organizations for their help and support in providing data and information.

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Part IV: Field Processing and Treatment of Natural Gas

Chapter 10. Overview of Gas Field Processing

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Chapter 12. Gas Dehydration

Chapter 13. Recovery, Separation, and Fractionation of Natural Gas Liquids

Introduction

Fluids produced from oil and gas wells genrerally constitute mixtures of crude oil, natural gas, and salt water. These mixtures are very difficult to handle, meter, or transport. In addition to the difficulty, it is also unsafe and uneconomical to ship or to transport these mixtures to refineries and gas plants for processing. Further, hydrocarbon shipping tankers, oil refineries, and gas plants require certain specifications for the fluids that each receive. Also, environmental constraints exist for the safe and acceptable handling of hydrocarbon fluids and disposal of produced salt water. It is therfore necessary to process the produced fluids in the field to yield products that meet the specifications set by the customer and are safe to handle.

I. CRUDE OIL PROCESSING

Crude oil–gas–water mixtures produced from wells are generally directed, through flow lines and mainfold system, to a central processing and treatment facility normally called the gas–oil separation plant (GOSP). The first step in processing of the produced stream is the separation of the phases (oil, gas, and water) into separte streams. This takes place in mechanical devices known as *two-phase gas–oil separators* when the produced stream contains no water or *three-phase separators* when the produced stream contains water. Gas–oil separation carried out in these separators is recognized as the backbone process in a train of field processing units of oil and gas operations. The separators are used to relieve the excess pressure due to the gas associated with the produced crude and, consequently, separating it from the oil. When water exists in the produced stream, separators are also used to separate the *free water* from the oil. Once separation is done, each stream undergoes the proper processing for further field treatment, as shown in Fig. 1.

Oil leaving the separator does not generally meet the purchaser's specifications. Oil may still contain between 10% and 15% water that exists

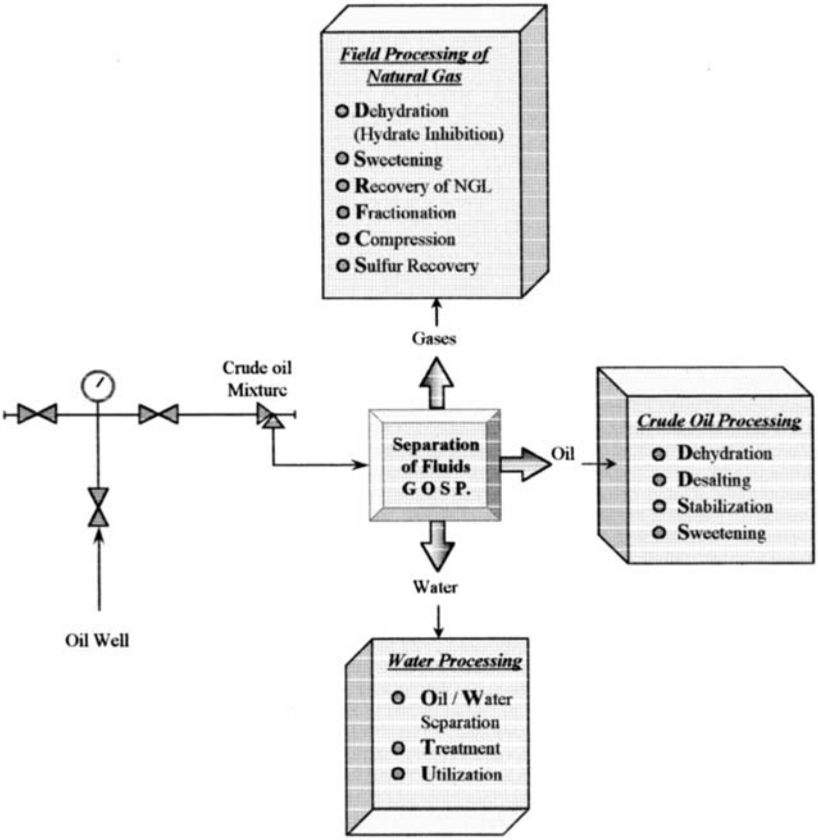


Figure 1 An outline of the processing surface field operations.

mostly as emulsified water. The presence of this salt water presents serious corrosion and scaling problems in transportation and refinery operations. Water remaining in the oil is known as the *basic sediments and water* (BS&W). A maximum of 1% BS&W and in some cases less than 0.5% BS&W is acceptable. The limit on the salt content of the *remnant water* in oils is usually in the range of 10 to 15 PTB (pounds of salt per thousand barrels of oil). If these specifications are not met, then further treatment of the oil leaving the separator will be needed. Such treatment involves *emulsion treatment/dehydration* and *desalting processes*.

After oil treating, there may be a need to *stabilize* the crude oil to optimize the oil recovery and reduce its volatility. Some produced crude oils contain hydrogen sulfide and other sulfur products. When it contains more than 400 ppm of H_2S gas, the oil is classified as *sour crude*. Sour crude oils present serious safety and corrosion problems. In such cases, another treatment known as the *sweetening process* is needed to remove hydrogen sulfide or reduce its content to acceptable limits.

A quick comparison between the properties of a typical crude oil produced from a field in the Middle East before and after treatment illustrates the role of field processing the crude oil [1].

Raw crude oil has the following materials present:

Water: present in two forms: emulsions (10%) and free water (30%)
Salt: 50,000–250,000 mg/L formation water
Gas: dissolved gas (600 scf/bbl crude oil)
H_2S: 1000 ppm

Treated crude oils have the following final specifications:

Water content: 0.3 vol% maximum
Salt content: 10 lbs (as NaCl) per 1000 barrels of oil
H_2S: 70 ppm
Vapor pressure: 10 psig (4–5 psi RVP)

II. FIELD PROCESSING OF NATURAL GAS

Nature gas associated with oil production (*associated gas*) or produced from gas fields generally contains undesirable components such as H_2S, CO_2, N_2, and water vapor. Field processing of natural gas implies the removal of such undesirable components before the gas can be sold in the market. Specifically, the gas contents of H_2S, CO_2, and water vapor must be removed or reduced to acceptable concentrations. N_2, on the other hand, may be removed if it is justifiable. Gas compression is usually needed after these treatment processes. In addition, the gas undergoes separation and fractionation for the recovery of some hydrocarbon components that are utilized as a feedstock for the petrochemical industry.

Several schemes can be recommended for field processing and separation of natural gas, but the specific solution is usually a function of the composition of the gas stream, the location of this source, and the markets available for the products obtained. Saudi Arabia's master gas system (MGS), as illustrated in Fig. 2, is a good example.

Nonassociated gases, including gas condensates, are also a potential source of hydrocarbons for many oil-producing countries. The processing of the gas condensates, however, involves a low-temperature technique to be described later in the text.

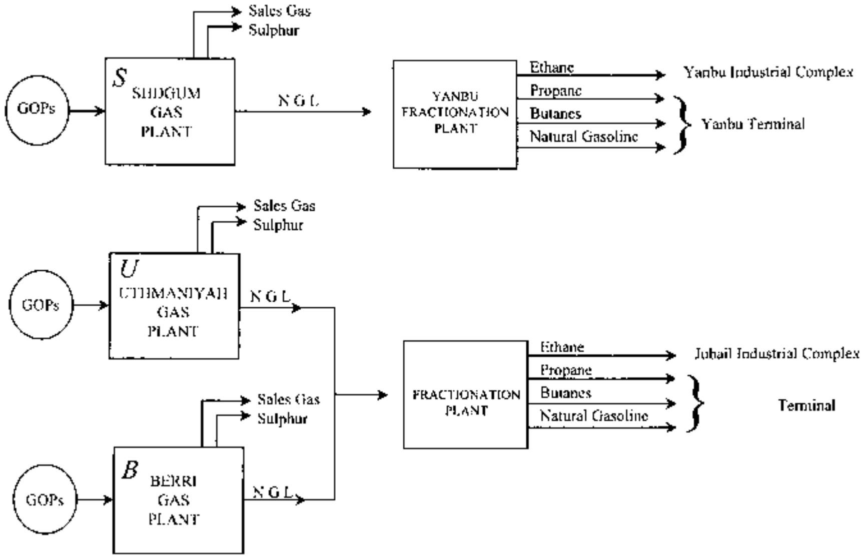


Figure 2 Master gas system (MGS).

Field processing of natural gas, in general, is carried out for two main reasons:

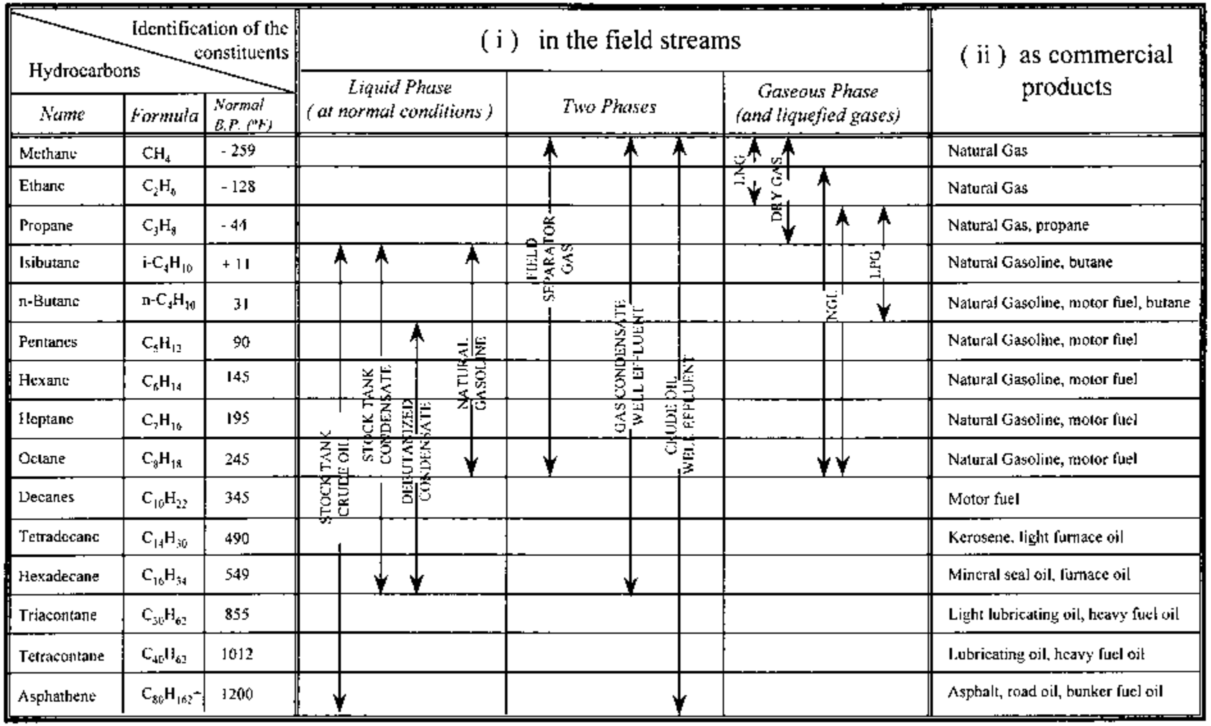
1. The necessity to remove impurities from the gas
2. The desirability of increasing liquid product recovery above that obtained by conventional gas processing

Gas field processing units usually include the following:

1. Removal of water vapor: *Gas dehydration* represents the most common needed unit in gas field processing. Water vapor seems to be a universal impurity in natural gas. Water vapor is not objectionable as such; however, the liquid or solid phase of water is very troublesome. This is experienced when the gas is compressed or cooled.

   The liquid from of water accelerates corrosion of pipelines and other related equipment. It also reduces pipeline capacity, because it accumulates in low-point regions. Solid hydrates, on the other hand, plug valves, fittings, and in some cases the pipe itself.
2. Acid gas separation: Acid gases basically imply hydrogen sulfide and carbon dioxide (H_2S and CO_2). Both gases are considered impurities that must be removed from raw natural gas before it can be marketed. Hydrogen sulfide is extremely toxic; when

Table 1 Constituents of Crude Oil and Associated Gases*



Hydrocarbons / Identification of the constituents			(i) in the field streams			(ii) as commercial products
Name	*Formula*	*Normal B.P. (°F)*	*Liquid Phase (at normal conditions)*	*Two Phases*	*Gaseous Phase (and liquefied gases)*	
Methane	CH_4	- 259		FIELD SEPARATOR GAS GAS CONDENSATE WELL EFFLUENT CRUDE OIL WELL EFFLUENT	LNG DRY GAS	Natural Gas
Ethane	C_2H_6	- 128			NGL	Natural Gas
Propane	C_3H_8	- 44			LPG	Natural Gas, propane
Isibutane	$i\text{-}C_4H_{10}$	+ 11	STOCK TANK CRUDE OIL STOCK TANK CONDENSATE NATURAL GASOLINE			Natural Gasoline, butane
n-Butane	$n\text{-}C_4H_{10}$	31				Natural Gasoline, motor fuel, butane
Pentanes	C_5H_{12}	90	DEBUTANIZED CONDENSATE			Natural Gasoline, motor fuel
Hexane	C_6H_{14}	145				Natural Gasoline, motor fuel
Heptane	C_7H_{16}	195				Natural Gasoline, motor fuel
Octane	C_8H_{18}	245				Natural Gasoline, motor fuel
Decanes	$C_{10}H_{22}$	345				Motor fuel
Tetradecane	$C_{14}H_{30}$	490				Kerosene, light furnace oil
Hexadecane	$C_{16}H_{34}$	549				Mineral seal oil, furnace oil
Triacontane	$C_{30}H_{62}$	855				Light lubricating oil, heavy fuel oil
Tetracontane	$C_{40}H_{62}$	1012				Lubricating oil, heavy fuel oil
Asphathene	$C_{80}H_{162}{}^{+}$	1200				Asphalt, road oil, bunker fuel oil

***LPG = Liquefied Petroleum Gases.**
NGL= Natural Gas Liquids (normally C_3^+).
LNG= Liquefied Natural Gas.

combusted, it produces sulfur oxides, which are a nuisance to consumers. Both H_2S and CO_2 are corrosive, especially in the presence of water. Once removed, H_2S could be commercially utilized to produce sulfur.

3. Heavy hydrocarbon separation: It is desirable to remove hydrocarbons heavier than methane from natural gas, especially for fuel gasses. Heavier hydrocarbons, specifically C_3^+ tend to condense, forming two-phase flow and thus creating pipeline operating problems.

Irrespective of the source of natural gas, the final specifications set for market sales requirements are usually the following [1]:

H_2S	0.25–0.3 grain per 100ft^3 (one grain = 64.799 mg)
Total sulfur	20 grains per 100 ft^3
Oxygen (air)	0.2% by volume
Carbon dioxide	2% by volume
Liquefiable hydrocarbons	0.2 gal per 1000 ft^3
Water content	7 lbs/MMSCF (in a 1000-psia gas line)
Thermal heating value	1150 Btu/ft^3

Identification of the hydrocarbon constituents of crude oils and associated natural gas along with their corresponding commercial products are summarized in Table 1 [1].

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1

Oil and Gas

From Formation to Production

1.1 INTRODUCTORY REMARKS

This book deals with the field surface operations and facilities for handling and processing the produced oil, gas, and water. This is an important aspect in the overall planning and development of the field and must be considered and integrated into the early stages of planning, economic evaluation, and development of the field. Before considering the surface production operations and facilities, however, extensive work and studies are first made to characterize and evaluate the reservoir, determine the production strategy for the life of the field, design the well completions that are compatible with the production strategy, and design the well-drilling programs.

Petroleum engineering students would, normally, have covered the subject matters related to the reservoir, well completion, drilling, and subsurface production methods before taking a course on surface production operations. Non-petroleum-engineering students taking this course, however, would be lacking such important and useful knowledge and background. This chapter, therefore, provides brief background information for the non-petroleum-engineering students to appreciate all operations related to the production of oil and gas.

A brief description of how oil and gas were formed and accumulated underground is first presented. A description of the various types of petroleum reservoir according to their geologic and production classifications is then provided. The exploration activities used in finding (discovering) petroleum reservoirs are then highlighted. Finally, an overview of the field development work, including the reservoir, drilling, and production engineering aspects of the development, are summarized.

1.2 FORMATION AND ACCUMULATION OF OIL AND GAS

1.2.1 Formation of Oil and Gas

Several theories have been proposed to explain the formation and origin of oil and gas (petroleum); these can be classified as the organic theory of petroleum origin and the inorganic theory of origin. The organic theory provides the explanation most accepted by scientists and geologists.

It is believed, and there is evidence, that ancient seas covered much of the present land area millions of years ago. The Arabian Gulf and the Gulf of Mexico, for example, are parts of such ancient seas. Over the years, rivers flowing down to these seas carried large volumes of mud and sedimentary materials into the sea. The mud and sedimentary materials were distributed and deposited layer upon layer over the sea floor. The buildup of thousands of feet of mud and sediment layers caused the sea floors to slowly sink and be squeezed. This eventually became the sedimentary rocks (the sandstones and shales, and the carbonates) where petroleum is found today.

The very large amount of small plant and animal life, which came into the sea with river mud and sedimentary materials, and the much larger amount of small marine life remains already on the sea floors constituted the source of petroleum. These small organisms died and were buried by the depositing silt and, thus, were protected from ordinary decay. Over many years, pressure, temperature, bacteria, and other reactions caused these dead organisms to change into oil and gas. The gas was formed under the higher-temperature conditions, whereas the oil was formed under the lower-temperature conditions. The rocks where oil and gas were formed are known as the *source rock*.

1.2.2 Accumulation of Oil and Gas

The oil, gas, and salt water occupied the pore spaces between the grains of the sandstones, or the pore spaces, cracks, and vugs of the limestones and dolomites. Whenever these rocks were sealed by a layer of impermeable rock, the *cap rock*, the petroleum accumulating within the pore spaces of the source rock was trapped and formed the petroleum reservoir. However, when such conditions of trapping the petroleum within the source rocks did not exist, oil gas moved (*migrated*), under the effects of pressure and gravity, from the source rock until it was trapped in another capped (sealed) rock.

Because of the differences in density, gas, oil, and water segregated within the trap rock. Gas, when existed, occupied the upper part of the trap and water occupied the bottom part of the trap, with the oil between the gas and water. Complete displacement of water by gas, or oil, never occurred. Some salt water stayed with the gas and/or oil within the pore spaces and as a film covering the surfaces of the rock grains; this water is known as the *connate water*, and it may occupy from 10% up to 50% of the pore volume.

The geologic structure in which petroleum has been trapped and has accumulated, whether it was the source rock or the rock to which petroleum has migrated, is called the *petroleum reservoir*.

In summary then, the formation of a petroleum reservoir involves first the accumulation of the remains of land and sea life and their burial in the mud and sedimentary materials of ancient seas. This is followed by the decomposition of these remains under conditions that recombine the hydrogen and carbon to form the petroleum mixtures. Finally, the formed petroleum is either trapped within the porous source rock when a cap rock exists or it migrates from the source rock to another capped (sealed) structure.

1.3 TYPES OF PETROLEUM RESERVOIR

Petroleum reservoirs are generally classified according to their geologic structure and their production (*drive*) mechanism.

1.3.1 Geologic Classification of Petroleum Reservoirs

Petroleum reservoirs exist in many different sizes and shapes of geologic structures. It is usually convenient to classify the reservoirs according to the conditions of their formation as follows:

1. *Dome-Shaped and Anticline Reservoirs:* These reservoirs are formed by the folding of the rock layers as shown in Figure 1. The dome is circular in outline, and the anticline is long and narrow. Oil and/or gas moved or migrated upward through the porous strata where it was trapped by the sealing cap rock and the shape of the structure.
2. *Faulted Reservoirs:* These reservoirs are formed by shearing and offsetting of the strata (*faulting*), as shown in Figure 2. The movement of the nonporous rock opposite the porous formation containing the oil/gas creates the sealing. The tilt of the petroleum-bearing rock and the faulting trap the oil/gas in the reservoir.

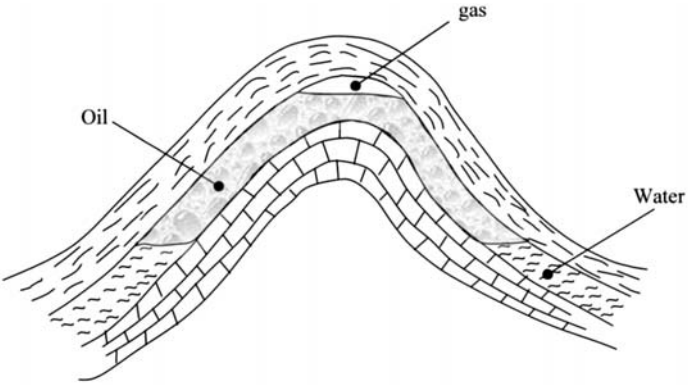


Figure 1 A reservoir formed by folding of rock layers.

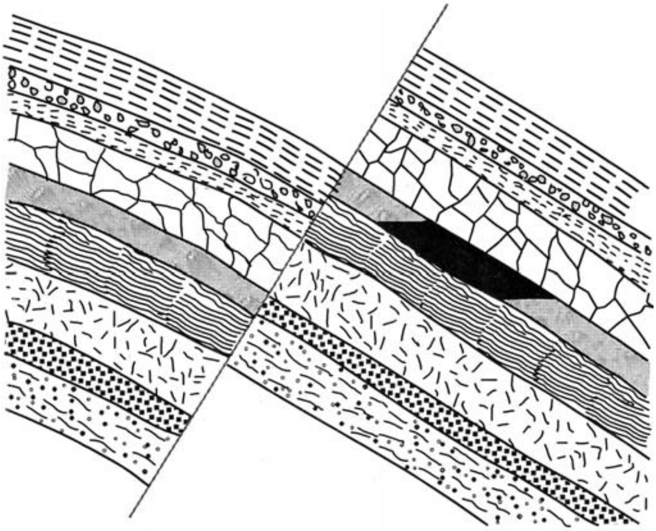


Figure 2 A cross section of a faulted reservoir.

3. *Salt-Dome Reservoirs:* This type of reservoir structure, which takes the shape of a dome, was formed due to the upward movement of large, impermeable salt dome that deformed and lifted the overlying layers of rock. As shown in Figure 3, petroleum is trapped between the cap rock and an underlying impermeable rock layer, or between two impermeable layers of rock and the salt dome.

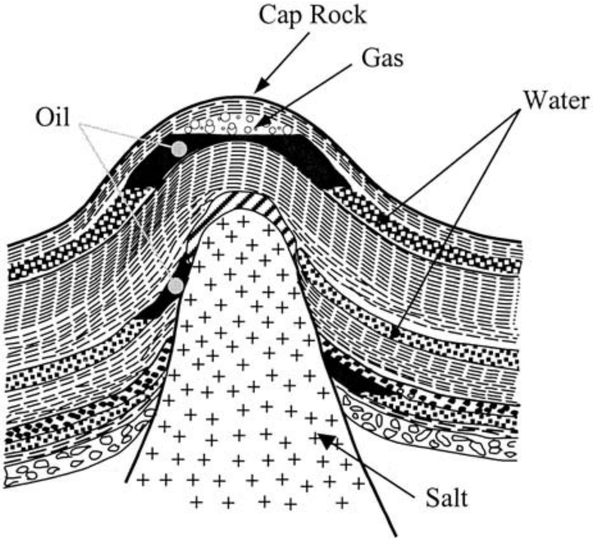


Figure 3 Section in a salt-dome structure.

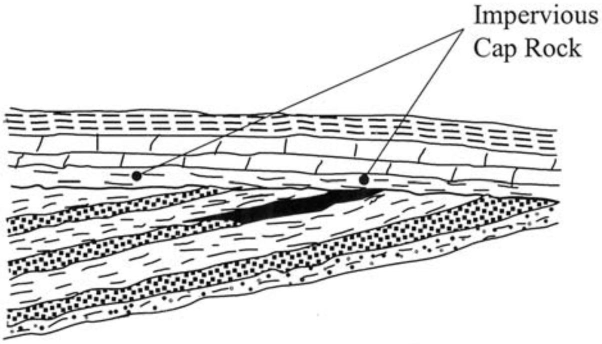


Figure 4 A reservoir formed by unconformity.

4. *Unconformities:* This type of reservoir structure, shown in Figure 4, was formed as a result of an unconformity where the impermeable cap rock was laid down across the cutoff surfaces of the lower beds.
5. *Lense-Type Reservoirs:* In this type of reservoir, the petroleum-bearing porous formation is sealed by the surrounding, nonporous formation. Irregular deposition of sediments and shale at the time the formation was laid down is the probable

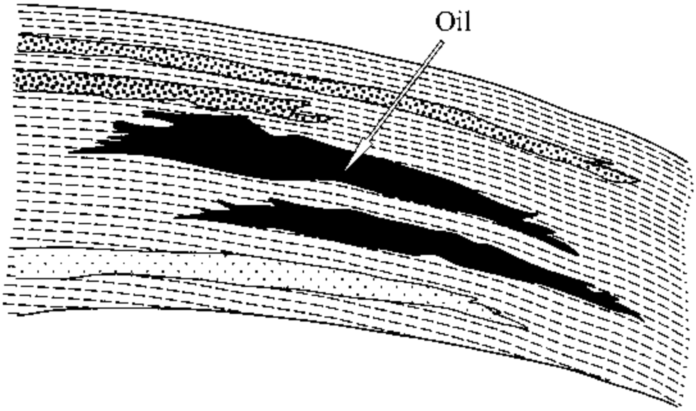


Figure 5 An example of a sandstone lense-type reservoir.

cause for this abrupt change in formation porosity. An example of this type of reservoirs is shown in Figure 5.

6. *Combination Reservoirs:* In this case, combinations of folding, faulting, abrupt changes in porosity, or other conditions that create the trap, from this common type of reservoir.

1.3.2 Reservoir Drive Mechanisms

At the time oil was forming and accumulating in the reservoir, the pressure energy of the associated gas and water was also stored. When a well is drilled through the reservoir and the pressure in the well is made to be lower than the pressure in the oil formation, it is that energy of the gas, or the water, or both that would displace the oil from the formation into the well and lift it up to the surface. Therefore, another way of classifying petroleum reservoirs, which is of interest to reservoir and production engineers, is to characterize the reservoir according to the production (*drive*) mechanism responsible for displacing the oil from the formation into the wellbore and up to the surface. There are three main drive mechanisms:

I. *Solution-Gas-Drive Reservoirs:* Depending on the reservoir pressure and temperature, the oil in the reservoir would have varying amounts of gas dissolved within the oil (*solution gas*). Solution gas would evolve out of the oil only if the pressure is lowered below a certain value, known as the *bubble point pressure*, which is a property of the oil. When a well is drilled through the reservoir and the pressure conditions are controlled to create a pressure that is lower than the bubble point pressure, the liberated gas expands and drives the oil out of the formation and assists in lifting it to the surface. Reservoirs with the energy of the escaping and expanding dissolved gas as the only source of energy are called *solution-gas-drive reservoirs*. This drive mechanism is the least effective of all drive mechanisms; it generally yields recoveries between 15% and 25% of the oil in the reservoir.

II. *Gas-Cap-Drive Reservoirs:* Many reservoirs have free gas existing as a gas cap above the oil. The formation of this gas cap was due to the presence of a larger amount of gas than could be dissolved in the oil at the pressure and temperature of the reservoir. The excess gas is segregated by gravity to occupy the top portion of the reservoir.

In such a reservoirs, the oil is produced by the expansion of the gas in the gas cap, which pushes the oil downward and fills the pore spaces formerly occupied by the produced oil. In most cases, however, solution gas is also contributing to the drive of the oil out of the formation. Under favorable conditions, some of the solution gas may move upward into the gas cap and, thus, enlarge the gas cap and conserves its energy. Reservoirs produced by the expansion of the gas cap are known as *Gas-cap-drive reservoirs*. This drive is more efficient than the solution-gas drive and could yield recoveries between 25% and 50% of the original oil in the reservoir.

III. *Water-Drive Reservoirs:* Many other reservoirs exist as huge, continuous, porous formations with the oil/gas occupying only a small portion of the formation. In such cases, the vast formation below the oil/gas is saturated with salt water at very high pressure. When oil/gas is produced, by lowering the pressure in the well opposite the petroleum formation, the salt water expands and moves upward, pushing the oil/gas out of the formation and occupying the pore spaces vacated by the produced oil/gas. The movement of the water to displace the oil/gas retards the decline in oil, or gas pressure, and conserves the expansive energy of the hydrocarbons.

Reservoirs produced by the expansion and movement of the salt water below the oil/gas are known as *water-drive reservoirs*. This is the most efficient drive mechanism; it could yield recoveries up to 50% of the original oil.

1.4 FINDING OIL AND GAS EXPLORATION

As explained in the previous sections, oil and gas exist in reservoirs located thousands of feet below the Earth's surface and ocean floors. These reservoirs would exist only in certain locations depending on the geologic history of the Earth. Therefore, determining the location of petroleum reservoirs is a very difficult task and is, probably, the most challenging aspect of the petroleum industry. Finding, or discovering, a petroleum reservoir involves three major activities: *geologic surveying*, *geophysical surveying*, and *exploratory drilling* activities. The following subsections provide a brief background on each of these activities.

1.4.1 Geologic Survey

Geologic surveying is the oldest and first used tool for determining potential locations where there would be a possibility of finding underground petroleum reservoirs. It involves examination of the surface geology, formation outcrops, and surface rock samples. The collected information is used in conjunction with geologic theories to determine whether petroleum reservoirs could be present underground at the surveyed location. The results of the geologic survey are not conclusive and only offer a possibility of finding petroleum reservoirs. The rate of success of finding petroleum reservoirs using geologic surveys alone has been historically low. Currently, geologic surveys are used together with other geophysical surveys to provide higher rates of success in finding petroleum reservoirs.

1.4.2 Geophysical Surveys

There are mainly four types of geophysical surveys used in the industry: *gravity survey*, *magnetic survey*, *seismic survey*, and *remote sensing*.

The *gravity survey* is the least expensive method of locating a possible petroleum reservoir. It involves the use of an instrument, a *gravimeter*, which picks up a reflection of the density of the subsurface rock. For example, because salt is less dense than rocks, the gravimeter can detect the presence of salt domes, which would indicate the presence of an anticline structure. Such a structure is a candidate for possible accumulation of oil and gas.

The *magnetic survey* involves measurement of the magnetic pull, which is affected by the type and depth of the subsurface rocks. The magnetic survey can be used to determine the existence and depth of subsurface volcanic formations, or basement rocks, which contain high concentrations of magnetite. Such information is utilized to identify the presence of sedimentary formations above the basement rocks.

The *seismic survey* involves sending strong pressure (sound) waves through the earth and receiving the reflected waves off the various surfaces of the subsurface rock layers. The sound waves are generated either by using huge land vibrators or using explosives. The very large amount of data collected, which include the waves' travel times and characteristics, are analyzed to provide definitions of the subsurface geological structures and to determine the locations of traps that are suitable for petroleum accumulation.

This type of survey is the most important and most accurate of all of the geophysical surveys. Significant technological developments in the field of seismic surveying have been achieved in recent years. Improvements in the data collection, manipulation, analysis, and interpretation have increased the significance and accuracy of seismic surveying. Further, the development of three-dimensional (3D) seismic surveying technology has made it possible to provide 3D descriptions of the subsurface geologic structures.

Remote sensing is a modern technique that involves using infrared, heat-sensitive, color photography to detect the presence of underground mineral deposits, water, faults, and other structural features. The sensing device, normally on a satellite, feeds the signals into special computers that produces maps of the subsurface structures.

1.4.3 Exploratory Drilling

The data collected from the geologic and geophysical surveys are used to formulate probable definitions and realizations of the geologic structure that may contain oil and/or gas. However, we still have to determine whether petroleum exists in these geologic traps, and if it does exist, would it be available in such a quantity that makes the development of the oil/gas field economical? The only way to provide a definite answers is to drill and test exploratory well(s).

The exploratory well, known as the *wildcat well*, is drilled in a location determined by the geologists and geophysicists. The well is drilled with insufficient data available about the nature of the various rock layers that will be drilled or the fluids and pressures that may exist in the various formations. Therefore, the well completion and the drilling program are usually overdesigned to assure safety of the operation. This first well, therefore, does not represent the optimum design and would probably cost much more than the rest of the wells that will be drilled in the field.

As this exploratory well is drilled, samples of the rock cuttings are collected and examined for their composition and fluid content. The data are used to identify the type of formation versus depth and to check on the presence of hydrocarbon materials within the rock. Cores of the formations are also obtained, preserved, and sent to specialized laboratories for analysis. Whenever a petroleum-bearing formation is drilled, the well is tested while placed on controlled production. After the well has been drilled, and sometimes at various intervals during drilling, various *logs* are taken. There are several logging tools, or techniques, (*electric logs*, *radioactivity logs*, and *acoustic logs*) that are used to gather information about the drilled formations. These tools are lowered into the well on a *wireline* (electric cable) and, as they are lowered, the measured signals are transmitted to the surface and recorded on computers. The signals collected are interpreted and produced in the form of rock and fluid properties versus depth.

The exploratory well will provide important data on rock and fluid properties, type and saturation of fluids, initial reservoir pressure, reservoir productivity, and so forth. These are essential and important data and information, which are needed for the development of the field. In most situations, however, the data provided by the exploratory well will not be sufficient. Additional wells may need to be drilled to provide a better definition of the size and characteristics of the new reservoir. Of course, not every exploratory well will result in a discovery. Exploratory wells may result in hitting dry holes or they may prove the reservoir to be an uneconomical development.

1.5 DEVELOPMENT OF OIL AND GAS FIELDS

The very large volume of information and data collected from the various geologic and geophysical surveys and the exploratory wells are used to construct various types of map. *Contour maps* are lines drawn at regular intervals of depth to show the geologic structure relative to reference points called the *correlation markers*. *Isopach maps* illustrate the variations in thickness between the correlation markers. Other important maps such as *porosity maps*, *permeability maps*, and maps showing variations in rock characteristics and structural arrangements are also produced. With all data and formation maps available, conceptual models describing the details of the structure and the location of the oil and gas within the structure are prepared.

The data available at this stage will be sufficient to estimate the petroleum reserves and decide and plan for the development of the field for commercial operation.

The development of petroleum fields involves the collective and integrated efforts and experience of many disciplines. Geologists and geophysicists are needed, as described earlier, to define, describe, and characterize the reservoir. Reservoir engineers set the strategy for producing the petroleum reserves and managing the reservoir for the life of the field. Production and completion engineers design the well completions and production facilities to handle the varying production methods and conditions, and drilling engineers design the well-drilling programs based on well-completion design. In the past, each group used to work separately and deliver its product to the next group. That is, when geologists and geophysicists finish their work, they deliver the product to the reservoir-engineering group. Then, reservoir engineering would deliver the results of their work to production engineering, and so on. In almost all cases, it was necessary for each group to go back to the previous group for discussion, clarification, or requesting additional work. This has been realized as a very inefficient operation. In recent years, most major companies have adopted what is known as the *multidisciplinary team approach* for field developments. In this approach, a team consisting of engineers and scientists covering all needed disciplines is formed. The team members work together as one group throughout the field development stage. Of course, other specialists such as computer scientists, planners, cost engineers, economists, and so forth work closely with the team or may become an integral part of the team. Experience has shown that this field development approach is very efficient; more and more companies are moving in this direction.

The following sections provide brief descriptions of the roles and functions of drilling, reservoir, and production engineering.

1.6 DRILLING ENGINEERING AND OPERATIONS

Following the preparation stage of field development (i.e., setting the production strategy, determining the locations of the wells in the field, and designing the well completions), the drilling-related activities begin. The drilling program is first designed. Then, plans are prepared and executed to acquire the required equipment and materials. The drilling sites in the field are then prepared for the equipment and materials to be moved in, and the drilling operations begin. Depending on the organization of activities within the oil company, drilling engineers may only be responsible for drilling and casing of the well, and production engineers will be responsible for completion of the well. Alternatively, drilling engineers may be responsible for drilling and completion of the wells.

The drilling program consists of three main stages: (1) drilling the hole to the target depth, (2) setting the various casings, and (3) cementing the casing.

1.6.1 Drilling the Well

Well drilling has gone through major developments of drilling methods to reach the modern method of rotary drilling. In this method, a *drilling bit* is attached to the bottom end of a string of pipe joints known as the *drilling string*. The drilling string is rotated at the surface, causing rotation of the drilling bit. The rotation of the bit and the weight applied on it through the drilling string causes the crushing and cutting of the rock into small pieces (*cuttings*). To remove the cuttings from the hole, a special fluid, called the *drilling fluid* or the *drilling mud*, is pumped down through the drilling string, where it exists through nozzles in the bit as jets of fluid. This fluid cleans the bit from the cuttings and carries the cuttings to the surface through the annular space between the drilling string and the wall of the hole. At the surface, the mud is screened to remove the cuttings and is circulated back into the drilling string. The drilling operation is performed using huge and complex equipment known as the *drilling rig*. This is briefly described next.

The Drilling Rig

Figure 6 shows a schematic of a rotary drilling rig. It consists of two main sections, the *substructure* (bottom section) and the *derrick* (top section). The substructure, which ranges from 15 to 30 ft in height, is basically a rigid platform that supports the derrick. The rig is composed of several systems and components; the major systems are as follows:

1. *Power System:* It consists of diesel-engine-driven electric generators to supply electric power to the various systems and components of the rig. About 85% of the power generated is consumed by the drilling-mud circulating system.
2. *Hoisting System:* The function of this system is to lower and raise the drilling string into and out of the well. The main components of the system are the *crown block*, the *drawworks*, the *traveling block*, the *hook*, the *swivel*, the *elevator*, the *drilling line*, and the *dead-line anchor*.
3. *Rotating System:* The system consists mainly of a motor-driven rotating table that is used to rotate a pipe with a square or hexagonal cross section, called the *kelly*, while allowing it to slide

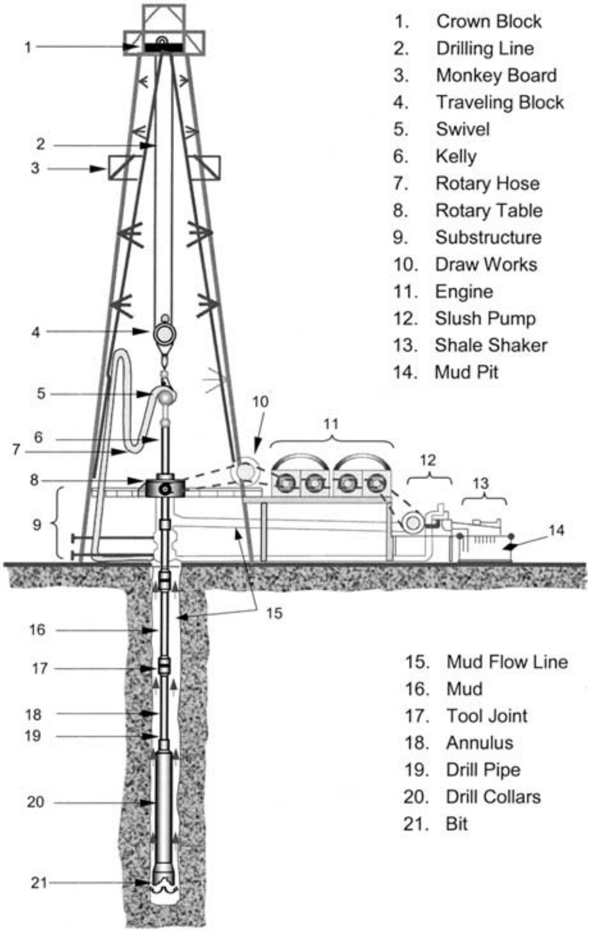


Figure 6 A schematic of a rotary drilling rig.

through. The kelly is suspended by the hoisting system and is connected at its bottom to the drill pipe and the bit.

4. *Circulating System:* The system consists mainly of drilling-mud storage tanks, high-pressure pumps, circulating hoses and pipes, and a shale shaker. Its function is to circulate the mud through the well to bring the cuttings to the surface.
5. *Drilling Bit:* The drilling bit is the device that does the actual drilling by crushing and cutting the rock as it rotates with some force applied by the drilling string. There are different types and shapes used for different types of rock.

Drilling Fluid (Mud)

The drilling fluid is a very important element of the drilling operation. Its importance stems from the many essential functions it serves. Some of these functions are as follows:

1. Transporting the cuttings from the bottom of the hole to the surface
2. Cooling of the bit and lubrication of the drill string
3. Exerting hydrostatic pressure to overbalance the pressure of the formation and thus prevent flow of formation fluids into the well
4. Supporting the walls of the hole to prevent it from caving in
5. Enhancing drilling by its jetting action through the bit nozzles

The drilling fluid can be prepared in different formulations to provide the desired properties (density, viscosity, and filtration) under the bottom hole conditions. The basic drilling fluid consists of water and clay (*water-base mud*). Other materials and chemicals are also added to control the properties of the fluid. Other fluids such as foam and air have also been used in drilling operations.

1.6.2 Casing the Well

The casing is a steel pipe that is placed in the drilled hole (well) to support the wall of the hole and prevent it from collapsing. When cemented to the wall, it seals the subsurface formation layers and prevents communications between the various layers.

Normally, four strings of casing of different diameters are installed in the well at various depths that are specified by the geologist. These are the *conductor*, the *surface casing*, the *intermediate casing* and the *production casing*. The conductor has the largest diameter and shortest length of the four casing strings; the production casing has the smallest diameter and longest casing.

Casings of various outside diameters are available in different grades and weights. The grade refers to the type of casing steel alloy and its minimum yield strength. Commonly available grades are H-40, J-55, N-80, C-75, L-80, and P-105. The letter (H, J, etc.) identifies the type of alloy and heat treatment; the number (40, 55, etc.) refers to the minimum yield strength in thousands of pounds per square inch (psi). For a given outside diameter and grade, casings are available in different weights (i.e., various inside diameters) expressed in pounds per linear foot of casing. The weight and grade of the casing specify its resistance to various loads such as burst, collapse, and tension loads. In designing casing strings, weight and grade must be selected such that the casing string will not fail under all loads to which it will be subjected during drilling, setting casing, and production.

To set a casing string, the drilling operation is stopped when the desired depth is reached and the drill string and the pit are pulled out of the hole. The casing string is then lowered into the hole, joint by joint, using the hoisting system of the rig, until the total length of casing is in the hole. A round, smooth object called the *guide shoe* is attached to the bottom of the first joint of casing to ease and guide the movement of the casing into the hole. The casing is then cemented to the wall and drilling operation is resumed until the target depth for the next casing string is reached. Normally, before setting the production casing, the petroleum formation is logged and evaluated. The casing will be set only if the logging results indicate the presence of a productive formation. Otherwise, the well will be abandoned.

1.6.3 Cementing the Casing

To cement the casing, the annulus between the casing and the wall of the hole must be filled with cement. To achieve this, the required volume of the cement slurry (prepared on location) is pumped through the casing. A special rubber plug is normally inserted ahead of the cement to separate it from the mud and prevent any contamination of the cement with mud. Another plug is inserted after pumping the specified volume of cement. This is followed by pumping a fluid (normally mud) to displace the cement. When the first plug reaches the bottom, pumping pressure is increased to rupture the plug and allow the flow of cement behind the casing. When the top plug reach the bottom, the cement must have filled the annulus to the surface. The cement is then left undisturbed until it sets and acquires enough compressive strength before resuming the drilling operation for the next casing string.

Once the casing is cemented, it becomes permanently fixed into the hole. It is very important to have a good cement bond between the casing and the wall of the hole. For this purpose, a special log (*cement bond log*) is conducted to check the integrity of the cementing operation. Failure to have a good cementing job will necessitate expensive remedial cementing operations.

1.7 RESERVOIR ENGINEERING ROLE AND FUNCTIONS

Reservoir engineers play a major role in field development and operation. Some of the major functions of reservoir engineering are discussed next.

1.7.1 Estimation of Reserves

Estimation of oil and gas reserves in a discovered reservoir is one of the most important factors in evaluating the discovery and deciding on its viability for commercial development. To determine the volume of oil and/or gas present in the reservoir, the bulk volume of the reservoir (V_b) is first determined using the available reservoir description data. The volume of fluids in the pore spaces of the reservoir rock is then calculated by multiplying the bulk volume by the rock porosity (φ); this is also known as the *pore volume* of the rock. *Porosity* is a property of the rock defined as the ratio of the volume of the pore spaces within the rock to the bulk volume of the rock. The pore volume is normally occupied by oil (or gas) and water. The fraction of the pore volume occupied by water is known as the *water saturation* (S_w). The porosity and initial water saturation are determined from the logs and core samples obtained from the exploratory wells. Therefore, the initial volume of oil (V_0) at the reservoir conditions is determined by

$$V_0 = V_b \varphi (1 - S_w)$$

This volume of oil is called the initial oil in place (IOIP) or the original oil in place (OOIP). This method of estimating the OOIP is known as the volumetric method. Other methods (the material balance method and decline curve method) also exist for estimating the OOIP.

It is impossible to recover all of the OOIP; certain forces within the reservoir rock prevent the movement of some oil from the rock to the well. The fraction of the OOIP that could be recovered is called the recovery factor (E_r), and the total recoverable volume of oil ($E_r V_0$) is called the *proven reserves*.

The *proven reserve* for a reservoir changes with time as a result of three factors. First, the volume of oil in place decreases as oil is produced from the reservoir. Second, as more oil is produced, more reservoir data become available, which could change or modify the initial estimate of the OOIP. Third, new developments and improvements in recovery and production methods may increase the possible recovery factor. Therefore, the proven reserve of any field is continuously updated.

1.7.2 Well Location, Spacing, and Production Rates

Another important function of reservoir engineering is to determine the optimum locations of the wells to be drilled and the production rate from each well for the most effective depletion of the reservoir. Where the wells should be drilled depends largely on the structural shape of the reservoir and the reservoir drive mechanism. As a general rule, no wells should be drilled into locations of the reservoir where water or gas is expected to invade that part of the reservoir early after starting production. Figures 7–10 illustrate the preferred locations of wells for various reservoir structures and drive mechanisms.

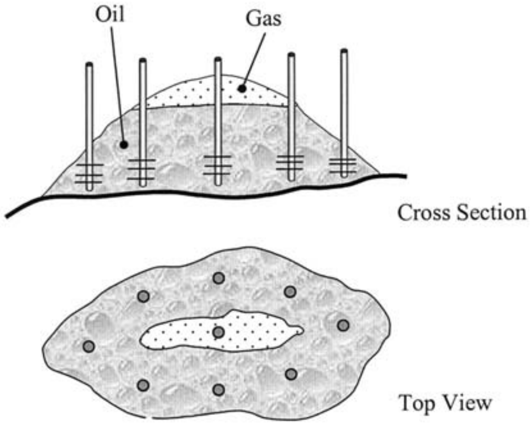


Figure 7 Gas-cap drive reservoir: completion near bottom of oil zone.

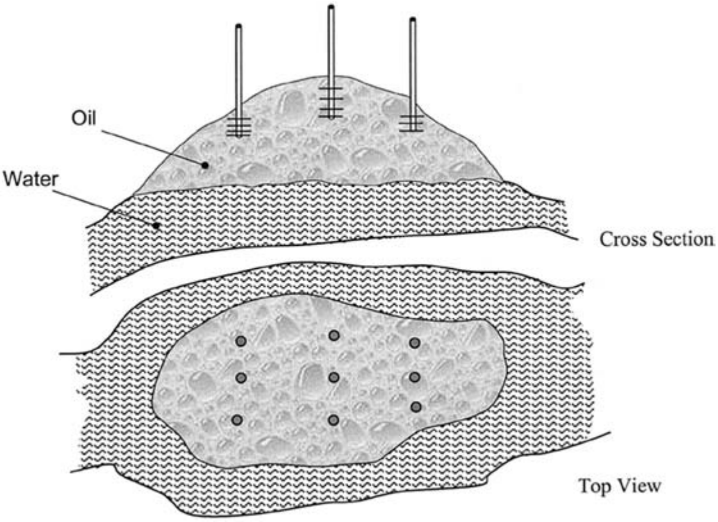


Figure 8 Bottom-water drive reservoir: completion near top of oil zone.

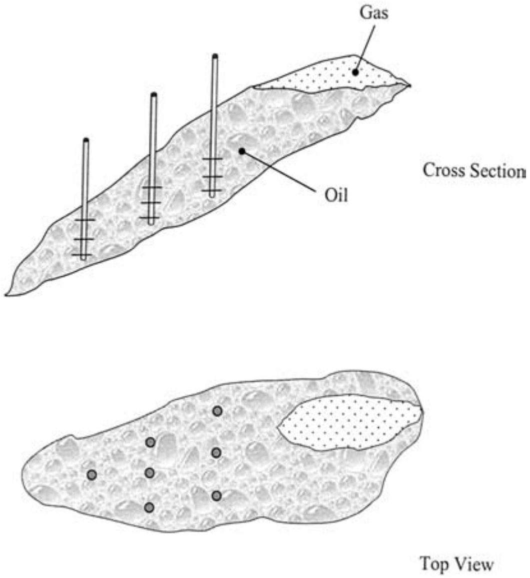


Figure 9 Gas-cap drive, high angle of dip: wells located at the lower section of the structure.

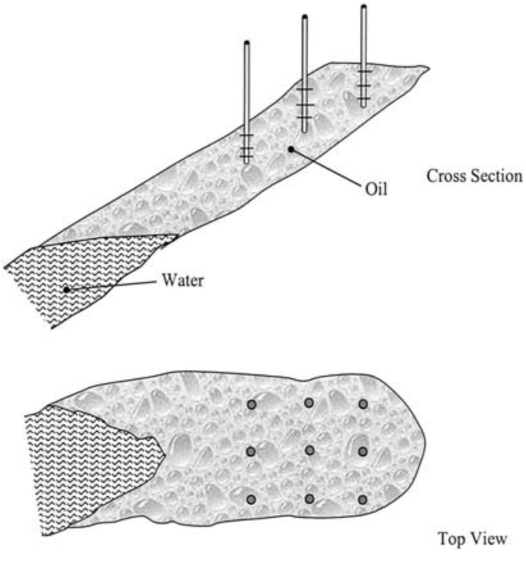


Figure 10 Water drive, high angle of dip: wells located near the upper section of the structure.

The spaces between wells and the production rate from each well are determined with the objective of recovering the maximum volume of petroleum in the most economical way. This is not an easy task and requires extensive reservoir simulation studies. Such studies continue after the development stage and, as more data become available, there studies may result in having to drill additional wells (*infill wells*) for higher recovery.

1.7.3 Reservoir Simulation

The complexity of the reservoir structures and the large number of nonlinear equations governing the flow of fluids in the reservoir make it impossible to obtain analytical solutions to the fluid flow problem. Reservoir simulators implementing various numerical techniques are constructed for solving such a problem. Running a simulator requires inputting of many parameters that are related to the reservoir rock and fluids. The majority of these parameters may either be known with various degrees of certainty or may have to be assumed. Therefore, the accuracy of the simulation results depends, first, on the accuracy of the input data.

Simulation studies are first conducted during the development stage to predict and evaluate the performance of the reservoir for the life of the field. Once production starts, data are collected and the actual reservoir performance is checked against the simulation-predicted performance. In almost all cases, the actual and predicted results do not match. Therefore, reservoir (simulation) engineers modify the simulation by changing some or all of the input data using either newly acquired data or making new assumptions, or both. This is done until a match between the simulation results and the actual results, obtained from the producing field, is obtained. This process is known as *history matching*. The modified simulator is then used to predict the field performance and direct the production strategy for better management of the reservoir. The whole process is repeated on a continuous basis for every specified period of production.

1.7.4 Reservoir Management

The aspects of reservoir management are numerous and involve extensive and complicated studies. In simple terms, however, the role of reservoir management is to specify specific producing strategies that will control the movements of the various fluids within the reservoir in order to achieve maximum recovery of the hydrocarbon materials while minimizing the production of undesired fluids. The production strategies may include assignment of production rates to individual wells in the field, shutting off certain wells for a specific duration, drilling of new wells at specific locations, and implementing specific pressure maintenance or improved recovery operations. The importance of reservoir management has been recognized in recent years. This led to the formation of specialized groups and departments for reservoir management in most major oil companies.

1.7.5 Improved Petroleum Recovery

The natural reservoir drive mechanisms described in Section 1.3.2 normally result in low-recovery factors. In fact, if reservoirs were produced solely by the natural driving forces, very poor recoveries would be obtained. The specific rock and fluids properties and the forces that control the movements of the fluids within the reservoir are responsible for such poor recoveries. Petroleum recovery by these natural drive mechanisms is called *primary recovery*.

In order to achieve recoveries higher than the primary recovery, we must intervene into the reservoir to artificially control, or alter, the natural driving forces and/or the rock and fluids properties. Reservoir engineers have developed, and continued research is being conducted for further developments, refinement, and improvement, and various techniques to achieve higher than primary recoveries. Some techniques involve supplementing the natural driving force by injecting high-pressure fluids into the reservoir. Other techniques aim at changing the fluids and/or rock properties to enhance the mobility of the petroleum fluids and suppress the forces that hinder their movements. All such techniques are known as *enhanced recovery methods* or *improved recovery methods*. A brief description of some of the improved recovery methods follows.

Pressure Maintenance

As oil is produced, the reservoir pressure declines at a rate that depends on the reservoir drive mechanism, the strength of that drive, and the amount of oil produced. Solution-gas-drive reservoirs experience the highest rate of pressure decline, followed by the gas-cap-drive reservoirs, with the water-drive reservoirs being the least affected. The decrease in reservoir pressure reduces the ability of the formation to produce oil. The loss of productivity becomes very severe if the reservoir pressure drops below the bubble point pressure.

One way of maintaining a high reservoir pressure to maintain productivity and increase recovery is to inject a fluid into the reservoir at such quantities and pressure that it will keep the reservoir pressure at the desired high level. Depending on the type of reservoir, pressure maintenance may be achieved by either the injection of water through wells drilled at the periphery of the reservoir or the injection of gas at the top of the reservoir.

Water Flooding

In this method of improved recovery, injection wells are drilled between the oil producing wells in a specific regular pattern. For example, for the *5-spot flooding pattern*, one injection well is drilled between and at equal distances from four producing wells. Water is injected into the injection well to drive (push) the oil toward the four producers. Simulations of the process are usually made before implementation to determine the optimum well spacing, or pattern, and the injection rate and pressure to achieve efficient displacement of oil by the injected water. Water-flooding operations are always associated with a high water cut in the produced fluid. The process continues until the oil production rate becomes too low for economic operation.

Because the water viscosity is lower than the oil viscosity, water may finger through the oil and reach the producing well prematurely, bypassing a significant amount of oil that would be very difficult to recover. Engineers have to be careful in determining the injection rate for the specific reservoir and rock and fluid properties to avoid water fingering. Water flooding could be improved if the water viscosity were altered to become higher than the oil viscosity. This would create a more favourable mobility conditions, make the oil–water interface more stable, and avoid fingering. This is achieved by adding certain polymers to the injection water. This is known as *polymer flooding*.

Chemical Recovery

Some liquids have a higher affinity than others to adhere to solid surfaces. When two liquids are in contact with a common solid surface, one of the two liquids (the one with the higher adhesion affinity) will spread over the surface at the expense of the other. The solid surface is then identified as *wettable* by that liquid. Most reservoir rocks are wetted by water than by oil and are, therefore, identified as *water-wet* rocks. Such conditions are favorable for displacing oil by water. There are situations, however, where the reservoir rocks are *oil-wet*. In these cases, water would not be able to displace the oil. Reservoir engineers and surface-chemistry scientists have developed methods in which chemicals are used to change the wettability of the rock for effective water-flooding operations. These methods are known as the *chemical recovery methods*. The two most common chemical recovery methods are as follows:

1. *Surfactant Flooding:* In this method, a slug of water–surfactant solution is first injected through the injection well into the formation. This is followed by injection of ordinary water as in regular water-flooding operations. The surfactant ahead of the floodwater causes changes in the interfacial tension and mobilizes the oil that would otherwise adhere to the surface of the rock. Thus, the displacement of the oil by the floodwater becomes possible. Again, adding polymers to the water to create favorable mobility conditions could increase the flooding effectiveness. In this case, the method can be called *surfactant–polymer flooding*.
2. *Caustic Flooding:* This method is essentially a surfactant flooding, with the surfactant being generated within the reservoir rather than being injected. The method is applicable in situations where the reservoir oil contains high concentrations of natural acids, which can react with alkaline to produce surfactants. The most common approach is to inject a slug of caustic soda (NaOH) solution ahead of the floodwater. The alkaline reacts with the acids in the oil to in situ produce surfactants. Then, the process is converted into a surfactant-flooding process. This method is less expensive than the regular surfactant-flooding process.

Miscible Recovery

In this method, a slug of a substance that is miscible in the reservoir oil is injected into the reservoir at pressures high enough to achieve good miscibility. This is then followed by water injection. The process has been used with carbon dioxide, rich natural gas, nitrogen, flue gases, and light hydrocarbon liquids as the miscible fluid. Miscible flooding could achieve very high recovery factors.

Thermal Recovery

Heavy oil reservoirs present a unique production problem. The high viscosity of the oil makes it difficult, and in some cases impossible, to produce the oil, even with the above-discussed improved recovery methods. The best method to mobilize the oil is to heat the formation in order to reduce the oil viscosity. When heating is used in recovering the oil, the recovery method is called *thermal recovery*. The three most common thermal recovery methods are as follows:

1. *Steam Stimulation:* In this method, steam is injected into the producing well for a specified period of time (normally more than a month); then, the well is shut off for another period of time (normally a few days). The injected steam heats up the surrounding formation, causing significant reduction in oil viscosity. The well is then put on production for a period of time until the oil flow declines. The process is then repeated through the same cycle of injection, shutting off, and production. This process is also known as the *Hugh and Pugh method.*
2. *Steam Flooding:* This method is similar to the water-flooding process, except that steam is used instead of water. The steam is injected into an injection well to reduce the oil viscosity while the condensed steam (hot water) displaces the oil toward the producing wells.
3. *In Situ Combustion:* In this process, air is injected into the formation through an injection well under conditions that initiate ignition of the oil within the nearby formation. The combustion zone creates a front of distilled oil, steam, and gases. Continued air injection drives the combustion front toward the producing wells. The combination of heating and displacement by the steam, gases, and condensed liquids enhances the recovery of the oil.

1.8 PRODUCTION ENGINEERING ROLE AND FUNCTIONS

Production engineers have probably the most important role in both the development and operating stages of the field. They are responsible for making the development and production strategies prepared by the reservoir engineers a reality. Production engineers are responsible for designing and installing the well completions that are capable of producing the desired volumes of oil/gas with the prescribed methods of production. They are also responsible for maintaining the wells at their best producing conditions throughout the life of the field. These two major responsibilities are classified as *subsurface production engineering*. Still, production engineers are responsible for designing, installing, operating, and maintaining all surface production facilities starting from the flow lines at the wellhead and ending with the delivery of oil and gas to the end user. This is classified as *surface production engineering*, which is the main theme of this book. Both the subsurface and surface production engineering aspects of the field are tied together and considered as one production system. The main objective of that system is to *obtain maximum recovery in the most economical and safe manner*.

In the following subsections, the functions are operations related to subsurface production engineering are briefly described.

1.8.1 Well Completion Design

The well completion is the subsurface mechanical configuration of the well which provides the passage for the produced fluids from the face of the formation to the wellhead at the surface. The well completion design is needed by the drilling engineers to properly design, plan, and execute the drilling of the well that is compatible with the well completion.

Types of Well Completion

There are three major types of well completion:

1. *Open Hole Completions:* For this type of completion, the well is drilled down to a depth that is just above the target petroleum formation. The production casing is then lowered into the well and cemented. The target formation is then drilled and is left uncased (*open*). Depending on the production rate and the properties of the produced fluids, the well may be produced through the production casing or through production tubing placed above the producing formation with a packer that provides a seal between the tubing and casing. One of the functions of the packer is to protect the casing from the produced fluids. Figure 11 illustrates this type of well completion.
2. *Cased Hole* (*Perforated) Completions:* For this type of completion, the well is drilled all the way through the producing zone and the production casing is lowered and cemented. The casing is then perforated across the producing zone to establish communication between the formation and the well, as illustrated in Figure 12. Again, depending on the producing conditions, production could be either through the casing or through a tubing.
3. *Liner Completion:* As illustrated in Figure 13, the production casing is set and cemented above the petroleum formation similar to an open hole completion. A liner (basically a smaller diameter casing) is then set and cemented across the producing formation. The linear is then perforated to establish communication between the well and the formation. In some cases, an already perforated (or slotted) liner may be set across the formation without cementing.

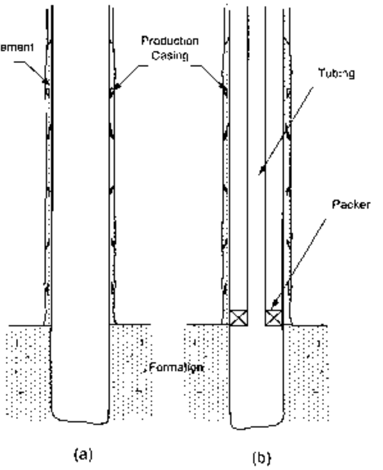


Figure 11 Open hole completion: (a) production through casing; (b) with tubing and packer.

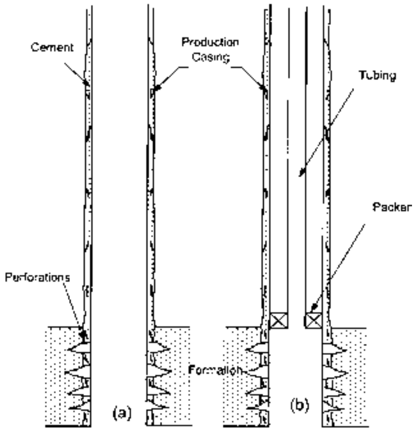


Figure 12 Cased hole completions: (a) without tubing; (b) with tubing packer.

Tubing–Casing Configurations

There are several different tubing–casing configurations. The selection of a particular configuration is normally governed by the characteristics of the reservoir and cost. The type of the reservoir and drive mechanism, the rock and fluid properties, and the need for artificial lift and improved recovery are among the technical factors which influence the selection of

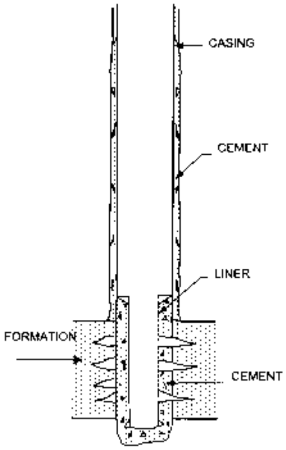


Figure 13 Liner completion.

well completion and configuration. Tubing–casing configurations can be grouped under two main categories: the *conventional* and the *tubingless* completions.

A *conventional completion* is the completion with a production tubing inside the production casing. A packer may or may not be used and the production could be through the tubing, the annulus, or both. Depending on the number of petroleum formations to be produced through the well, we may have either *single* or *multiple completion*. Figure 14 illustrates a multiple (triple) conventional completion.

The *tubingless completion* is a special type of completion where a relatively small-diameter production casing is used to produce the well without the need for production tubing. Such completions are low-cost completions and are used for small and short-life fields. Again, we may have multiple tubingless completions or single as illustrated in Figure 15.

1.8.2 Tubing Design

Tubing design involves the determination of the size, the grade, and weight of the production tubing. The size of the production tubing is controlled by the production rate, the types of flowing fluid, and the pressures at the bottom of the well and at the surface. The grade and weight of the tubing reflect its strength and are determined through analysis of the various loads that act on the tubing under all expected conditions.

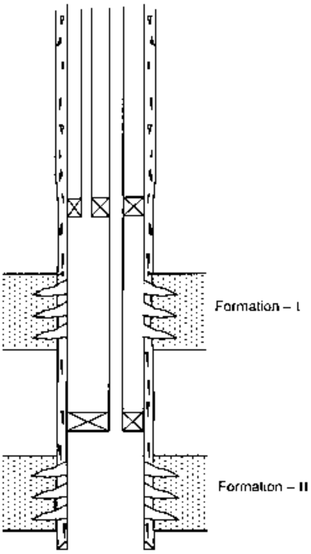


Figure 14 Multiple conventional completion.

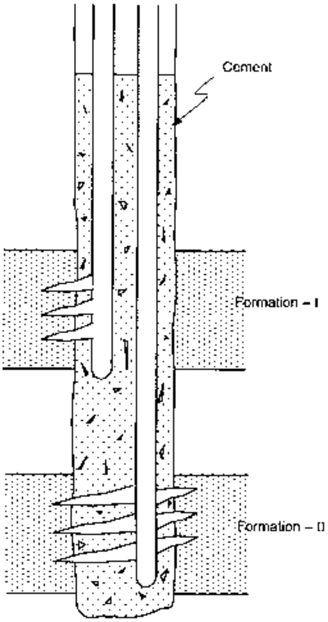


Figure 15 Multiple tubingless completion.

Determining Tubing Size

Determining the size of production tubing is the starting point for completion and drilling design, as it directly affects the sizes of all casing strings and, accordingly, the design and cost of the drilling program. The tubing size must be selected to handle the desired production rates under the varying producing conditions for the life of the well. To properly determine the tubing size, the whole production system (from the formation to the surface separator) must be considered. The ability of the formation to produce fluids by the natural drive and improved recovery methods from the start of production until depletion need to be considered. The flow of produced fluids through the production tubular, the wellhead restrictions, and the surface flow line over the life of the well need to be analyzed taking into account possible means of artificially lifting the fluids.

The flow from the formation to the bottom of the well (bottom hole) is governed by what is known as the *inflow performance relationship* of the well, whereas the flow from the bottom hole to the surface is represented by the *outflow performance relationship*.

1. *Inflow Performance Relationship (IPR):* This is the relationship between the flow rate (q) and the flowing bottom-hole pressure (P_{wf}). The relationship is linear for reservoirs producing at pressures above the bubble point pressure (i.e., when P_{wf} is greater than or equal to the bubble point pressure). Otherwise, the relationship takes the shape of a curve, as illustrated in Figure 16. When the IPR is linear, it can be represented with what is known as the *productivity index* (PI), which is the inverse of the slope of the IPR. The PI is basically the production rate per unit *drawdown* (the difference in pressures between the average reservoir static pressure, P_R, and P_{wf}). The IPR depends on the reservoir rock and fluid characteristics and changes with time, or cumulative production as illustrated in Figure 17. Methods exist for determination of the IPR and for predicting future IPRs.
2. *Outflow Performance Relationship (OPR):* Outflow performance involves fluid flow through the production tubular, the wellhead, and the surface flow line. In general, analyzing fluid flow involves the determination of the pressure drop across each segment of the flow system. This is a very complex problem, as it involves the simultaneous flow of oil, gas, and water (multiphase flow), which makes the pressure drop dependent on many variables, some of which are interdependent. There is no

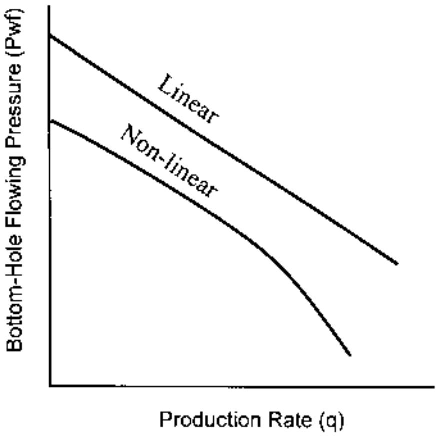


Figure 16 Inflow performance relations.

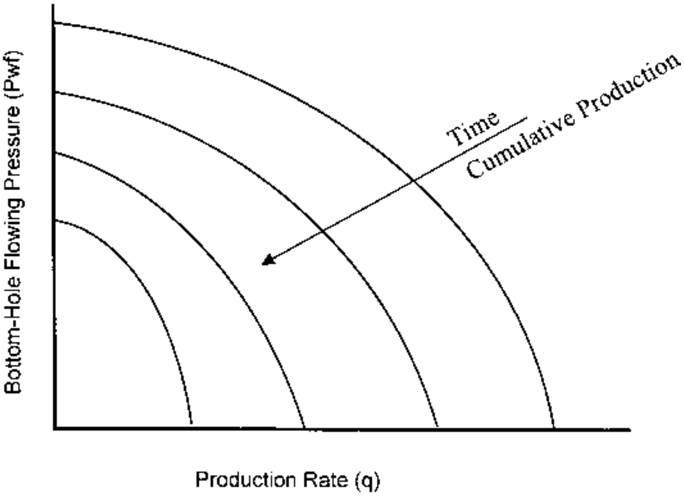


Figure 17 Dependence of the IPR on time, or cumulative production.

analytical solution to this problem. Instead, empirical correlations and mechanistic models have been developed and used for predicting the pressure drop in multiphase flow. Computer programs based on such correlations and models are now available for the determination of pressure drops in vertical, inclined, and horizontal pipes.

3. *Sizing Tubing and Flow Line:* Determination of the tubing and flow line sizes is a complex process involving the determination and prediction of future well productivity, analysis of multiphase flow under varying production conditions, and economic analysis. However, a simplied approach is summarized in the following steps and illustrated in Figure 18:

   (i) Determine (predict) the present and future IPR. Plot the results as P_{wf} versus q.
   (ii) Selecting tubing and flow line diameters and starting with a specified value for the surface separator pressure, determine the flowing bottom-hole pressure for a specified production rate, water cut (W.C.) and gas–oil ratio (GOR), using available multiphase flow correlations or models.
   (iii) Step (ii) is repeated for different values of production rates (q) with W.C. and GOR being fixed. Thus, a relation between q and P_{wf} is established and presented on the same plot of the IPR.
   (iv) Steps (ii) and (iii) are repeated for various combinations of tubing and flow line sizes.

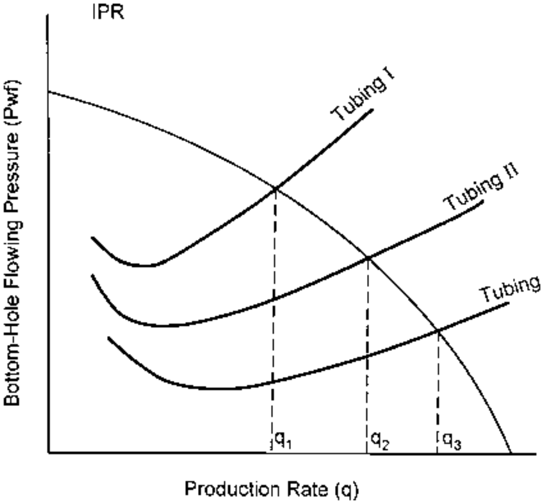


Figure 18 IPR and outflow performance for different tubings.

(v) Steps (ii)–(iv) are repeated for various expected values of W.C. and GOR, and for various producing conditions (natural flow and artificial lift)

(vi) On the P_{wf}–q plot, the intersection of the OPR curve for a particular tubing–flow line combination and producing conditions with the IPR curve of any specific time represents the maximum possible production rate and corresponding P_{wf}. If the OPR curve does not intersect with a particular IPR curve, then the well will not flow when such an IPR and producing conditions are reached.

(vii) If possible, select the tubing–flow line combination whose OPR curves intersects with present and all corresponding future IPR curves and provides values for q that are consistent with the planned production strategy.

(viii) When more than one tubing–flow line combinations are possible, select the one that provides the best economics.

Determining Tubing Grade and Weight

Once the tubing size has been determined, the next step is to determine the grade and weight of the tubing. Similar to casing, tubing grade refers to the type of steel alloy and its minimum yield strength. Tubing are available in the same grades as casing (i.e., H-40, J-55, N-80, C-75, L-80, and P-105). Other high-strength grades that are resistance to sulfide stress cracking are also available.

To determine the tubing grade and weight, the maximum collapse and burst loads that act on the tubing are first calculated and multiplied by a safety factor. These values are then used to make an initial selection of tubing grade and weight that provides sufficient collapse and burst resistance. With the selected weight, the tension load is calculated and multiplied by a safety factor. This is then compared to the tensile strength of the selected tubing. The selected tubing is accepted if its tensile strength matches or exceeds the tensile load. Otherwise, another grade and/or weight are selected and the calculations are repeated until a final selection is made.

1.8.3 Completion and Workover Operations

Completion and workover operations are basically similar. However, they are given different associations based on when the operation is performed. Completion operations are any and all operations performed on the well to get if ready for production. Workover operations, however, refer to such operations that are performed to resolve specific problems that are found after production has started. Some of the important and common operations are briefly described in the following subsections.

Perforating Operation

For cased hole completions, perforations are made through the casing and cement and into the formation to establish communication between the formation and the wellbore. It is essential to have clean perforations with relatively large diameters and deep penetrations to achieve high well productivity. Further, perforating should be done only through the clean and productive zones within the formation, as determined from the formation evaluation logs. Therefore, extreme care is taken in locating the perforating gun at the right locations. The selection of the type of perforating gun, explosive charges, and completion fluid and the control of the pressure in the well at the time of perforating are very important elements in achieving effective and productive perforations.

Perforations are made by detonating specially shaped explosive charges. The shaped charge consists of a body called the case, a linear that is made of a powder alloy of lead, copper, and tungsten, the explosive material that is contained between the case and the liner, and a detonating cord. Figure 19 shows a schematic cross section of a shaped charge. Upon explosion, the case expands and ruptures and the liner collapses into a carrot-shaped jet consisting of lead, copper, and tungsten particles. The jet travels at very high velocity and impacts upon the casing with an extremely high pressure. The high energy of the jet causes the jet to penetrate through the casing, cement, and formation, thus creating the perforation.

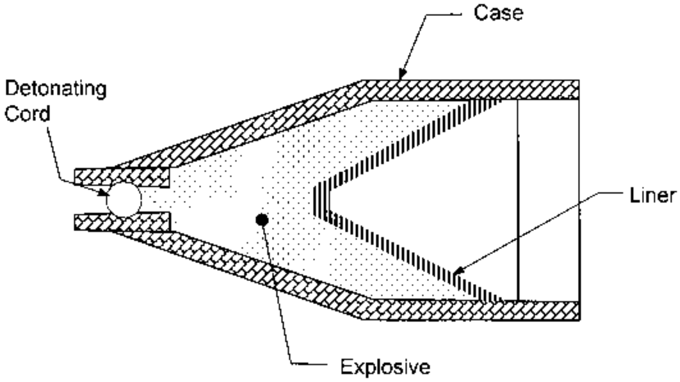
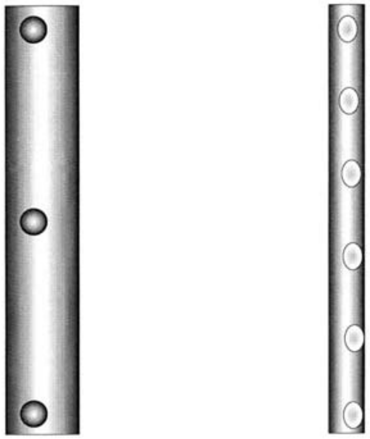
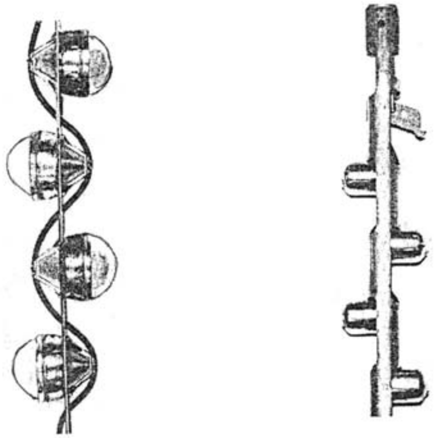


Figure 19 Shaped charge (cutaway).



Casing Gun

Through-Tubing Gun



Semi-Expendable
Wire Carrier & Strip

Fully Expendable

Figure 20 Types of perforating gun.

Perforating guns are classified as *hollow steel carrier*, *semi expendable*, and *fully expendable* guns (Fig. 20). Hollow steel carrier guns are made of steel cylinders that carry the explosive charges and needed accessories. They come in small diameters that can go through the production tubing, and in large diameters that can go only through the production casing. They are rigid, can withstand high pressure and temperature conditions, and have the advantage of containing all of the debris that is left after perforating. However, because of their rigidity, they would not go through sharp bends. For semiexpendable guns, the charges are fixed to a flexible metal strip. After detonation, most of the debris is expelled into the well, but the metal strip, any unfired charges, and the electronic detonator are retrieved. The flexibility of the guns allows movement through bends and highly deviated holes. Fully expendable guns have advantages similar to those of semiexpendable guns. However, because, after detonation, the gun breaks completely into pieces that are left in the hole, there is no way to know whether all charges have been fired.

To obtain clean and relatively undamaged perforations, the perforating operation should be conducted with the bottom-hole pressure much less than the reservoir pressure. Upon perforating, the differential pressure causes fluids to surge from the formation into the well; this effectively cleans the perforations from any debris. This method of perforating is known as *underbalance perforating*. To perforate underbalance, the production tubing must be installed in the well and the peforating mode is, therefore, identified as the *through-tubing perforating mode*.

In the early days, large through-casing hollow steel carrier guns were always needed to obtain large-diameter perforations with deep penetration. This required that the tubing be out of the well, and in order to have control over the well, perforating had to be conducted with the bottom-hole pressure being higher than the reservoir pressure by a safe margin. This mode of *perforating overbalance* did not offer an effective means for cleaning the perforations of the debris. Another mode of perforating which offers the opportunity to use large casing guns and perforating underbalance is known as *tubing-conveyed perforating*. Recent technological developments, however, have made it possible to obtain such desired perforations using small-size charges on a semiexpendable gun.

Well Stimulation Operations

Well stimulation operations could be classified into two main categories: *matrix acidizing* and *formation fracturing*. The objective of matrix acidizing is to remove near-wellbore formation damage that might have been caused by drilling or other workover operations in order to restore or enhance well productivity. Formation fracturing, however, is conducted on formations having very low permeability in order to increase well productivity.

For carbonate formation, matrix acidizing is achieved by injecting hydrochloric acid (HCl) into the formation at low pressure. As acid is injected, it preferentially flows through more permeable passages, reacting with and dissolving the carbonate. This creates irregular highly conductive channels, known as a *wormholes*, for easier flow of fluid from the formation to the well.

For sandstone matrix acidizing, a solution of hydrofluoric and hydrochloric acids (known as *mud acid*) is injected into the formation to dissolve clay and, to some extent, silica. This removes the damage to restore the near-wellbore permeability.

Fracturing is performed mostly on wells completed in very tight (very low permeability) formations. In carbonate formations, a fracturing fluid followed by HCl is injected at pressures exceeding the formation fracturing pressure. The fracturing fluid initiates a fracture and the acid reacts with the carbonate walls of the fracture, leaving the walls as rough surface. Therefore, when the pressure is reduced, the fracture will not close and will provide a very conductive passage for fluids to flow from the formation into the well. This operation is known as *acid fracturing*. Fracturing tight sandstone formations is known as *hydraulic fracturing*. As with carbonate fracturing, a fracturing fluid is injected at high pressure to initiate the fracture. The fracture is kept open by filling it with highly permeable, high-compressive-strength sand known as *proppant*.

Sand Control

When wells are completed in unconsolidated or weakly consolidated formations, sand is likely to be produced with the fluids. Sand production is a very serious problem. Produced sand erodes subsurface and surface equipment, necessitating very costly frequent replacement. Sand also settles in the bottom of the well and in surface processing facilities. This requires periodic shutdowns to clean the well and facilities.

Methods are available to control or prevent sand production. These are classified as *mechanical retention methods* and *chemical*, or *plastic*, *consolidation methods*. The simplest method of mechanical retention involves the installation of screens opposite the producing zone. Screens, however, are subject to erosion and corrosion and would need to be replaced frequently. Therefore, screens are used only as a temporary solution to sand production problems. The best and most commonly used method of mechanical sand control is known as *gravel packing*. In simple terms, the method involves the installation of a screen having a smaller diameter than the casing diameter opposite the producing zone. The annular space between the casing and screen is packed with specially sized gravel (sand). Formation sand will bridge against the gravel, which bridges against the screen. This prevents formation sand from flowing with the fluids into the well.

Chemical or plastic consolidation involves the injection of the polymer and catalyst solution into the formation to coat the sand grains around the wellbore with the polymer. The polymer is then cured to harden and bonds the sand grains together and thus consolidates the sand around the wellbore.

Each sand control method has its limitations and care should be taken to select the most suitable method for the situation in hand. All methods of sand control result in loss of productivity; this, however, is acceptable in comparison to the problems associated with sand production.

Remedial Cementing

Remedial cementing refers to any cementing operation performed on the well after placing it in production. These are basically repair jobs executed to resolve specific problems in the well. The most common applications of remedial cementing are the following:

1. *Control of GOR and WOR*. Squeezing cement into the perforations producing gas or water can stop excessive gas or water production with the oil and thus reduce an undesired high gas–oil ratio (GOR) or water–oil ratio (WOR).
2. *Channel Repair*. A channel is a void or crack that developed during the primary cementing of the production casing string. When the channel behind the casing communicates two or more zones, unwanted fluids (such as water) may flow through the channel into the formation and be produced with the oil. Alternatively, the oil may flow through the channel and be lost into another lower-pressure formation. A channel is repaired by creating a few perforations at the top and bottom of the channel and squeezing cement into the channel to seal if off.
3. *Recompletion*. When one well is used to produce from multiple-reservoir or multiple zones with one reservoir or zone produced at a time, the depleted zone must be plugged off before producing the next zone. Cement is squeezed into the old perforations to seal the depleted zone completely; then, the new zone is perforated.
4. *Casing Leak Repair*. Depending on the nature of the leak, cement may be used to seal parts of the casing, leaking unwanted fluids into the well or allowing produced oil to leak into other formations.
5. *Setting Liners*. Cementing liners to convert an open hole completion into a perforated liner completion.

6. *Plug and Abandonment*. For abandoning depleted wells, cement is used to squeeze and seal off all perforations and to set several cement plugs in the well.

1.8.4 Producing the Well

Producing the well means bringing the fluids that flowed from the formation into the borehole, from the bottom of the well to the surface. Fluids need to be brought to the surface at the desired rate and with sufficient pressure to flow through the surface-treating facilities. Some reservoirs possess such a high pressure that it can produce the desired rates at high bottom-hole pressures that could push the fluids to the surface at the desired wellhead pressure. This mode of production is known as *natural flow*. Reservoirs with high initial pressures and with strong pressure support (i.e., from bottom water) can be produced under natural flow for extended period of time.

When the reservoir pressure declines, pressure support may be provided by injecting water or gas into the reservoir to maintain natural flow at desired rates and surface pressures. In some cases, however, pressure support may not be sufficient to maintain desired natural flow. This usually occurs due to increased water production, which increases the hydrostatic head and friction losses in the tubing. Therefore, the fluids may reach the surface at lower than desired pressures, or may not even reach the surface. In such cases, external means of lifting the fluids to the surface will be needed. These means are known as the *artificial lift methods*. Production engineers are responsible for selecting, designing, installing, and operating artificial lifting facilities.

Artificial Lift Methods

The objective of artificial lift is to create low bottom-hole pressure to allow high rates of production from the formation into the bottom of the well and artificially lift the fluids to the surface with the desired surface pressure. The main methods of artificial lift are as follows:

1. *Sucker Rod Pumping (SRP):* This is the oldest and most common method of artificial lift used in the industry. The method is, however, limited to vertical and straight wells with relatively low productivity. The pumping facility consists mainly of surface equipment that provide reciprocating up-and-down movement to a rod string (the sucker rod) that is connected to

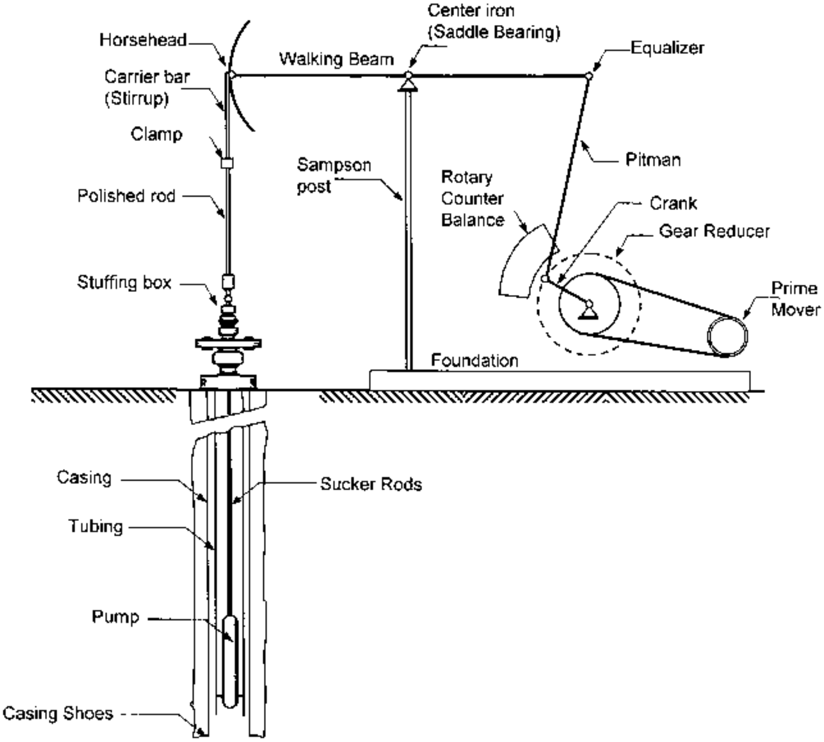


Figure 21 Schematic of sucker rod pumping.

the bottom-hole pump (Fig. 21). The pump may be thought of as a cylinder-plunger system with one-way valves. Produced fluids enter the cylinder above the plunger during the down stroke and are lifted to the surface during the up stroke.

2. *Hydraulic Pumping (HP):* In this method, a pump located at the bottom of the well is powered by high-pressure fluid. The function of the pump is to increase the produced fluid pressure to lift it to the surface with the desired wellhead pressure. Figure 22 shows a schematic of hydraulic pumping facility. The surface facility provides the high-pressure hydraulic fluid that powers the pump. Hydraulic fluid may flow through a closed circuit and does not come in contact with the produced fluids. Oil may also be used as the hydraulic fluid, allowing its mixing with the produced fluids. The major limitations of this method are the need for expensive centralized hydraulic power units and expensive clean hydraulic fluid.

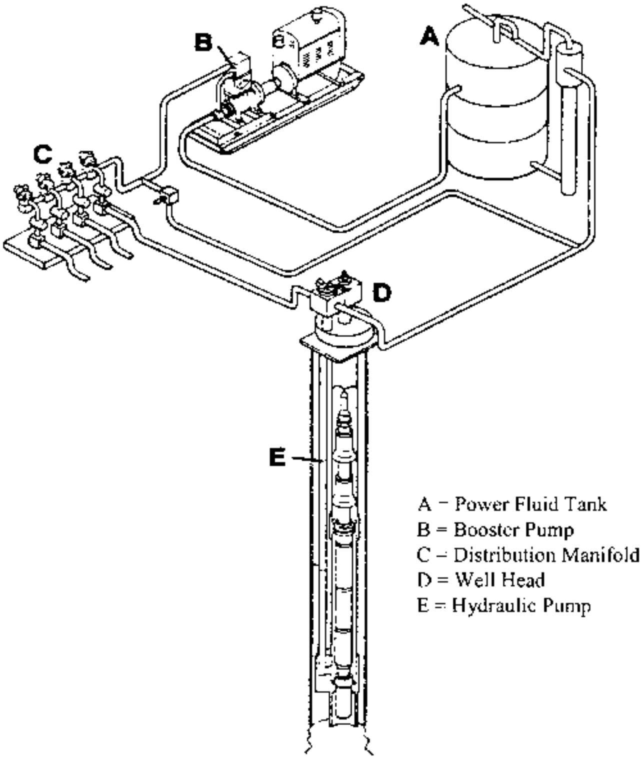


Figure 22 Schematic of hydraulic pumping installation.

3. *Electric Submersible Pumps (ESP):* This method uses centrifugal pumps powered by electric motors. The pump and motor are both located at the bottom of the well and electric power is provided to the motor from the surface through a special cable, (Fig. 23). As with the HP, the function of the ESP is to increase the pressure of the fluid so that it can move and reach the surface with high pressure. The ESP is capable of producing very high rates with a high surface pressure.
4. *Gas Lift (GL):* Although many different types of gas-lift installation exist, the lifting concept is the same. High-pressure gas is injected into the annular space between the casing and tubing and enters the tubing, through special gas-lift valves (Fig. 24). The gas mixes with fluids in the tubing, reducing its density and, consequently, reducing the hydrostatic head imposed by the fluid at the bottom of the well. The reduction

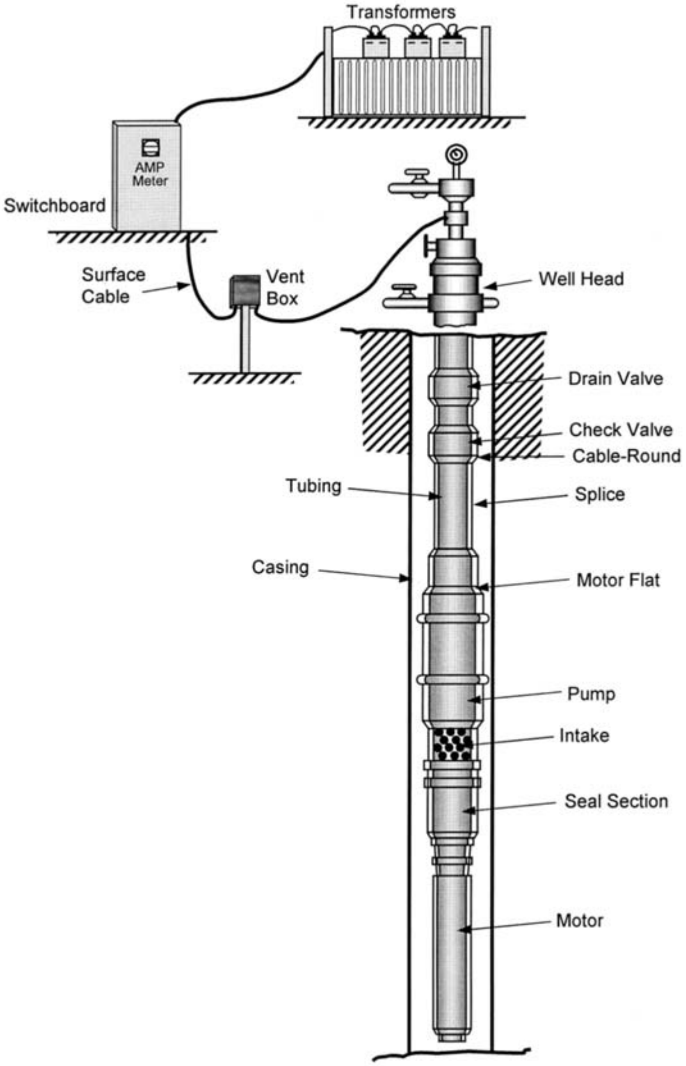


Figure 23 Typical standard complete pumping system. (Courtesy of Bryon Jackson, Centrilift.)

in bottom-hole pressure causes more production of fluids from the formation. The reduced hydrostatic head reduces the pressure losses in the tubing and thus enables the fluid to reach the surface with relatively high pressure.

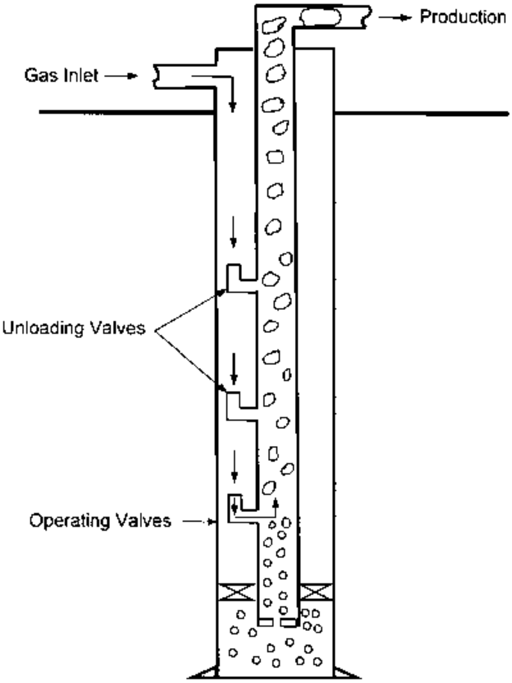


Figure 24 Schematic of a gas lift.

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REVIEW QUESTIONS

1. Describe briefly the organic theory of petroleum formation.
2. Define the petroleum reservoir.
3. What are the necessary conditions that lead to the accumulation of petroleum to form a petroleum reservoir?
4. What is meant by the "source rock"?
5. Name and briefly describe, with illustrations, the different geologic types of petroleum reservoir.
6. What is meant by "reservoir drive mechanism"?
7. Describe the different types of reservoir drive mechanism.
8. Name and briefly describe the techniques used for petroleum exploration.
9. What are the functions of drilling fluids?
10. What is the purpose of the well casting?
11. Describe briefly the role and functions of reservoir engineering in petroleum field development and operation.
12. Show, using illustrations and giving reasons, where would you locate and complete wells for an anticline gas-cap-drive reservoir.
13. Show, using illustrations and giving reasons, where would you locate and complete wells for an anticline water-drive reservoir.
14. What are the roles of reservoir simulation?
15. What is meant by "improved recovery"?
16. Describe the various methods of improved recovery.
17. Describe briefly the role and functions of production engineering in petroleum fields development and operation.
18. What are the main types of well completion?
19. Define the "productivity index" of the well.
20. What is the purpose of "perforating" a well?
21. Explain the function of each of the following "workover operations":

    (i) Stimulation
    (ii) Sand control
    (iii) Remedial cementing

22. Describe the difference between "artificial lift" and improved recovery.
23. What are the different types of artificial lift method?

2

Composition and Characteristics of Crude Petroleum

A Brief Review

2.1 GENERAL

Crude oils are complex mixtures of a vast number of hydrocarbon compounds. Properties of crude petroleum vary appreciably and depend mainly on the origin. In this chapter, the chemical composition of the crude oils is viewed, including the hydrocarbon series as well as the nonhydrocarbon compounds. Physical methods generally used for indentifying types of crude oils are described next. Characterization and classification of crude oils based on correlation indexes and crude assays are presented, followed by a comparison between some of the well-known types of oil.

2.2 CRUDE OIL COMPOSITION

The raw material that we deal with—referred to as crude petroleum—is, by definition, the naturally occurring rock oil produced as was explained in Chapter 1. In general, composition of crude oil may be studied by two methods:

1. Chemical approach
2. Physical methods

Chemical composition describes and identifies the individual chemical compounds isolated from crude oils over the years. Physical representation, on the other hand, involves considering the crude oil and its products as mixtures of hydrocarbons and describing physical laboratory tests or methods for characterizing their quality.

2.2.1 Chemical Approach

Nearly all petroleum deposits are made up of a mixture of chemical compounds that consist of hydrogen and carbon, known as hydrocarbons, with varying amounts of nonhydrocarbons containing S, N_2, O_2, and other some metals. The composition of crude oil by elements is approximated as shown in Table 1 [1]. It could be further stated that these hydrocarbon compounds making up oils are grouped chemically into different series of compounds described by the following characteristics:

- Each series consists of compounds similar in their molecular structure and properties (e.g., the alkanes or parafin series).
- Within a given series, there exists a wide spectrum of compounds that range from extremely light or simple hydrocarbon to a heavy or complex one. An example, CH_4 for the former and $C_{40}H_{82}$ for the latter in the paraffinic series.

Hydrocarbon Series

The major constituents of most crude oils and its products are hydrocarbon compounds, which are made up of hydrogen and carbon only. These compounds belong to one of the following subclasses:

1. *Alkanes or Paraffins:* Alkanes are saturated compounds having the general formula C_nH_{2n+2}. Alkanes are relatively nonreactive compounds in comparison to other series. They may either be straight-chain or branched compounds, the latter are more valuable than the former, because they are useful for the production of high-octane gasoline.
2. *Cycloalkanes or Cycloparaffins (Naphtenes):* Cycloalkanes and bicycloalkanes are normally present in crude oils and its fractions in variable proportions. The presence of large amounts

Table 1 Composition of Petroleum Crude

Element	Percent by weight
Carbon	83–87
Hydrogen	11–14
Sulfur	0.05–2.5
Nitrogen	0.1–2
Oxygen	0–2

Note: Sulfur, nitrogen, and oxygen are regarded as impurities.

of these cyclic compounds in the naphtha range has its significance in the production of aromatic compounds. Naphtha cuts with a high percentage of naphthenes would make an excellent feedstock for aromatization.

3. *Alkenes or Olefins:* Alkenes are unsaturated hydrocarbon compounds having the general formula C_nH_n. They are practically not present in crude oils, but they are produced during processing of crude oils at high temperatures.

   Alkenes are very reactive compounds. Light olefinic hydrocarbons are considered the base stock for many petrochemicals. Ethylene, the simplest alkene, is an important monomer in this regard. For example, polyethylene is a well-known thermoplastic polymer and polybutadiene is the most widely used synthetic rubber.
4. *Aromatics:* Aromatic compounds are normally present in crude oils. Only monomolecular compounds in the range of C6–C8 (known as B-T-X) have gained commercial importance. Aromatics in this range are not only important petrochemical feedstocks but are also valuable for motor fuels.

   Dinuclear and polynuclear aromatic compounds are present in heavier petroleum fractions and residues. *Asphaltenes*, which are concentrated in heavy residues and in some asphaltic crude oils, are, in fact, *polynuclear aromatics* of complex structures. It has been confirmed by mass spectroscopic techniques that condensed-ring aromatic hydrocarbons and heterocyclic compounds are the major compounds of asphaltenes.

Nonhydrocarbon Compounds

So far, a brief review of the major classes of the hydrocarbon compounds that exist in crude oils and their products was presented. For completeness, we should mention that other types of nonhydrocarbon compound occur in crude oils and refinery streams. Most important are the following:

- Sulfur compounds
- Nitrogen compounds
- Oxygen compounds
- Metallic compounds

Sulfur Compounds. In addition to the gaseous sulfur compounds in crude oil, many sulfur compounds have been found in the liquid phase in the form of organosulfur. These compounds are generally not acidic.

Sour crude oils are those containing a high percentage of hydrogen sulfide. However, many of the organic sulfur compounds are not thermally stable, thus producing hydrogen sulfide during crude processing.

High-sulfur crude oils are in less demand by refineries because of the extra cost incurred for treating refinery products. Naphta feed to catalytic reformers is hydrotreated to reduce sulfur compounds to very low levels (1 ppm) to avoid catalyst poisoning.

The following sulfur compounds are typical:

1. *Mercaptans (H–S–R):* Hydrogen sulfide, H–S–H, may be considered as the simple form of mercaptan; however, the higher forms of the series are even more objectionable in smell. For example, butyl mercaptan ($H{-}S{-}C_4H_9$) is responsible for the unusual odor of the shank.
2. *Sulfides (R–S–R):* When an alkyl group replaces the hydrogen in the sulfur-containing molecule, the odor is generally less obnoxious. Sulfides could be removed by the hydrotreating technique, which involves the hydrogenation of the petroleum streams as follows:

   $$R{-}S{-}R + 2H{-}H \rightarrow 2R{-}H + H{-}S{-}H$$
   $$R{-}S{-}R + H{-}H \rightarrow R{-}R + H{-}S{-}H$$

   The hydrogen sulfide may be removed by heating and may be separated by using amine solutions.
3. *Polysulfides (R–S–S–R):* These are more complicated sulfur compounds and they may decompose, in some cases depositing elemental sulfur. They may be removed from petroleum fractions, similar to the sulfides, by hydrotreating.

Nitrogen Compounds. Nitrogen compounds in crude oils are usually low in content (about 0.1–0.9%) and are usually more stable than sulfur compounds. Nitrogen in petroleum is in the form of heterocyclic compounds and may be classified as basic and nonbasic. Basic nitrogen compounds are mainly composed of pyridine homologs and have the tendency to exist in the high-boiling fractions and residues. The nonbasic nitrogen compounds, which are usually of the pyrrole and indole, also occur in high-boiling fractions and residues. Only a trace amount of nitrogen is found in light streams.

During hydrotreatment (hydrodesulfurization) of petroleum streams, hydrodenitrogeneation takes place as well, removing nitrogen as ammonia gas, thus reducing the nitrogen content to the acceptable limits for feedstocks to catalytic processes.

It has to be stated that the presence of nitrogen in petroleum is of much greater significance in refinery operations than might be expected from the very small amounts present. It is established that nitrogen compounds are responsible for the following:

1. Catalyst poisoning in catalytic processes
2. Gum formation in some products such as domestic fuel oils

Oxygen Compounds. Oxygen compounds in crude oils are more complex than sulfur compounds. However, oxygen compounds are not poisonous to processing catalysts. Most oxygen compounds are weakly acidic, such as phenol, cresylic acid and naphthenic acids. The oxygen content of petroleum is usually less than 2%, although larger amounts have been reported.

Metallic Compounds. Many metals are found in crude oils; some of the more abundant are sodium, calcium, magnesium, iron, copper, vanadium, and nickel. These normally occur in the form of inorganic salts soluble in water—as in the case of sodium chloride—or in the form of organometallic compounds—as in the case of iron, vanadium, and nickel.

The occurrence of metallic constituents in crude oils is of considerably greater interest to the petroleum industry than might be expected from the very small amounts present. The organometallic compounds are usually concentrated in the heavier fractions and in crude oil residues. The presence of high concentration of vanadium compounds in naphtha streams for catalytic reforming feeds will cause permanent poisons. These feeds should be hydrotreated not only to reduce the metallic poisons but also to desulfurize and denitrogenate the sulfur and nitrogen compounds.

Hydrotreatment may also be used to reduce the metal content in heavy feeds to catalytic cracking.

2.2.2 Physical Methods

Having discussed the various chemicals found in crude oils and realizing not only the complexity of the mixture but the difficulty of specifying a crude oil as a particular mixture of chemicals, we can understand why the early petroleum producers adopted the physical methods generally used for classification.

As may be seen, crude oils from different locations may vary in appearance and viscosity and also vary in their usefulness as producers for final products. It is possible by the use of certain basic tests to identify the quality of crude oil stocks. The tests included in the following list are primarily physical (except sulfur determination):

1. Distillation
2. Density, specific gravity, and API (American Petroleum Industry) gravity
3. Viscosity
4. Vapor pressure
5. Flash and fire points
6. Cloud and pour points
7. Color
8. Sulfur content
9. Basic sediments and water (B.S.&W.)
10. Aniline point
11. Carbon residue

The details of some of these tests are described next.

API Gravity

Earlier, density was the principal specification for petroleum products. However, the derived relationships between the density and its fractional composition were only valid if they were applied to a certain type of petroleum. Density is defined as the mass of a unit volume of material at a specified temperature. It has the dimensions of grams per cubic centimeter.

Another general property, which is more widely, is the specific gravity. It is the ratio of the density of oil to the density of water and is dependent on two temperatures, those at which the densities of the oil sample and the water are measured. When the water temperature is 4°C (39°F), the specific gravity is equal to the density in the cgs system, because the volume of 1 g of water at that temperature is, by definition, 1 mL. Thus, the density of water, for example, varies with temperature, whereas its specific gravity is always unity at equal temperatures. The standard temperatures for specific gravity in the petroleum industry in North America are 60/60°F and 15.6/15.6°C.

Although density and specific gravity are used extensively in the oil industry, the API gravity is considered the preferred property. It is expressed by the following relationship:

$$^\circ\mathrm{API} = \frac{141.5}{\gamma} - 131.5$$

where γ is the oil specific gravity at 60°F. Thus, in this system, a liquid with a specific gravity of 1.00 will have an API of 10 deg. A higher API gravity indicates a lighter crude or oil product, whereas a low API gravity implies a heavy crude or product.

Carbon Residue

Carbon residue is the percentage of carbon by weight for coke, asphalt, and heavy fuels found by evaporating oil to dryness under standard laboratory conditions. Carbon residue is generally referred to as CCR (Conradson carbon residue). It is a rough indication of the asphaltic compounds and the materials that do not evaporate under conditions of the test, such as metals and silicon oxides.

Viscosity

The viscosity is the measure of the resistance of a liquid to flow, hence indicating the ‘‘pumpability’’ of oil.

Pour Point

This is defined as the lowest temperature (5°F) at which the oil will flow. The lower the pour point, the lower the paraffin content of the oil.

Ash Content

This is an indication of the contents of metal and salts present in a sample. The ash is usually in the form of metal oxides, stable salts, and silicon oxides. The crude sample is usually burned in an atmosphere of air and the ash is the material left unburned.

Reid Vapor Pressure

The Reid Vapor Pressure (RVP) is a measure of the vapor pressure exerted by oil or by light products at 100°F.

Metals

In particular, arsenic, nickel, lead, and vanadium are potential poisons for process catalysts. Metal contents are reported in parts per million (ppm).

Nitrogen

It is the weight of total nitrogen determined in a liquid hydrocarbon sample (in ppm). Nitrogen compounds contribute negatively to process catalysts.

Salt Content

Salt content is typically expressed as pounds of salt (sodium chloride, NaCl) per 1000 barrels of oil (PTB). Salts in crude oil and in heavier products may create serious corrosion problems, especially in the top-tower zone and the overhead condensers in distillation columns.

Sulfur

This is the percentage by weight (or ppm) of total sulfur content determined experimentally in a sample of oil or its product. The sulfur content of crude oils is taken into consideration in addition to the API gravity in determining their commercial values. It has been reported that heavier crude oils may have a high sulfur content [2].

Hydrogen Sulfide

Hydrogen sulfide dissolved in a crude oil or its products is determined and measured in parts per million. It is a toxic gas that can evolve during storage or in the processing of hydrocarbons.

The above tests represent many properties for the crude oils that are routinely measured because they affect the transportation and storage facilities. In addition, these properties define what products can be obtained from a crude oil and contribute effectively to safety and environmental aspects. The price of a crude oil is influenced by most of these properties.

To conclude, it can be stated that light and low-sulfur crude oils are worth more than heavy and high-sulfur ones. One can summarize the two approaches of examining crude oils as follows:

1. Chemical composition
2. Physical properties:
    - (a) API, S, salt, metals, nitrogen and so forth
    - (b) Distillation: ASTM, TBP, EFV
    - (c) Correlations: Kw, Ind

where TBP is true boiling point, EFV is equilibrium flash vaporization, Kw is Watson characterization factor, IND is U.S. Bureau of Mines correlation index.

2.3 CRUDE OIL COMPARISONS AND CRUDE OIL ASSAY

In order to establish a basis for the comparison between different types of crude oil, it is necessary to produce experimental data in the form of what is known as an “assay.” Crude assays are the systematic compilation of data for the physical properties of the crude and its fractions, as well as the yield. In other words, a crude assay involves the determination of the following:

1. The properties of crude oil
2. The fractions obtained: (a) their percentage yield and (b) properties

Analytical testing only without carrying out distillation may be considered an assay. However, the most common assay is a comprehensive one that involves all of the above-stated parameters.

The basis of the assay is the distillation of a crude oil under specified conditions in a batch laboratory distillation column, operated at high efficiency [column with 14 plates and reflux ratio (RR)]. Pressure in column is reduced in stages to avoid thermal degradation of high boiling components.

A comparison of the characteristics of different types of crude oil over the distillation range could be made via a graph that relates the following:

- The density of distillate fractions
- Their mid-boiling points

Such a comparison is illustrated in Figure 1 [3]. The density level of a crude at given boiling point on the curve is a function of the relative proportions of the main three hydrocarbon series: aromatics, cycloparafins, paraffins; their densities decrease in that order.

In order to show how the properties of crude oils affect strongly processing requirements, product expectations, storage and transportation, and others, a comparison is presented as given in Table 2.

Table 3, on the other hand, gives the percent yield and other characteristics of the fractions obtained by the distillation of a typical Arabian crude oil having an API gravity of about 34–37.

2.4 CRUDE OIL CLASSIFICATIONS AND CHARACTERIZATION

2.4.1 Classifications

Although there is no specific method for classifying crude oils, it would be useful to establish simple criteria to quantify the quality of a crude. Numerous attempts have been made to devise a system to classify crude oils into types based on the predominant hydrocarbon series present in the

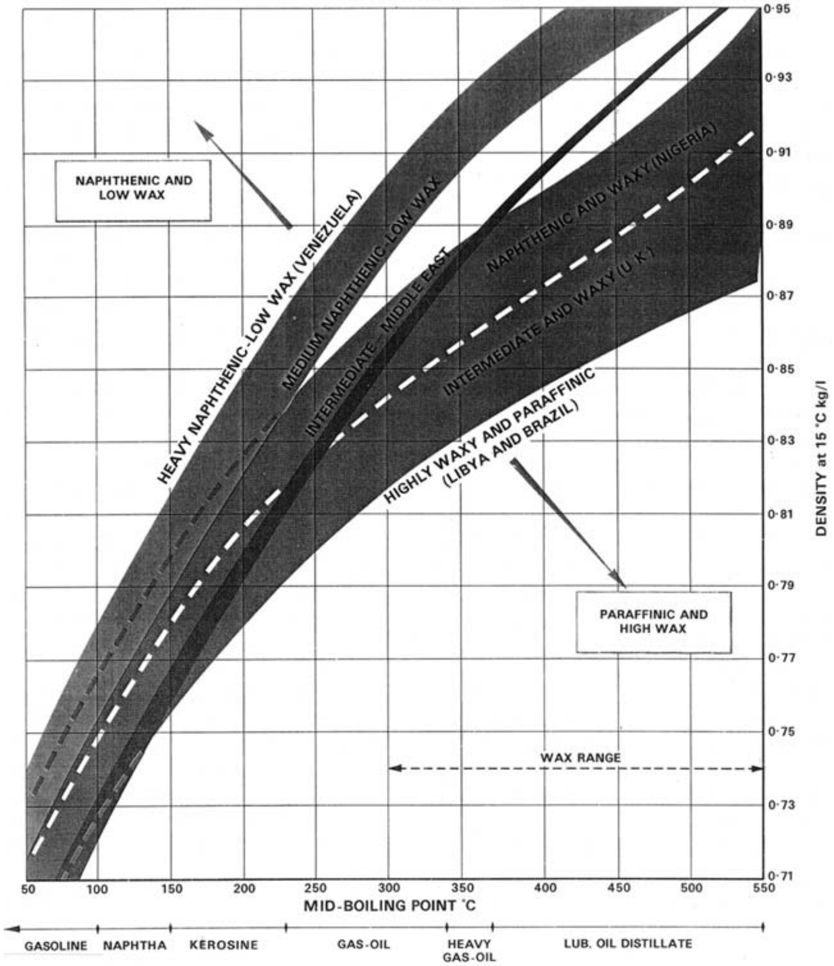

Figure 1 Comparison of crude oils density/mid-boiling point basis. (From Ref. 4.)

crude. Such attempts have only partially succeeded. In the United States, crude oils are classified into three types:

1. Paraffinic: paraffinic hydrocarbons with a relatively lower percentage of aromatics and naphthenes
2. Naphthenic: cycloparaffins in a higher ratio and a higher amount of asphalt than in paraffinic crudes
3. Asphaltic: fused aromatic compounds and asphalt in higher amounts

Table 2 Properties of Some Reference Crude Oils

Property	Arabian light	Arun indonesia	Beryl N.S	Nigerian light	SJV Calif.
API (gravity)	33.9	54.1	36.5	37.6	15.2
Pour Point (°F)	− 45	− 55	20	5	− 5
CCR (wt%)	3.6	0.01	1.3	1.1	7.0
Sulfur (wt%)	1.8	> 0.1	0.42	0.13	1.05
Nitrogen (ppm)	60	50	880	0.06	6200
Nickel (ppm)	3	0.65	0.8	3.6	63
Vanadium (ppm)	19	0.15	3.7	0.3	60
Salt Content (PTB)	10	3	7.4	5	14

Table 3 Fractions Obtained from Arabian Crude

Fractions	Percent yield	No. of carbon atoms in molecule	Boiling range (°F)
Gases (dry/wet)	2	1–2	−260 to −130 dry
		3–4	−55 to 30 wet
Naphtha	20–26	5–12	90–360
Kerosene	7–12	10–15	320–460
Diesel oil	10–14	12–20	400–600
Wax distillate	15–20	17–22	500–700
Residuum	35–40	20–90	600+

Another method of classification is the following:

1. Parafinic base
2. Mixed base
3. Naphthenic base

Based on this classification, a rating for the processing of crude oils is envisaged as follows for the production of certain products and their treatment:

Products:	Lub. oil	Asphalt	Gasoline	Treatment of products
Type of oil:				
Parafinic	1	3	3	1
Mixed	2	2	2	2
Naphthenic	3	1	1	3

Rating: 1. Excellent; 2. good; 3. poor.

2.4.2 Characterization Factors

Correlation indexes or characterization factors are used in the petroleum industry to indicate the crude type or class. There are several correlations between yield and type of crude in terms of aromaticity and parafinicity. The two most widely accepted relationships are the following:

1. Watson characterization factor: 0.333

   $$K_w = G\mathrm{Tb}$$

2. U.S. Bureau of Mines Correlation Index:

   $$\mathrm{Ind} = \mathrm{Tb}87.552 + 473.7G - 456.8$$

where G is the specific gravity at 60°F and Tb is the mean average boiling point (°R).
The Watson factor ranges from 10.5, for highly naphthenic crude oils, to 12.9, for the paraffinic type [4].

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3

Two-Phase Gas–Oil Separation

3.1 INTRODUCTION

At the high pressure existing at the bottom of the producing well, crude oil contains great quantities of dissolved gases. When crude oil is brought to the surface, it is at a much lower pressure. Consequently, the gases that were dissolved in it at the higher pressure tend to come out from the liquid. Some means must be provided to separate the gas from oil without losing too much oil.

In general, well effluents flowing from producing wells come out in two phases: vapor and liquid under a relatively high pressure. The fluid emerges as a mixture of crude oil and gas that is partly free and partly in solution. Fluid pressure should be lowered and its velocity should be reduced in order to separate the oil and obtain it in a stable form. This is usually done by admitting the well fluid into a gas–oil separator plant (GOSP) through which the pressure of the gas–oil mixture is successively reduced to atmospheric pressure in a few stages.

Upon decreasing the pressure in the GOSP, some of the lighter and more valuable hydrocarbon components that belong to oil will be unavoidably lost along with the gas into the vapor phase. This puts the gas–oil separation step as the initial one in the series of field treatment operations of crude oil. Here, the primary objective is to allow most of the gas to free itself from these valuable hydrocarbons, hence increasing the recovery of crude oil.

Crude oil as produced at the wellhead varies considerably from field to field due not only to its physical characteristics (as explained in Chapter 2) but also to the amount of gas and salt water it contains. In some fields, no salt water will flow into the well from the reservoir along with the produced oil. This is the case we are considering in this chapter, where it is only necessary to separate the gas from the oil; (i.e., two-phase separation).

When, on the other hand, salt water is produced with the oil, it is then essential to use three-phase separators—a case studied in Chapter 4.

Accordingly, oil-field separators can be classified into two types based on the number of phases to separate:

1. Two-phase separators, which are used to separate gas from oil in oil fields, or gas from water for gas fields.
2. Three-phase separators, which are used to separate the gas from the liquid phase, and water from oil.

Oil from each producing well is conveyed from the wellhead to a gathering center through a flow line. The gathering center, usually located in some central location within the field, will handle the production from several wells in order to process the produced oil–gas mixture.

Separation of the oil phase and the gas phase enables the handling, metering, and processing of each phase independently, hence producing marketable products.

3.2 THE SEPARATION PROBLEM

High-pressure crude oils containing large amount of free and dissolved gas flow from the wellhead into the flow line, which routes the mixture to the GOSP. In the separator, crude oil separates out, settles, and collects in the lower part of the vessel. The gas, lighter than oil, fills the upper part of the vessel. Crude oils with a high gas–oil ratio (GOR) must go through two or more stages of separation.

Gas goes out the top of the separators to a gas collection system, a vapor recovery unit (VRU), or a gas flow line. Crude oil, on the other hand, goes out the bottom and is routed to other stages of separation, if necessary, and then to the stock tank (Fig. 1).

Movement of the crude oil within the GOSP takes place under the influence of its own pressure. Pumps, however, are used to transfer the oil in its final trip to the tank farm, or pipeline (Fig. 2). Pressure reduction in moving the oil from stage to stage is illustrated in Fig. 3.

In order to visualize the changes that occur during the gas–oil separation process, two illustrations are presented.

First, consider an experiment where a sample of oil–gas mixture is placed in a cylinder with movable piston to simulate oil reservoir conditions. The weight exerted on the piston will be proportional to the pressure found in a reservoir. Initial conditions in the reservoir are as presented by case 1 in Figure 4, where 1.620 bbl of reservoir liquid with 1200 standard cubic feet (SCF) of dissolved gases are placed in the

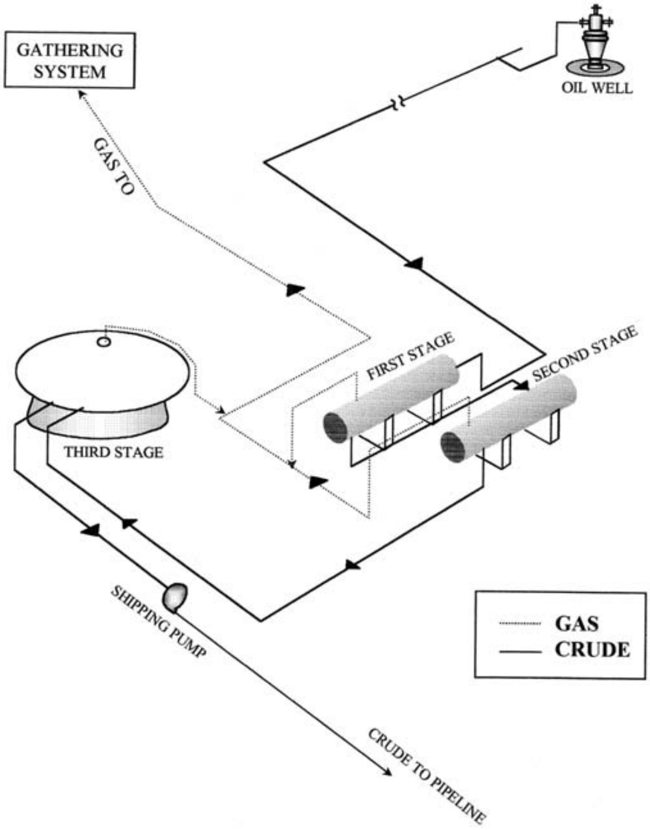


Figure 1 Flow of crude oil from oil well through GOSP.

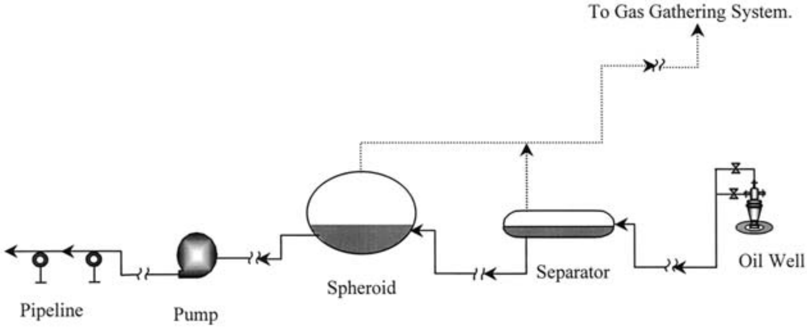


Figure 2 Separation of gas from oil.

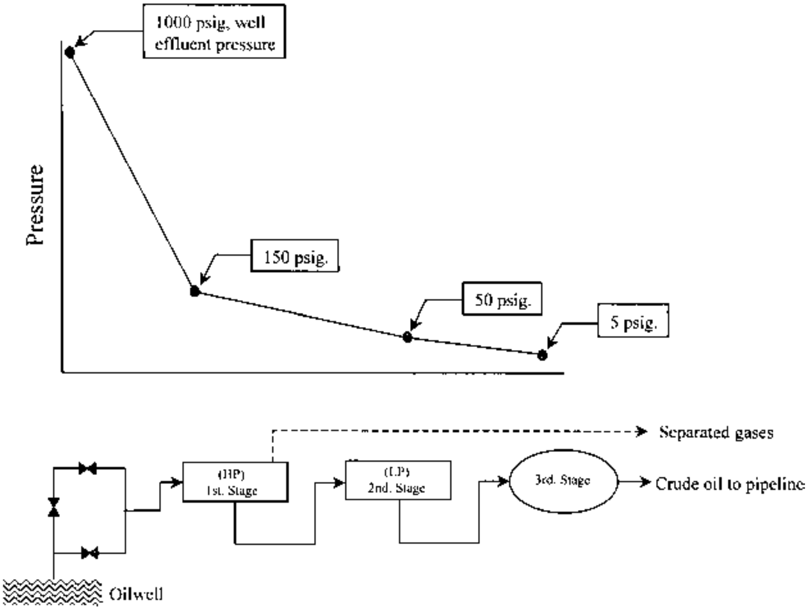
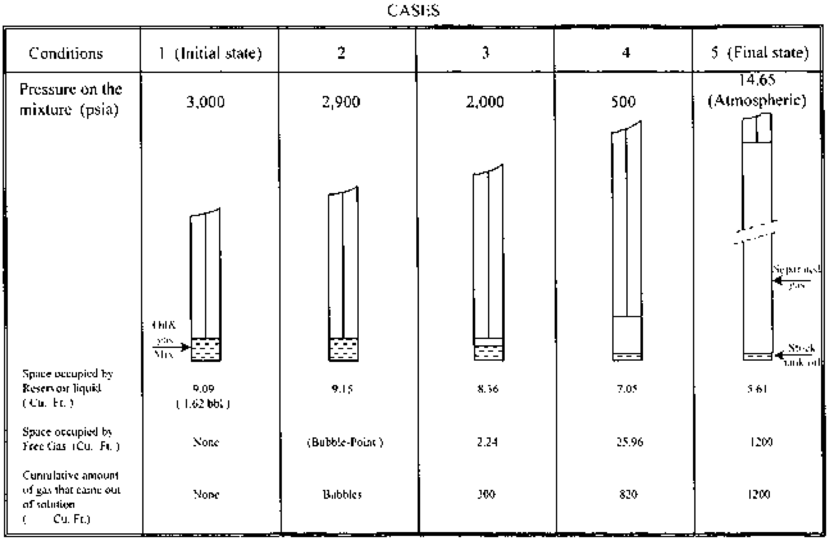


Figure 3 Pressure-drop profile for a typical GOSP in the middle east.



CASES

Conditions	1 (Initial state)	2	3	4	5 (Final state)
Pressure on the mixture (psia)	3,000	2,900	2,000	500	14.65 (Atmospheric)
Space occupied by Reservoir liquid (Cu. Ft.)	9.09 (1.62 bbl.)	9.15	8.36	7.05	5.61
Space occupied by Free Gas (Cu. Ft.)	None	(Bubble-Point)	2.24	25.96	1200
Cumulative amount of gas that came out of solution (Cu. Ft.)	None	Bubbles	300	820	1200

Figure 4 Demonstration of gas–oil separation process.

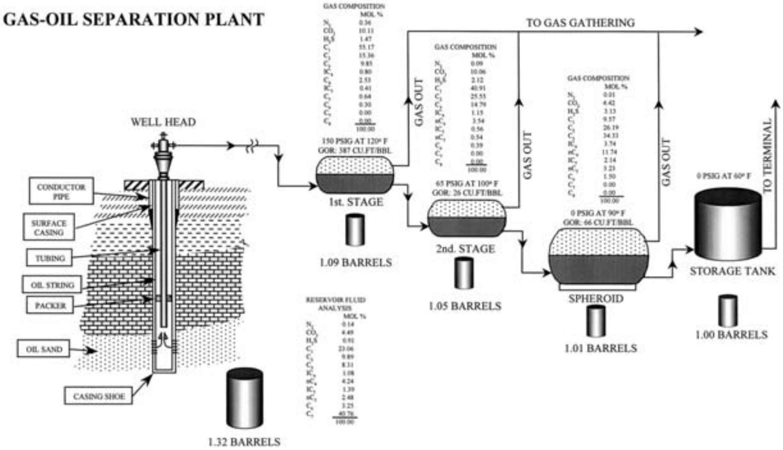


Figure 5 Gas–oil separation plant.

cylinder under 3000 psia pressure and the temperature is 60°F. The space occupied by the mixture under these conditions is 0.09 ft^3.

Now, in a series of repeated runs, the piston is raised to allow for a decrease in pressure, keeping the temperature constant. The volume of the gases separated in each run is measured and reported until reaching atmospheric conditions (14.5 psia and 60°F) in the final stage, corresponding to the surface conditions attained in the final stage in a GOSP.

In this demonstration, it is found that the original volume of oil of 1.620 bbl shrank to 1.0 bbl during its production to marketable products. The loss in volume of 0.620 bbl represents the space in the reservoir mixture that was occupied by 1200 ft^3 of the dissolved gas.

Second, consider Figure 5, which summarizes the results of a three-stage gas–oil separation pilot plant. The results are reported for an initial reservoir oil mixture of 1.32 bbl in place. Data reported for each stage show the flow rates as well as gas composition. One barrel of oil is collected in the storage tank, and the loss of 0.32 bbl represents the amount of gas separated.

3.3 THEORY OF GAS–OIL SEPARATION

In order to understand the theory underlying the separation of well-effluent hydrocarbon mixtures into a gas stream and oil product, it is

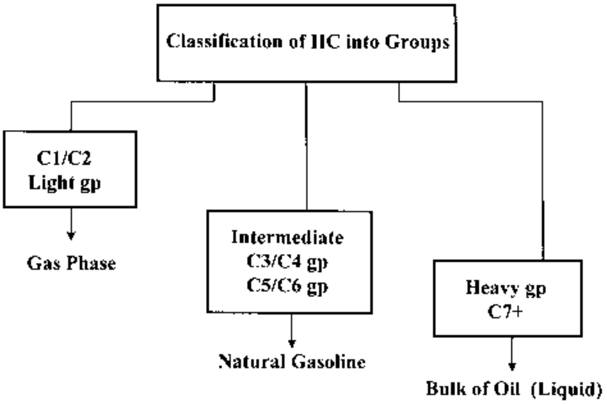


Figure 6 Classification of hydrocarbons found in wellhead fluids.

assumed that such mixtures contain essentially three main groups of hydrocarbon, as illustrated in Figure 6:

1. Light group, which consists of CH_4 (methane) and C_2H_6 (ethane)
2. Intermediate group, which consists of two subgroups: the propane/butane (C_3H_8/C_4H_{10}) group and the pentane/hexane (C_5H_{12}/C_6H_{14}) group.
3. Heavy group, which is the bulk of crude oil and is identified as C_7H_{16}.

In carrying out the gas–oil separation process, the main target is to try to achieve the following objectives:

1. Separate the C_1 and C_2 light gases from oil
2. Maximize the recovery of heavy components of the intermediate group in crude oil
3. Save the heavy group components in liquid product

To accomplish these objectives, some hydrocarbons of the intermediate group are unavoidably lost in the gas stream. In order to minimize this loss and maximize liquid recovery, two methods for the mechanics of separation are compared:

1. Differential or enhanced separation
2. Flash or equilibrium separation

In differential separation, light gases (light group) are gradually and almost completely separated from oil in a series of stages, as the total pressure on the well-effluent mixture is reduced. Differential separation is characterized by the fact that light gases are separated as soon as they are liberated (due to reduction in pressure). In other words, light components do not come into contact with heavier hydrocarbons; instead, they find their way out.

For flash separation, on the other hand, gases liberated from the oil are kept in intimate contact with the liquid phase. As a result, thermodynamic equilibrium is established between the two phases and separation takes place at the required pressure.

Comparing the two methods, one finds that in differential separation, the yield of heavy hydrocarbons (intermediate and heavy groups) is maximized and oil-volume shrinkage experienced by crude oil in the storage tank is minimized. This could be explained by the fact that separation of most of the light gases take place at the earlier high-pressure stages; hence, the opportunity of loosing heavy components with the light gases in low-pressure stages is greatly minimized. As a result, it may be concluded that flash separation is inferior to differential separation because the former experiences greater losses of heavy hydrocarbons that are carried away with the light gases due to equilibrium conditions. Nevertheless, commercial separation based on the differential concept is very costly and is not a practical approach because of the many stages required. This would rule out differential separation, leaving the flash process as the only viable scheme to affect gas–oil separation using a small number of stages. As illustrated in Table 1 [1], a close approach to differential separation is reached by using four to five flash separation stages. A comparison between the mechanisms of separation by the two methods is schematically simplified and presented in Table 2.

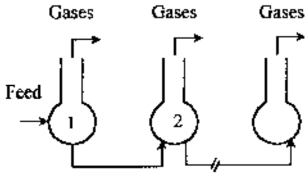
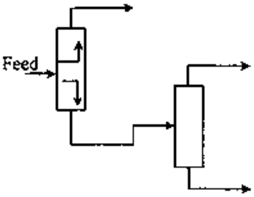
3.4 METHODS OF SEPARATION

The traditional process for recovering crude oil from high-pressure well streams is based on the flash separation concept explained in Section 3.3,

Table 1 Flash and Differential Separation

No. of flash stages	Percent approach to differential
2	0
3	75
4	90
5	96
6	98.5

Table 2 Comparison Between Mechanisms of Separation

Parameter	Differential separation	Flash separation
Process arrangement		
Losses of hydrocarbon	Low	High
No. of stages	Too many (can reach 100 stages)	Few (2–4 stages)
Commercial application	Not applied	Applied

which consists of a series of flash separators operating over a pressure range from wellhead pressure to atmospheric pressure. However, with the increased desirability of recovering natural gas and natural gas liquids (NGLs), other methods have been proposed as modification to the basic flash separation technique [2].

Separation methods could be broadly classified as follows:

1. Conventional methods
2. Modified methods
    - (a) Adding vapor recompression unit to the conventional methods
    - (b) Replacing the conventional methods by a stabilizer and a recompression unit

The conventional method is a multistage flash separation system and is recommended for comparatively high-pressure fluids. Several stages operated at successively lower pressures affect the separation of oil from gas, thus increasing the oil recovery. In general, the number of stages in a multistage conventional separation process is a function of the following:

1. The API gravity of the oil
2. The gas–oil ratio (GOR)
3. The flowing pressure

Consequently, high-API-gravity oils with high GOR flowing under high pressure would require the greatest number of stages (from three to four), as depicted in Figure 7.

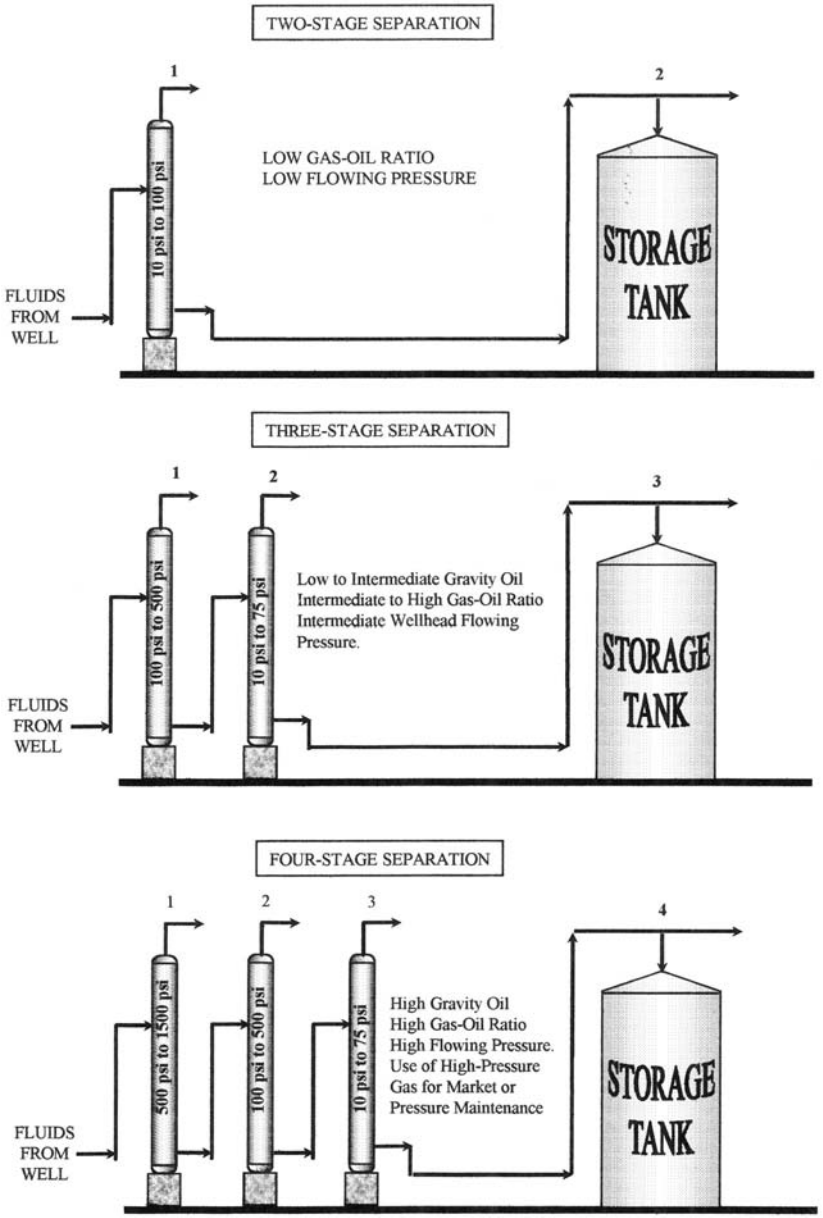


Figure 7 Stage separation flow diagrams.

The first modified method of separation implies adding several stages of gas compression to recompress the separated gas from each flash stage. Liquids from interstage vessels between the compressors can be collected and processed as liquid natural gas (LNG) stock. Natural gas will be delivered at the desired pressure depending on its usage. The second modified method of separation is different in concept from the conventional (flash) separation. It makes use of crude stabilizer columns. Normally, these columns have top feed trays with no rectifying section and no condenser, but are provided with interstage reboilers and feed preheaters. Crude stabilization systems are advantageous as GOSPs if space is critical, as may be encountered on an offshore platform, because they occupy less space than conventional GOSPs.

3.5 GAS–OIL SEPARATION EQUIPMENT

The conventional separator is the very first vessel through which the well-effluent mixture flows. In some special cases, other equipment (heaters, water knockout drums) may be installed upstream of the separator.

The essential characteristics of the conventional separator are the following:

1. It causes a decrease in the flow velocity, permitting separation of gas and liquid by gravity.
2. It always operates at a temperature above the hydrate point of the flowing gas.

The choice of a separator for the processing of gas–oil mixtures containing water or without water under a given operating conditions and for a specific application normally takes place guided by the general classification illustrated in Figure 8.

3.5.1 Functional Components of a Gas–Oil Separator

Regardless of their configuration, gas–oil separators usually consist of four functional sections, as shown in Figure 9:

1. *Section A:* Initial bulk separation of oil and gas takes place in this section. The entering fluid mixture hits the inlet diverter. This causes a sudden change in momentum and, due to the gravity difference, results in bulk separation of the gas from the oil. The gas then flows through the top part of the separator and the oil through the lower part.

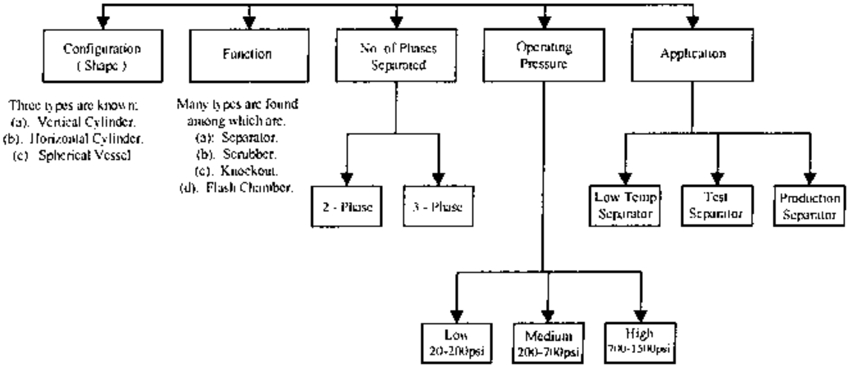


Figure 8 Classification of separators.

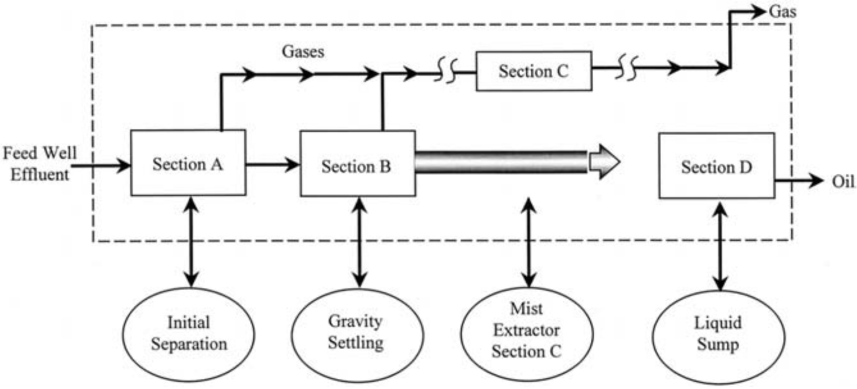


Figure 9 Schematic outline of the main components in a gas–oil separator

2. *Section B:* Gravity settling and separation is accomplished in this section of the separator. Because of the substantial reduction in gas velocity and the density difference, oil droplets settle and separate from the gas.
3. *Section C:* Known as the mist extraction section, it is capable of removing the very fine oil droplets which did not settle in the gravity settling section from the gas stream.
4. *Section D:* This is known as the liquid sump or liquid collection section. Its main function is collecting the oil and retaining it for a sufficient time to reach equilibrium with the gas before it is discharged from the separator.

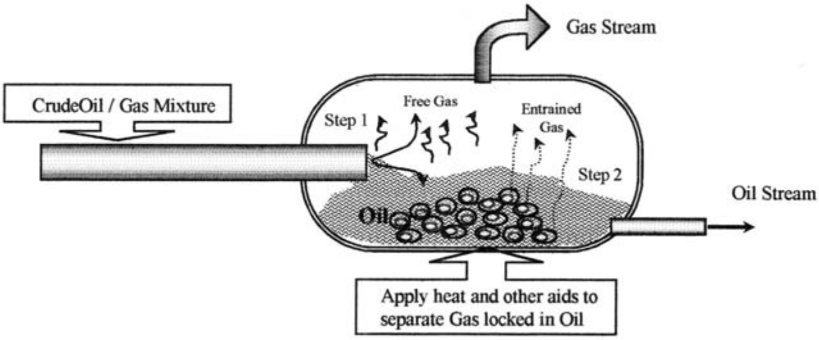


Figure 10 Two-step mechanism of separating gas from oil.

In separating the gas from oil, a mechanical mechanism could be suggested [3], as shown in Figure 10, which implies the following two steps:

(a) To separate oil from gas: Here, we are concerned primarily with recovering as much oil as we can from the gas stream. Density difference or gravity differential is responsible for this separation. At the separator's operating condition of high pressure, this difference in density between oil and gas becomes small (gas law). Oil is about eight times as dense as the gas. This could be a sufficient driving force for the liquid particles to separate and settle down. This is especially true for large-sized particles, having diameter of 100 μm or more. For smaller ones, mist extractors are needed.

(b) To remove gas from oil: The objective here is to recover and collect any non solution gas that may be entrained or "locked" in the oil. Recommended methods to achieve this are settling, agitation, and applying heat and chemicals.

3.5.2 Commercial Types of Gas–Oil Separator

Based on the configuration, the most common types of separator are horizontal, vertical, and spherical, as illustrated in Figures 11, 12, and 13, respectively. A concise comparison among these three types is presented in Table 3. Large horizontal gas–oil separators are used almost exclusively in processing well fluids in the Middle East, where the gas–oil ratio of the producing fields is high. Multistage GOSPs normally consists of three or more separators.

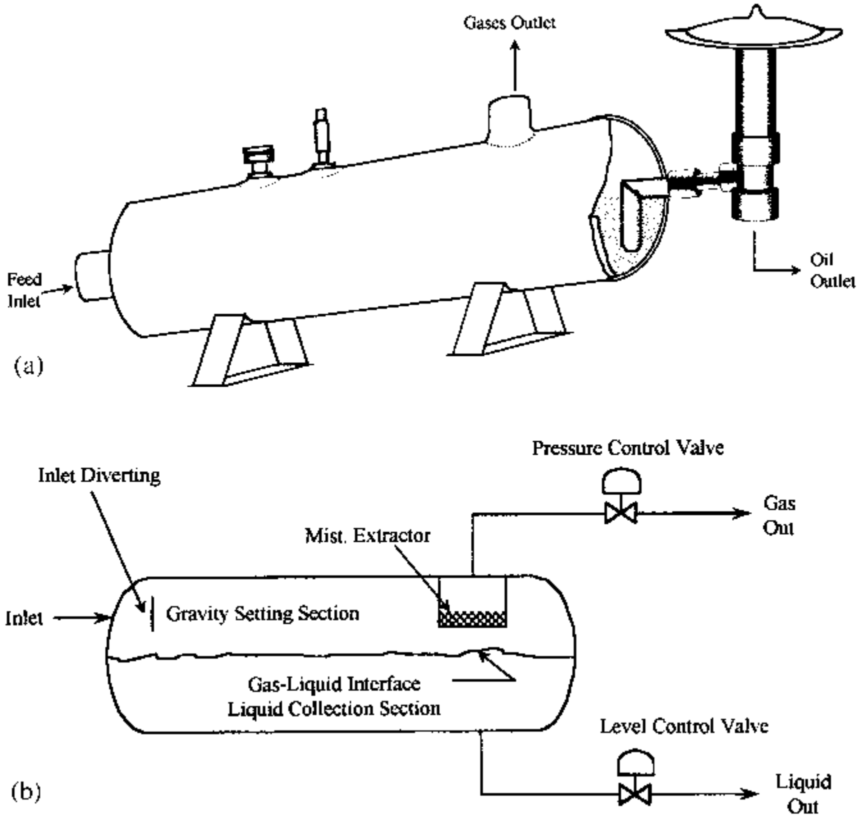


Figure 11 (a) Single-barrel horizontal separator; (b) horizontal separator schematic.

The following is a brief description of some separators for some specific applications. In addition, the features of what is known as “modern” GOSP are highlighted.

Test Separators

These units are used to separate and measure at the same time the well fluids. Potential test is one of the recognized tests for measuring the quantity of both oil and gas produced by the well in 24 hours period under steady state of operating conditions. The oil produced is measured by a flow meter (normally a turbine meter) at the separator’s liquid outlet and the cumulative oil production is measured in the receiving tanks.

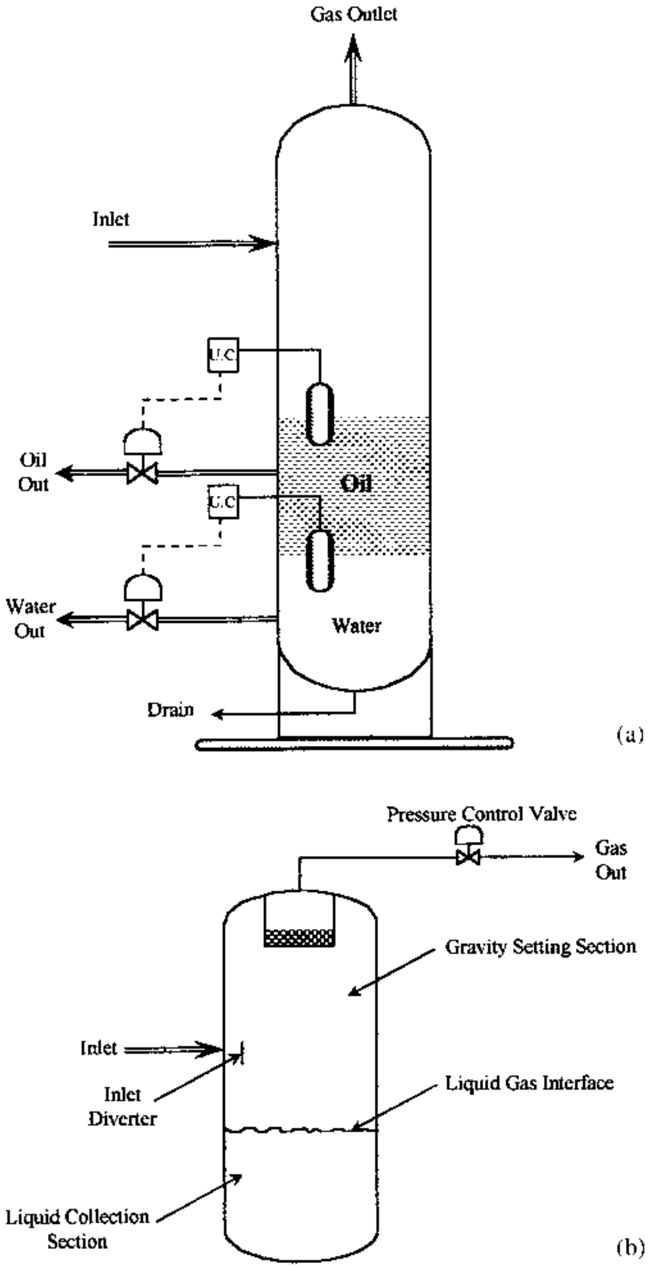


Figure 12 (a) Vertical separator, three-phase operation; (b) vertical separator schematic.

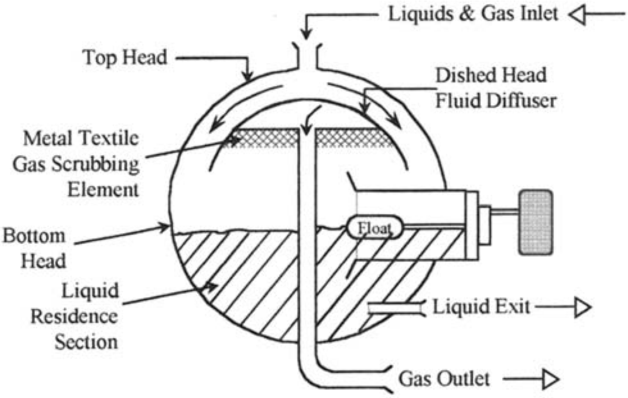
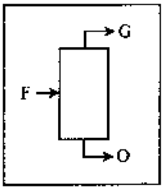
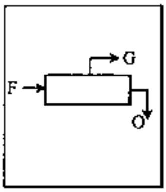
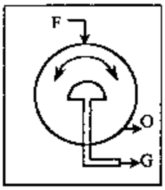


Figure 13 Spherical separator.

Table 3 Comparison Among Different Configurations of Gas–Oil Separators

Function	Vertical	Horizontal	Spherical
Usage	For low gas–oil ratio	For high gas–oil ratio	For small leases operating at moderate pressure
Location of inlet and outlet streams			
Capacity or efficiency	Large fluid capacity	Large gas capacity (handles high GOP)	Capacity rated less (low efficiency)
Handling foreign material	Rated No. 1	Rated No. 3	Rated No. 2
Separation efficiency	Rated No. 2	Rated No. 1	Rated No. 3
Ranking in use in Middle East	Rated No. 2	Rated No. 1	Rated No. 3
Handling foaming oil	Rated No. 2	Rated No. 1	Rated No. 3
Maintenance and inspection	Very difficult	Accessible	Average
Cost per unit capacity	Average	Least expensive	Most expensive
Installation	Most difficult	Average	Easy

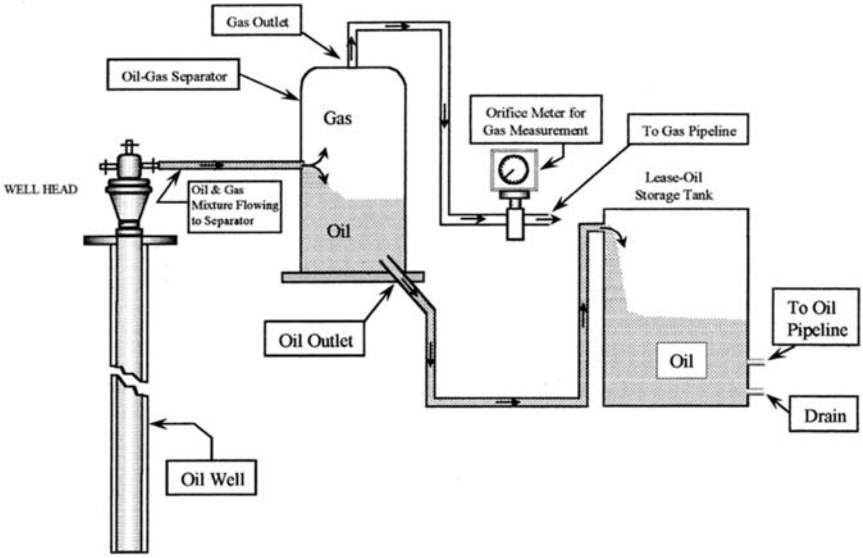


Figure 14 Main equipment for test separator.

An orifice meter at the separator's gas outlet measures the produced gas. Physical properties of the oil and GOR are also determined. Equipment for test units is shown in Figure 14.

Low-Temperature Separators

Low-temperature separators (LTSs) are used to effectively remove light condensable hydrocarbons from a high-pressure gas stream (gas condensate feed). Liquid (condensate) separation is made possible by cooling the gas stream before separation. Temperature reduction is obtained by what is known as the Joule–Thomson effect of expanding the well fluid as it flows through the pressure-reducing choke or valve into the separator. Condensation of the vapors takes place accordingly, where the temperature is in the range 0–10°F.

Modern GOSPs

Safe and environmentally acceptable handling of crude oils is assured by treating the produced crude in the GOSP and related crude-processing facilities. The number one function of the GOSP is to separate the associated gas from oil. As the water content of the produced crude increases, field facilities for control or elimination of water are to be added. This identifies the second function of a GOSP. If the effect of corrosion due to high salt content in the crude is recognized, then modern desalting equipment could be included as a third function in the GOSP design.

One has to differentiate between “dry” crude and “wet” crude. The former is produced with no water, whereas the latter comes along with water. The water produced with the crude is a brine solution containing salts (mainly sodium chloride) in varying concentrations.

The input of wet crude oil into a modern GOSP consists of the following:

1. Crude oil.
2. Hydrocarbon gases.
3. Free water dispersed in oil as relatively large droplets, which will separate and settle out rapidly when wet crude is retained in the vessel.
4. Emulsified water, dispersed in oil as very small droplets that do not settle out with time. Each of these droplets is surrounded by a thin film and held in suspension.
5. Salts dissolved in both free water and in emulsified water.

The functions of a modern GOSP could be summarized as follows:

1. Separate the hydrocarbon gases from crude oil
2. Remove water from crude oil
3. Reduce the salt content to the acceptable level [basic sediments and water]

It should be pointed out that some GOSPs do have gas compression and refrigeration facilities to treat the gas before sending it to gas processing plants. In general, a GOSP can function according to one of the following process operation:

1. Three-phase, gas–oil–water separation (See Chap. 4)
2. Two-phase, gas–oil separation
3. Two-phase, oil–water separation
4. Deemulsification
5. Washing
6. Electrostatic coalescence

To conclude, the ultimate result in operating a modern three-phase separation plant is to change “wet” crude input into the desired outputs, as given in Figure 15. Outputs from a three-phase regular separation plant, on the other hand, are as shown in Figure 16.

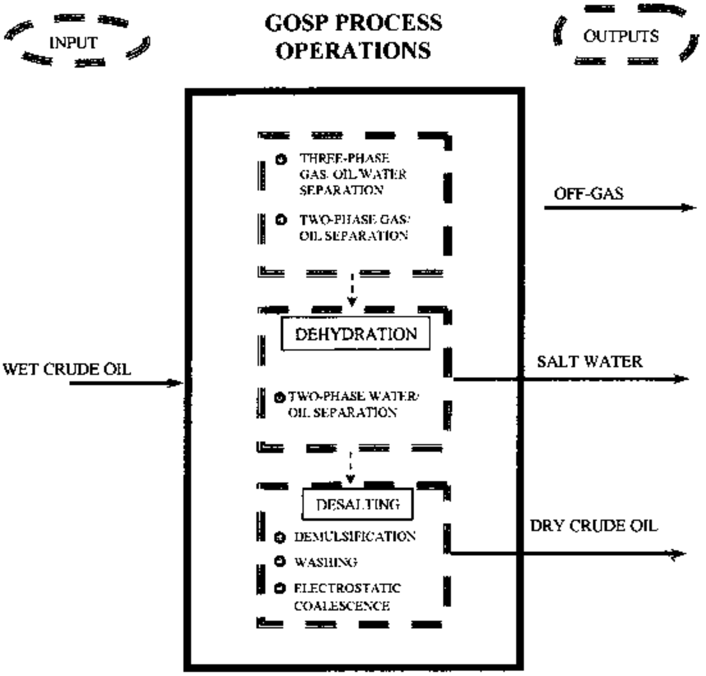


Figure 15 Functions of modern GOSPs.

3.5.3 Controllers and Internal Components of Gas–Oil Separators

Gas–oil separators are generally equipped with the following control devices and internal components [4].

Liquid Level Controller

The liquid level controller (LLC) is used to maintain the liquid level inside the separator at a fixed height. In simple terms, it consists of a float that exists at the liquid–gas interface and sends a signal to an automatic diaphragm motor valve on the oil outlet. The signal causes the valve to open or close, thus allowing more or less liquid out of the separator to maintain its level inside the separator.

Pressure Control Valve

The pressure control valve (PCV) is an automatic backpressure valve that exists on the gas stream outlet. The valve is set at a prescribed pressure.

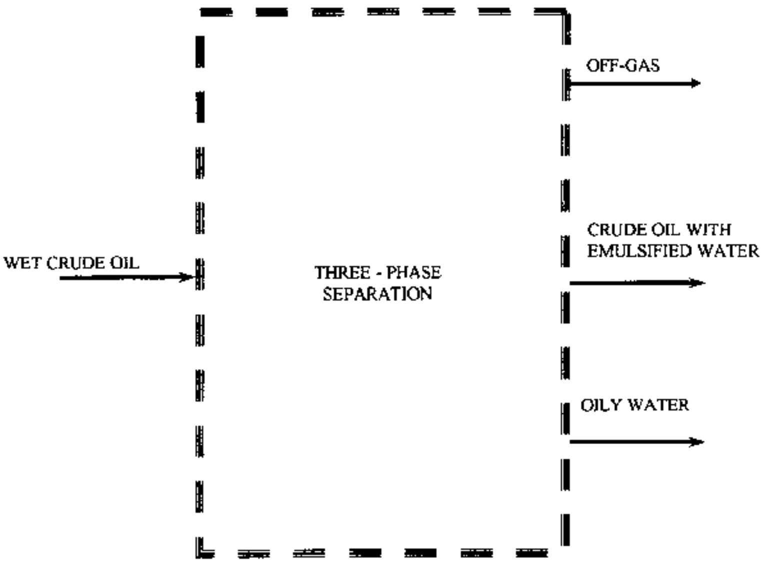


Figure 16 Products obtained from a three-phase GOSP.

It will automatically open or close, allowing more or less gas to flow out of the separator to maintain a fixed pressure inside the separator.

Pressure Relief Valve

The pressure relief valve (PRV) is a safety device that will automatically open to vent the separator if the pressure inside the separator exceeded the design safe limit.

Mist Extractor

The function of the mist extractor is to remove the very fine liquid droplets from the gas before it exits the separator. Several types of mist extractors are available:

1. *Wire-Mesh Mist Extractor:* These are made of finely woven stainless-steel wire wrapped into a tightly packed cylinder of about 6 in. thickness. The liquid droplets that did not separate in the gravity settling section of the separator coalesce on the surface of the matted wire, allowing liquid-free gas to exit the separator. As the droplets size grows, they fall down into the liquid phase. Provided that the gas velocity is reasonably low,

wire-mesh extractors are capable of removing about 99% of the 10-μm and larger liquid droplets. It should be noted that this type of mist extractor is prone to plugging. Plugging could be due to the deposition of paraffin or the entrainment of large liquid droplets in the gas passing through the mist extractor (this will occur if the separator was not properly designed). In such cases, the vane-type mist extractor, described next, should be used.

2. *Vane Mist Extractor:* This type of extractor consists of a series of closely spaced parallel, corrugated plates. As the gas and entrained liquid droplets flowing between the plates change flow direction, due to corrugations, the liquid droplets impinge on the surface of the plates, where they coalesce and fall down into the liquid collection section.
3. *Centrifugal Mist Extractor:* This type of extractor uses centrifugal force to separate the liquid droplets from the gas. Although it is more efficient and less susceptible to plugging than other extractors, it is not commonly used because of its performance sensitivity to small changes in flow rate.

Inlet Diverters

Inlet diverters are used to cause the initial bulk separation of liquid and gas. The most common type is the baffle plate diverter, which could be in the shape of a flat plate, a spherical dish, or a cone. Another type, is the centrifugal diverter; it is more efficient but more expensive. The diverter provides a means to cause a sudden and rapid change of momentum (velocity and direction) of the entering fluid stream. This, along with the difference in densities of the liquid and gas, causes fluids separation.

Wave Breakers

In long horizontal separators, waves may develop at the gas–liquid interface. This creates unsteady fluctuations in the liquid level and would negatively affect the performance of the liquid level controller. To avoid this, wave breakers, which consist of vertical baffles installed perpendicular to the flow direction, are used.

Defoaming Plates

Depending on the type of oil and presence of impurities, foam may form at the gas–liquid interface. This results in the following serious operational problems:

1. Foam will occupy a large space in the separator that otherwise would be available for the separation process; therefore, the separator efficiency will be reduced unless the separator is oversized to allow for the presence of foam.
2. The foam, having a density between that of the liquid and gas, will disrupt the operation of the level controller.
3. If the volume of the foam grows, it will be entrained in the gas and liquid streams exiting the separator; thus, the separation process will be ineffective. The entrainment of liquid with the exiting gas is known as *liquid carryover*. Liquid carryover could also occur as a result of a normally high liquid level, a plugged liquid outlet, or an undersized separator with regard to liquid capacity. The entrainment of gas in the exiting liquid is known as *gas blowby*. This could also occur as a result of a normally low liquid level, an undersized separator with regard to gas capacity, or formation of a vortex at the liquid outlet.

Foaming problems may be effectively alleviated by the installation of defoaming plates within the separator. Defoaming plates are basically a series of inclined closely spaced parallel plates. The flow of the foam through such plates results in the coalescence of bubbles and separation of the liquid from the gas.

In some situations, special chemicals known as foam depressants may be added to the fluid mixture to solve foaming problems. The cost of such chemicals could, however, become prohibitive when handling high production rates.

Vortex Breaker

A vortex breaker, similar in shape to those used in bathroom sink drains, is normally installed on the liquid outlet to prevent formation of a vortex when the liquid outlet valve is open. The formation of a vortex at the liquid outlet may result in withdrawal and entrainment of gas with the exiting liquid (gas blowby).

Sand Jets and Drains

As explained in Chapter 1, formation sand may be produced with the fluids. Some of this sand will settle and accumulate at the bottom of the separator. This takes up separator volume and disrupts the efficiency of separation. In such cases, vertical separators will be preferred over horizontal separators. However, when horizontal separators are needed, the separator should be equipped with sand jets and drains along the bottom of the separator. Normally, produced water is injected though the jets to fluidize the accumulated sand, which is then removed through the drains.

3.6 DESIGN PRINCIPLES AND SIZING OF GAS–OIL SEPARATORS

In this section, some basic assumptions and fundamentals used in sizing gas–oil separators are presented first. Next, the equations used for designing vertical and horizontal separators are derived. This will imply finding the diameter and length of a separator for given conditions of oil and gas flow rates, or vice versa. Solved examples are also given to illustrate the use of these equations.

3.6.1 Assumptions

1. No oil foaming takes place during the gas–oil separation (otherwise retention time has to be drastically increased as explained earlier).
2. The cloud point of the oil and the hydrate point of the gas are below the operating temperature.
3. The smallest separable liquid drops are spherical ones having a diameter of 100 μm.
4. Liquid carryover with the separated gas does not exceed 0.10 gallon/MMSCF (M = 1000).

3.6.2 Fundamentals

1. The difference in densities between liquid and gas is taken as a basis for sizing the gas capacity of the separator ($\rho_o - \rho_g$).
2. A normal liquid (oil) retention time for gas to separate from oil is between 30 s and 3 min. Under foaming conditions, more time is considered (5–20 min). Retention time is known also as the residence time ($= V/Q$, where V is the volume of vessel occupied by oil and Q is the liquid flow rate).
3. In the gravity settling section, liquid drops will settle at a terminal velocity that is reached when the gravity force F_g acting on the oil drop balances the drag force (F_d) exerted by the surrounding fluid or gas.

4. For vertical separators, liquid droplets (oil) separate by settling downward against an up-flowing gas stream; for horizontal ones, liquid droplets assume a trajectory like path while it flows through the vessel (the trajectory of a bullet fired from a gun).
5. For vertical separators, the gas capacity is proportional to the cross-sectional area of the separator, whereas for horizontal separators, gas capacity is proportional to area of disengagement (*LD*) (i.e., length × diameter).

3.6.3 Settling of Oil Droplets

In separating oil droplets from the gas in the gravity settling section of a separator, a relative motion exists between the particle, which is the oil droplet, and the surrounding fluid, which is the gas.

An oil droplet, being much greater in density than the gas, tends to move vertically downward under the gravitational or buoyant force, F_g. The fluid (gas), on the other hand, exerts a drag force, F_d, on the oil droplet in the opposite direction. The oil droplet will accelerate until the frictional resistance of the fluid drag force, F_d, approaches and balances F_g; and, thereafter, the oil droplet continues to fall at a constant velocity known as the *settling* or *terminal velocity*.

The drag force, F_d, is proportional to the droplet surface area perpendicular to the direction of gas flow, and its kinetic energy per unit volume. Hence,

$$F_d = C_d \frac{\pi}{4} d^2 \frac{\rho_g u^2}{2} \tag{1}$$

whereas F_g is given by

$$F_g = \frac{\pi}{6} d^3 (\rho_o - \rho_g) g \tag{2}$$

where C_d is the drag coefficient, d is the diameter of the oil droplet (ft), u is the settling velocity of the oil droplet (ft/s), ρ_o and ρ_g are the oil and gas densities, (lb/ft^3), respectively, and, g is gravitational acceleration (ft/s^2).

The settling terminal velocity, u, is reached when $F_d = F_g$. Therefore, equating Eqs. (1) and (2) and solving for u, the droplet settling velocity, we obtain

$$u^2 = \frac{8}{6} g \frac{(\rho_o - \rho_g)}{\rho_g} \left(\frac{d}{C_d} \right)$$

The droplet diameter d is normally expressed in microns, where 1 μm is equal to 3.2808×10^{-6} ft. Let d_m be the droplet diameter (in μm) and substitute 32.17 for g in the above equation to obtain the following expression for the settling velocity:

$$u = 0.01186\left[\left(\frac{\rho_o - \rho_g}{\rho_g}\right)\frac{d_m}{C_d}\right]^{1/2} \qquad \text{ft/s} \tag{3}$$

As mentioned earlier, in designing gas–oil separators, the smallest oil droplet to be removed from the gas in the gravity settling section is normally taken as 100 μm. Under such a condition, the mist extractor will be capable of removing oil droplets smaller than 100 μm without getting flooded.

There are other special separators known as *gas scrubbers* that are normally used to remove liquid from gas streams, which have been through regular gas–liquid separators. Such gas streams are mostly gas with little liquid that may have formed due to condensation. Examples of such separators are the gas scrubbers used at the inlet to gas compressors and gas dehydration facilities. Scrubbers are also used on vents and flares. Because the amount of liquid is very small, the design of such gas scrubbers could be based on separation of liquid droplets up to 500 μm in the gravity settling section with no danger of flooding the mist extractor.

3.6.4 Gas Capacity of Separators

The volumetric flow rate of the gas processed by a gas–oil separator is directly related to the cross-sectional area of flow and the maximum allowable gas velocity at which the oil droplets will be suspended and not carried over. Mathematically,

$$Q_g = A_g u \qquad \text{ft}^3/\text{s} \tag{4}$$

Equation (4) gives Q_g in ft^3/s under actual separator pressure and temperature. However, the volumetric flow rate of gas is normally reported at standard pressure and temperature of 14.7 psia and 520°R, respectively. Typically, Q_g is reported in units of millions of standard cubic feet per day (MMSCFD). Equation (4) could, therefore, be written as

$$\begin{aligned} Q_g &= (10^{-6} \times 60 \times 60 \times 24)A_g u\left(\frac{P}{14.7}\right)\left(\frac{520}{TZ}\right) \\ Q_g &= 3.056\left(\frac{P}{TZ}\right)uA_g \qquad \text{MMSCFD} \end{aligned} \tag{5}$$

Solving for the gas velocity, u, gives

$$u = 0.327 Q_g \left(\frac{TZ}{P}\right)\left(\frac{1}{A_g}\right) \frac{\text{ft}}{\text{s}} \tag{6}$$

where, Z is the gas compressibility at the operating pressure (P) and temperature (T) and A_g is the available area for gas flow (ft^2).

3.6.5 Liquid Capacity of Separators

The basic relationship that combines the oil flow rate or oil capacity of a separator, Q_o, the volume of separator occupied by oil, V_o, and the retention time or residence time, t, is

$$Q_o\left(\frac{\text{ft}^3}{\text{min}}\right) = \frac{V_o\,(\text{ft}^3)}{t(\text{min})} \tag{7}$$

Equation (7) is rewritten in terms of barrels per day instead of cubic feet per minute:

$$Q_o\left(\frac{\text{bbl}}{\text{day}}\right) = 257\frac{V_o}{t} \tag{8}$$

where $1\,\text{ft}^3/\text{min} = 257\,\text{bbl/day}$.

3.6.6 Sizing Vertical Gas–Oil Separators

The size (diameter and height or length) of a separator is normally determined by consideration of its required capacity for gas and oil as discussed in the following subsections.

Gas Capacity Constraint

For vertical separators, the upward average gas velocity should not exceed the downward terminal velocity of the smallest oil droplet to be separated. This condition is expressed mathematically by equating Eqs. (2) and (6):

$$0.327 Q_g\left(\frac{TZ}{P}\right)\left(\frac{1}{A_g}\right) = 0.01186\left[\left(\frac{\rho_o - \rho_g}{\rho_g}\right)\frac{d_m}{C_d}\right]^{1/2}$$

Substituting for A_g from

$$A_g = \frac{\pi}{4}\left(\frac{D}{12}\right)^2$$

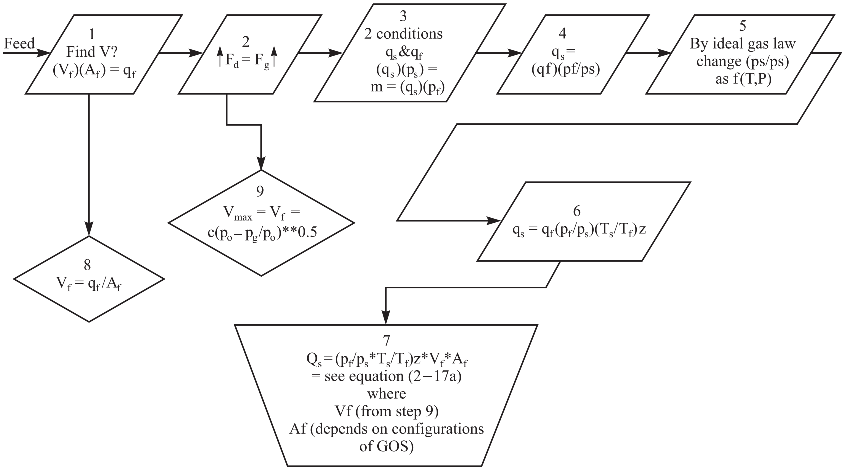


Figure 17 Outline of the calculated procedure for determining the gas capacity of GOSPs.

where D is the internal diameter of the separator in inches, and then solving for D, we obtain

$$D^2 = 5.058 Q_g \left(\frac{TZ}{P}\right) \left[\frac{\rho_g}{(\rho_o - \rho_g)} \frac{C_d}{d_m}\right]^{1/2} \text{in.}^2 \tag{9}$$

Equation (9) provides the minimum acceptable diameter of the separator. Larger diameters yield lower gas velocities and, thus, better separation of the oil droplets from the gas. Smaller diameters, on the other hand, result in higher gas velocities and, therefore, the liquid droplets will be carried over with the gas.

A summary of the steps involved in the calculation of the gas capacity of a gas–oil separator is outlined in Figure 17.

In solving Eq. (9), the value of the drag coefficient C_d must first be determined. C_d is related to the Reynolds number, Re, according to the following formula [4]:

$$C_d = 0.34 + \frac{3}{\text{Re}^{0.5}} + \frac{24}{\text{Re}} \tag{10}$$

where the Reynolds number is given by

$$\text{Re} = 0.0049 \frac{\rho_g d_m u}{\mu_g} \tag{11}$$

where μ_g is the gas viscosity.

The velocity, u, is given by Eq. (2) and is a function of C_d. Therefore, C_d could only be determined by an iterative procedure as follows:

1. Assume a value for C_d (a value of 0.34 could be used as a first assumption).
2. Calculate the velocity, u, from Eq. (2).
3. Calculate Re from Eq. (11).
4. Calculate C_d from Eq. (10) and compare to the assumed value.
5. If no match is obtained, use the calculated value of C_d and repeat steps 2–4 until convergence is obtained.

Oil Capacity Constraint

The oil has to be retained within the separator for a specific retention time, t. The volume of separator occupied by oil, V_o, is obtained by multiplying the cross-sectional area by the height of the oil column, H (in.). Equation (8) could, therefore, be rewritten as

$$Q_o = 257\left(\frac{\pi}{4}\right)\left(\frac{D}{12}\right)^2\left(\frac{H}{12}\right)\left(\frac{1}{t}\right)\left(\frac{\text{bbl}}{\text{day}}\right) \tag{12}$$

or

$$D^2H = 8.565 Q_o t \text{ in.}^3 \tag{13}$$

Sizing Procedure

In summary then, the size (diameter and seam-to-seam length or height) of a vertical separator is determined as follows:

1. Equation (9) is used to determine the minimum allowable vessel diameter.
2. For diameters larger than the minimum, Eq. (13) is used to determine combinations of D and H.
3. The seam-to-seam length, L_s, for each combination of D and H is determined using one of the following expressions as appropriate:

   $D < 36$ in.

   $$L_s = \frac{H + 76}{12} \text{ ft} \tag{14}$$

   $D > 36$ in.

   $$L_s = \frac{H + D + 40}{12} \text{ ft} \tag{15}$$

4. For each combination of D and L_s, the *slenderness ratio*, SR, defined as the ratio of length to diameter is determined. Separators with SR between 3 and 4 are commonly selected.

3.6.7 Sizing Horizontal Gas–Oil Separators

As with vertical separators, the size (diameter and length) of the horizontal separator is determined by consideration of its required capacity for gas and oil. It has been shown that the gas capacity constraint for vertical separators determines the minimum allowable vessel diameter. For horizontal separators, however, the gas capacity constraint yields, as shown in the following subsection, a relationship between the diameter and effective length of the separator. This along with a similar relationship derived from the liquid capacity constraint are used in determining the size of the separator. In reality, either the gas capacity constraint or the liquid capacity constraint governs the design and only one of the two constraints equations is used in determining the size.

In the following discussion, and as mentioned earlier, it is assumed that each of the gas and oil phases occupies 50% of the effective separator volume. Similar equations as those derived below could be obtained for other situations where either of the two phases occupies more or less than 50% of the separator effective volume.

Gas Capacity Constraint

Because the gas occupies the top half of the separator, its average flowing velocity within the separator, u_g, is obtained by dividing the volumetric flow rate, Q_g, by one-half of the separator cross-sectional area, A; that is,

$$u_g = \frac{Q_g}{0.5[(\pi/4)D^2]}$$

Q_g is usually reported in units of MMSCFD and should, therefore, be converted into actual ft^3/s; also D, which is usually given in inches, should be converted into feet in order to obtain the velocity in units of ft/s. The above equation, therefore, becomes

$$u_g = 120\frac{Q_g}{D^2}\left(\frac{TZ}{P}\right) \qquad \frac{\text{ft}}{\text{s}} \tag{16}$$

The gas travels horizontally along the effective length of the separator, L(ft), in a time t_g that is given by

$$t_g = \frac{L}{u_g} \text{ s} \tag{17}$$

This time must, at least, be equal to the time it takes the smallest oil droplet, to be removed from the gas, to travel a distance of $D/2$ to reach the gas–oil interface. This settling time, t_s, is obtained by dividing the distance ($D/2$) by the settling velocity from Eq. (2); therefore,

$$t_s = \left(\frac{D}{2 \times 12}\right)\left\{0.01186\left[\left(\frac{\rho_o - \rho_g}{\rho_g}\right)\frac{d_m}{C_d}\right]^{1/2}\right\}^{-1} \text{ s} \tag{18}$$

Equating Eqs. (18) and (17), substituting for u_g from Eq. (16), and solving for the product LD, we obtain

$$LD = 422\left(\frac{Q_g TZ}{P}\right)\left[\left(\frac{\rho_g}{\rho_o - \rho_g}\right)\left(\frac{C_d}{d_m}\right)\right]^{1/2} \text{ ft in.} \tag{19}$$

Equation (19) provides a relationship between the vessel diameter and effective length that satisfies the gas capacity constraint. Any combination of D and L satisfying Eq. (19) ensures that all oil droplets having diameter d_m and larger will settle out of the gas flowing at a rate of Q_g MMSCFD into the separator that is operating at P psia and T°R.

Liquid Capacity Constraint

As explained earlier, the separator has to have a sufficient volume to retain the liquid for the specified retention time before it leaves the separator. For a horizontal separator that is half full of liquid, the volume occupied by the liquid is given by

$$V_o = 0.5\left(\frac{\pi}{4}\right)\left(\frac{D}{12}\right)^2 L \text{ ft}^3$$

Substituting in Eq. (8), the following equation is obtained:

$$D^2 L = 1.428 Q_o t \text{ ft}^3 \tag{20}$$

Equation (20) provides another relationship between D and L that satisfies the liquid capacity (retention) time constraint.

Sizing Procedure

For a given set of operating conditions (pressure, temperature, gas and oil flow rates, gas and oil properties, and oil retention time), the size (diameter and seam-to-seam length) of a horizontal separator is determined as follows:

5. Assume various values for the separator diameter, D.
6. For each assumed value of D, determine the effective length, L_g, that satisfies the gas capacity constraint from Eq. (19) and calculate the seam-to-seam length, L_s, from

$$L_s = L_g + \frac{D}{12}\ \text{ft} \tag{21}$$

7. For each assumed value of D, determine the effective length, L_o, that satisfies the liquid capacity constraint from Eq. (20) and calculate the seam-to-seam length, L_s, from

$$L_s = \frac{4}{3} L_o\ \text{ft} \tag{22}$$

8. For each value of D used, compare the values of L_g and L_o to determine whether the gas capacity constraint or the oil capacity constraint governs the design of the separator. Of course, the larger required length governs the design.
9. Select reasonable combinations of D and L_s such that the slenderness ratio SR is in the range of 3–5. The cost and availability would then determine the final selection.

The above equations and sizing procedures are very sufficient for the determination of separator diameter and length as well as for the performance evaluation of existing separators. The students and practicing engineers should be familiar with such design equations and procedures before attempting the use of commercially available software. The retention time is an important parameter in designing gas–oil separators. It is best obtained from laboratory tests that simulate the field operating conditions. This, however, may not always be available. In such cases, experience and data from offset fields, if available, will be very valuable.

On the commercial side, computer programs for sizing gas–oil separators have been developed by Ellis Engineering Inc. [1]. Capacity curves based on computer programs have been introduced as well by Smith and Ellis [1]. These curves are suitable for preliminary sizing for estimating gas and liquid capacities.

For spherical separators, curves have been developed [5] for sizing gas capacity as illustrated by the diagram given in Figure 18.

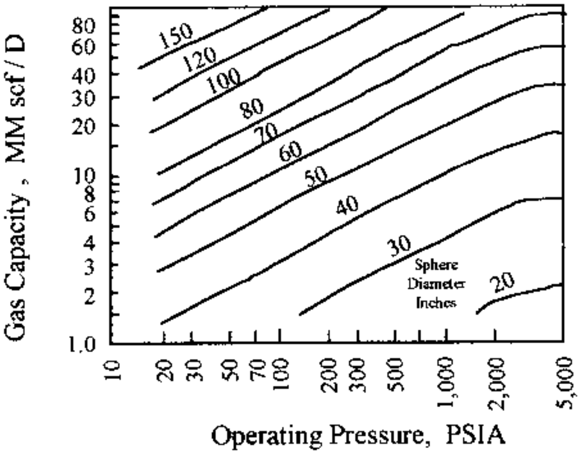


Figure 18 Gas capacity of spherical gas–oil separators.

3.6.8 Solved Examples on Sizing Gas–Oil Separators

Example 1: Design Problem, Vertical Separator

Determine the diameter and height (seam-to-seam length) of a vertical separator for the following operating conditions:

Gas rate:	15 MMSCFD
Gas specific gravity:	0.6
Oil rate:	3000 bbl/day
Oil gravity:	35°API
Operating pressure:	985 psi
Operating temperature:	60°F
Retention time:	3 min

Solution

1. *Determine gas and oil properties:* For 0.6 specific gravity gas at 60°F and 985 psi (1000 psia),

$$Z = 0.84 \qquad \text{and} \qquad \mu = 0.013\,\text{cP}$$

$$\rho_g = 2.7\gamma\frac{P}{TZ} = 2.7(0.6)\frac{1000}{520 \times 0.84} = \frac{3.708\,\text{lb}}{\text{ft}^3}$$

$$\rho_o = \rho_w\gamma_o = 62.4\frac{141.5}{131.5 + 35} = \frac{53.03\,\text{lb}}{\text{ft}^3}$$

2. *Determine C_d:* Assume $C_d = 0.34$, Using Eq. (2),

$$u = 0.01186\left[\left(\frac{\rho_o - \rho_g}{\rho_g}\right)\frac{d_m}{C_d}\right]^{1/2}$$

$$= 0.01186\left[\left(\frac{53.03 - 3.708}{3.708}\right)\left(\frac{100}{0.34}\right)\right]^{1/2}$$

$$u = 0.7418\ \frac{\text{ft}}{\text{s}}$$

$$\text{Re} = 0.0049\frac{\rho_g d_m u}{\mu_g} = 0.0049\frac{3.708 \times 100 \times 0.7418}{0.013} = 103.67$$

$$C_d = 0.34 + \frac{3}{\text{Re}^{0.5}} + \frac{24}{\text{Re}}$$

$$= 0.34 + \frac{3}{(103.67)^{0.5}} + \frac{24}{103.67} = 0.866$$

3. Repeat the above steps using the calculated value of C_d to obtain a new value of c_d and continue until convergence. The final value of c_d is 1.1709.
4. *Check for gas capacity constraint, Eq.* (9):

$$D^2 = 5,058 Q_g\left(\frac{TZ}{P}\right)\left[\frac{\rho_g}{(\rho_o - \rho_g)}\frac{C_d}{d_m}\right]^{1/2}$$

$$= 5058 \times 15\left(\frac{520 \times 0.84}{1000}\right)\left(\frac{3.708}{53.03 - 3.708}\frac{1.1709}{100}\right)^{1/2}$$

$$D^2 = 983.246;$$

therefore,

$$D_{\min} = 31.357\,\text{in.} \tag{E1}$$

This is the minimum allowable vessel diameter for separation of oil droplets down to 100 μm.

5. *Check for liquid capacity (retention time constraint), Eq. (13):*

$$D^2 H = 8.565 Q_o t = 8.565 \times 3000 \times 3 = 77,805 \tag{E2}$$

6. Select values for D greater than 31.357 in. and calculate the corresponding values of H from Eq. (E2) and the corresponding values of L_s from Eq. (14) for $D < 36$ or Eq. (15) for $D > 36$. The results are as follows:
From the results in the table, the first two diameters provide slenderness ratios within the commonly used range; therefore,

D (in.)	H (in.)	L_s(ft)	SR
36	60.03	11.34	3.78
42	44.10	10.51	3.00
48	33.77	10.15	2.54
54	26.68	10.06	2.23
60	21.61	10.13	2.03
66	17.86	10.32	1.88
72	15.01	10.58	1.76

both are acceptable. A preferred selection will be a 36-in. diameter by 12-ft separator as compared to a 42-in. by 11-ft separator. This is because the 36-in. separator is a standard unit and would probably be less costly than the larger-diameter, shorter separator.

Example 2: Performance Problem, Vertical Separator

Determine the actual gas and oil capacity of a 36-in. -diameter by 12-ft seam-to-seam length vertical separator operating under the following conditions:

Gas specific gravity:	0.6
Oil gravity:	35°API
Operating pressure:	985 psi
Operating temperature:	60°F
Retention time:	3 min
C_d:	1.1709

Solution

1. *To determine the actual gas capacity:* Rearranging Eq. (9) to solve for Q_g, we obtain

$$Q_g = 1.977 \times 10^{-4} D^2 \left(\frac{P}{TZ}\right)\left(\frac{(\rho_o - \rho_g)d_m}{\rho_g C_d}\right)^{1/2}$$

Because we have the same operating conditions of Example 1, we can substitute the same values of ρ_g, ρ_o, and Z; therefore,

$$Q_g = 1.977 \times 10^{-4}(36)^2\left(\frac{1000}{520 \times 0.84}\right)\left(\frac{(53.03 - 3.708)100}{3.708 \times 1.1709}\right)^{1/2}$$

2. *The determination of the pressure drop across each segment of the flow system*. This is a very complex problem, as it involves

the simultaneous flow of oil, gas, and water (multiphase flow), which makes the pressure drop dependent on many variables, some of which are interdependent. There is no analytical solution to this problem. Instead, empirical correlations and mechanistic models have been developed and used for predicting the pressure drop in multiphase flow. Computer programs based on such correlations and models are now available for the determination of pressure drops in vertical, inclined, and horizontal pipes.

3. *Sizing tubing and flow line:* Determination of the tubing and flow line sizes is a complex process involving the determination and prediction of future well productivity, analysis of multiphase flow under varying production conditions, and economic analysis. However, a simplified approach is summarized in the following steps and illustrated in Figure 18:
   - (i) Determine (predict) the present and future IPR. Plot the results as P_{wf} versus q.
   - (ii) Selecting tubing and flow line diameters and starting with a specified value for the surface separator pressure, determine the flowing bottom-hole pressure for a specified production rate, water cut (W.C.) and gas–oil ratio (GOR), using available multiphase flow correlations or models.

Determine the diameter and seam-to-seam length of a horizontal separator for the following operating conditions. Determine the actual gas and oil capacity of the designed separator.

Gas rate:	15 MMSCFD
Gas specific gravity:	0.6
Oil rate:	3,000 bbl/day
Oil gravity:	35°API
Operating pressure:	985 psi
Operating temperature:	60°F
Retention time:	3 min

Solution

From Example 1, we have

$$\mu_g = 0.013\ \mathrm{cP}$$
$$\rho_g = 3.708\ \mathrm{lb/ft^3}$$
$$\rho_o = 53.03\ \mathrm{lb/ft^3}$$
$$Z = 0.84$$
$$c_d = 1.1709$$

1. *Check for gas capacity constraint,* Eq. (18):

$$LD = 422\left(\frac{Q_g TZ}{P}\right)\left[\left(\frac{\rho_g}{\rho_o - \rho_g}\right)\left(\frac{c_d}{d_m}\right)\right]^{1/2}$$

$$= 422\left(\frac{15 \times 520 \times 0.84}{1000}\right)\left[\left(\frac{3.708}{53.03 - 3.708}\right)\left(\frac{1.1709}{100}\right)\right]^{1/2}$$

$$LD = 82.04 \tag{E3}$$

2. *Check for oil capacity (retention time),* Eq. (20):

$$D^2L = 1.428Q_o t = 1.428 \times 3000 \times 3$$

$$D^2L = 12{,}852 \tag{E4}$$

3. Assume values for D and determine corresponding effective length for gas capacity, L_g, from Eq. (E3) and seam-to-seam length from Eq. (21). For each assumed value of D, determine the corresponding effective length for oil capacity, L_o, from Eq. (E4) and seam-to-seam length from Eq. (22). The results are summarized as follows:

D (in.)	L_g (ft)	L_s (gas)	L_o (ft)	L_s (oil)	$\mathrm{SR} = 121_s$ (oil)$/D$
30	2.73	5.23	14.28	19.04	7.62
36	2.28	5.28	9.92	13.22	4.41
42	1.95	5.45	7.29	9.71	2.78
48	1.71	5.71	5.58	7.44	1.86
54	1.52	6.02	4.41	5.88	1.31

Comparing the value of L_s for the oil capacity to those for the gas capacity shows that the gas capacity does not govern the design. Investigating the values of the slenderness ratio shows that the 36-in. and 42-in. separators are the only possible selections. The recommended size would be a 36-in. diameter by 14-ft seam-to-seam length.

Recall that for the same conditions a vertical separator of the same diameter but shorter (12-ft) was suitable (Example 1). For such conditions, the vertical separator should be selected unless other operating conditions necessitate the selection of a horizontal separator.

Example 4: Performance Problem, Horizontal Separator

Determine the actual gas and oil capacity of a horizontal separator having a diameter of 36 in. and a seam-to-seam length of 14 ft given the following operating conditions:

Gas Specific gravity:	0.6
Oil gravity:	35°API
Operating pressure:	985 psi
Operating temperature:	60°F
Retention time:	3 min
C_d:	1.1709

Solution

From Example 1, because we have the same operating conditions, the fluid properties will be the same:

$$\rho_g = \frac{3.708\,\mathrm{lb}}{\mathrm{ft}^3}$$

$$\rho_o = \frac{53.03\,\mathrm{lb}}{\mathrm{ft}^3} \quad \text{and} \quad Z = 0.84$$

1. *To determine the actual gas capacity of the separator:* Determine the separator effective length for gas capacity from Eq. (21):

$$L_s = L_g + \frac{D}{12}$$
$$L_g = 14 - (12/36) = 14 - 3 = 11\,\mathrm{ft}$$

Rearranging Eq. (14) to solve for Q_g, we have

$$Q_g = 0.00237 L_g D\left(\frac{P}{TZ}\right)\left[\left(\frac{\rho_o - \rho_g}{\rho_g}\right)\left(\frac{d_m}{C_d}\right)\right]^{1/2}$$

Substitute $L_g = 11$ and $D = 36$ in the above equation and solve for Q_g:

$$Q_g = 0.00237(11 \times 36)\left(\frac{1000}{520 \times 0.84}\right) \times \left[\left(\frac{53.03 - 3.708}{53.03}\right)\left(\frac{100}{1.1709}\right)\right]^{1/2}$$
$$Q_g = 72.4\,\mathrm{MMSCFD}$$

2. *To determine the actual oil capacity of the separator:* Determine the separator effective length for oil capacity from Eq. (22):

$$L_o = 34 L_s = 3 \times 414 = 10.5\,\mathrm{ft}$$

Rearrange Eq. (20) to solve for Q_o:

$$Q_o = 0.7\left(\frac{D^2 L}{t}\right) D^2 L$$

Substitute $L_o = 10.5$ and $D = 36$ in the above equation and solve for Q_o:

$$Q_o = 3.176 \,\text{bbl/day}$$

Comparing the above actual oil and gas capacities to the design production rates of Example 3 shows that the separator can handle almost five times the design gas rate of Example 3. This is because the cross-sectional area for gas flow is much more than needed for settling. In such situations, it is evident that the assumption of allowing each phase to occupy one-half of the separator effective volume is not a good assumption. The size of the separator could certainly be reduced if the design was based on allowing the gas to flow through a smaller area than the 50% of the total area. It is recommended that the reader solve this problem assuming the gas to occupy a smaller volume than 50% of the separator volume. Care must be taken in modifying the above design equations used.

Geometrical relationships relating the cross-sectional area occupied by the gas, the height of the gas column, and the vessel diameter could be derived. For the cases where the cross-sectional area of the gas flow, A_g, is not equal to one-half of the vessel cross-sectional area, A; that is, the height of the gas, H_g, is not equal to the radius, $D/2$, Lockhart [6] presented the following simplified, but reasonably accurate, relationships.

(a) To determine H_g/D as a function of A_g/A:

$0 \le A_g/A \le 0.2$:

$$\frac{H_g}{D} = 2.481\frac{A_g}{A} - 12.29\left(\frac{A_g}{A}\right)^2 + 31.133\left(\frac{A_g}{A}\right)^3$$

$0.2 \le A_g/A \le 0.8$:

$$\frac{H_g}{D} = 0.8123\frac{A_g}{A} + 0.0924$$

$0.8 \le A_g/A \le 1.0$:

$$\left(1 - \frac{H_g}{D}\right) = 2.481\left(1 - \frac{A_g}{A}\right) - 12.29\left(1 - \frac{A_g}{A}\right)^2 + 31.133\left(1 - \frac{A_g}{A}\right)^3$$

(b) To determine A_g/A as a function of H_g/D:

$0 \leq H_g/D \leq 0.25$:

$$\frac{A_g}{A} = 0.21\left(\frac{H_g}{D}\right) + 3.52\left(\frac{H_g}{D}\right)^2 - 4.93\left(\frac{H_g}{D}\right)^3$$

$0.25 \leq H_g/D \leq 0.75$:

$$\frac{A_g}{A} = 1.231\left(\frac{H_g}{D}\right) - 0.1138$$

$0.75 \leq H_g/D \leq 1.0$:

$$\left(1 - \frac{A_g}{A}\right) = 0.21\left(1 - \frac{H_g}{D}\right) + 3.52\left(1 - \frac{H_g}{D}\right)^2 - 4.93\left(1 - \frac{H_g}{D}\right)^3$$

3.7 OPTIMUM PRESSURE FOR GAS–OIL SEPARATORS

3.7.1 Introduction

In order to study the effect of operating pressure in gas–oil separation in general, we will consider first the case of a single-stage separation plus a storage tank. Now, what is the effect of the operating pressure on the recovery of stock tank-oil for the following two extreme cases?

1. *High-pressure operation:* This will diminish the opportunity of light hydrocarbons in the feed to vaporize and separate. However, once this liquid stream is directed to the storage tank (normally operating at/or close to atmospheric pressure), violent flashing occurs due to the high-pressure drop, with subsequent severe losses of the heavier hydrocarbons into the gas phase.
2. *Low-pressure operation:* Here, large quantities of light hydrocarbons will separate from the gas–oil separator, carrying along with them heavier hydrocarbons, causing a loss in the recovered oil. Upon directing this liquid stream to the storage tank, it suffers very little loss in heavy components, because the bulk of light gases were separated in the separator.

From the above, it may be concluded that a proper operating pressure has to be selected and its value has to be between the two extreme cases as described in order to maximize the oil yield. This conclusion is illustrated in Figure 19, in which an optimum pressure of 45 psig is selected to give 0.75 bbl of oil yield.

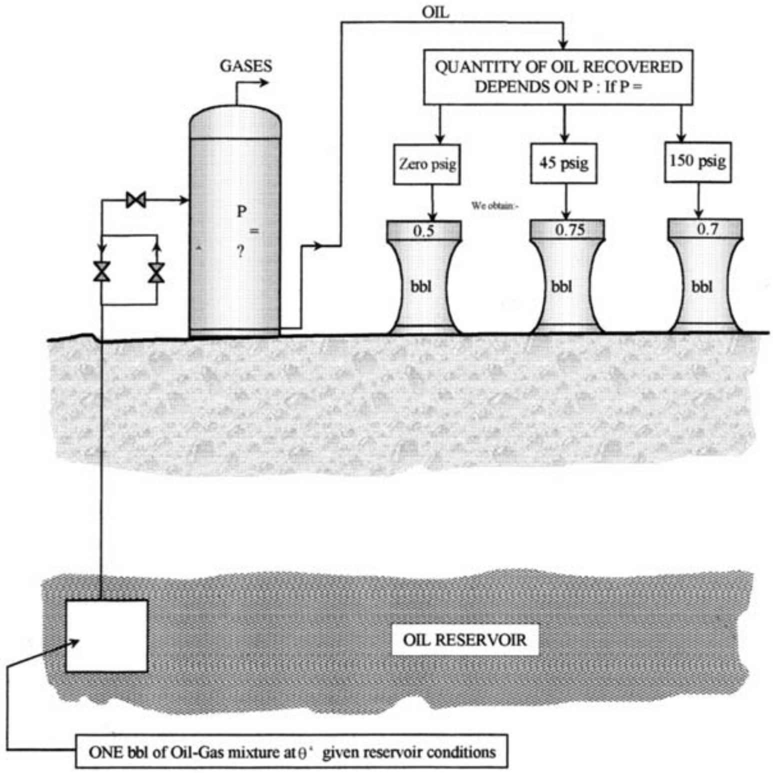


Figure 19 Effect of operating pressure on separation.

The same concept is discussed quantitatively using Rault's and Dalton's laws: For a multicomponent mixture of hydrocarbons, the following thermodynamic relationships exist:

1. In the liquid phase, the partial pressure exerted by each component i (P_i^l) is related to its vapor pressure (P_i^0) and mole fraction (X_i) by

$$P_i^l = P_i^0 X_i \tag{23}$$

2. In the vapor phase, the partial pressure of component $i\,(P_i^v)$ in the gas mixture is related to its mole fraction (Y_i) and total pressure (P_t) by

$$P_i^v = P_t Y_i \tag{24}$$

At given conditions of T and P, thermodynamic equilibrium exists, because the vapor and the liquid are in intimate contact. Consequently, the partial pressure of a component in the vapor is equal to the partial vapor pressure of the component in the liquid. Therefore, equating the two equations and rearranging, the following relationship is obtained:

$$\frac{P_i^0}{P_t} = \frac{Y_i}{X_i} = K_i = \text{Equilibrium constant} \tag{25}$$

where P_i^v is the partial pressure of component i in the mixture, P_i^0 is the vapor pressure of a pure component, i, X_i is the mole fraction of component i in the liquid phase, Y_i is the mole fraction of component i in the gas phase, and P_t is the total pressure on the system (inside the separator).

Now, it can be concluded that if the total pressure on the system is increased, the mole fraction in the gas phase has to decrease in accordance with Eq. (21). In other words, the tendency of vaporization diminishes as the pressure inside the gas–oil separator increases.

In considering the above relationship based on Rault's law, there are some shortcomings in using it, especially for gas mixtures containing methane (its critical temperature is $-116°\text{F}$). However, for mixture of C_3, C_4, and C_5, Rault's law could be applied fairly well at temperatures up to 150°F and pressures up to 100 psia.

3.7.2 Pressure Profile of a Three-Stage GOSP

In the determination of the optimum operating pressure for a GOSP consisting of three stages (high-, intermediate-, and low-pressure separators) it is the second-stage pressure that could be freely changed, hence optimized.

The pressure in the first stage (high pressure) is normally fixed under one of the following conditions:

1. Matching certain requirements to supply high-pressure gas for gas-injection facilities existing in the field
2. Selling the gas through pipelines
3. Flow conditions of the producing wells

Similarly, the pressure in the third stage (low pressure) is fixed for the following cases:

1. The last stage is a storage tank.
2. An existing gas-gathering or vapor recovery facilities that utilizes the gas.
3. The last stage operates at the relatively low pressure.

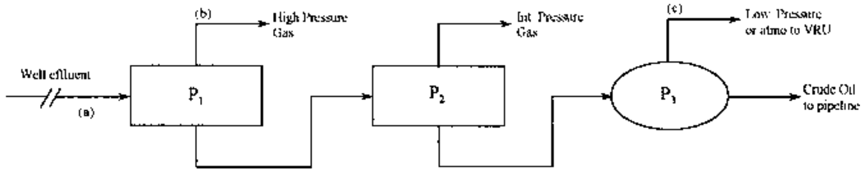


Figure 20 Conditions affecting the determination of $(P_2)_o$. P_1 is fixed in relation to streams at points (a) and (b); P_3 is fixed in relation to stream at point (c), and P_2 is determined by optimization.

Figure 20 illustrates a three-stage GOSP in which the pressure in the first and the third stages are fixed for the above reasons.

The optimum pressure of operating a GOSP is defined as the second-stage pressure that provides the desired gas–oil separation with maximum oil recovery in the stock tank and minimum gas–oil ratio (GOR).

If R designates the recovery of oil and is defined as

$$R = O/G \qquad \text{(bbl of oil/SCF of Gas)} \tag{26}$$

then the optimum pressure in the second-stage separator is the value that makes R maximum or $1/R$ minimum.

Apart from obtaining a high recovery of oil, operating pressures have other important considerations in the processing of the separated streams. A minimum pressure has to be maintained in order for the oil to be delivered to the next processing stage. In addition, using high pressure will deliver the gas stream for sales at higher output pressure, thus reducing the compressor horsepower used for gas pumping.

In general, pressure of 50–100 psia is considered an optimum value of the second-stage operation, whereas a minimum pressure for the third stage will be in the range of 25–50 psig [4].

3.7.3 Determination of the Optimum Second-Stage Operating Pressure

Four methods can be used to determine the optimum operating pressure for the second stage.

Experimental Measurements

In this method, experimental runs are carried out in which the composition of the gas leaving the separator is analyzed and the content of some key component (e.g., C_5^+) is determined. Now, while increasing the pressure in the second stage, we should calculate the ratio of gas to oil (G/O) for both second and third stages. A graphical plot for P_2 versus G/O is given in

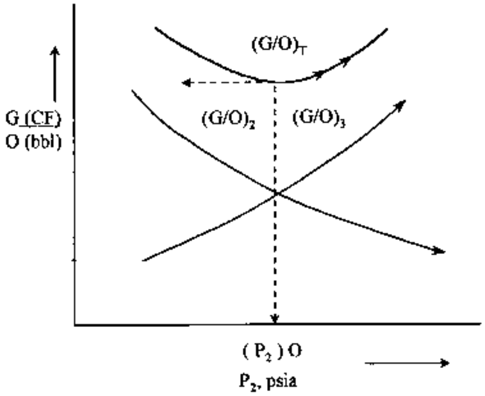


Figure 21 Variation of G/O with P_2.

Figure 21. It is observed that as the pressure increases, $(G/O)_2$ decreases because condensation of C_5^+ is enhanced at higher pressure, whereas $(G/O)_3$ increases because the pressure difference between stages 2 and 3 becomes higher, causing more hydrocarbons to vaporize from the third stage. The optimum operating pressure is the value that makes $(G/O)_T$ minimum.

Further verification of the value of $(P_2)_o$ is done as follows:

1. Determine experimentally the C_5^+ content in the gas samples leaving the top of separators 2 and 3, in gallons/MCF of the gas (gal C_5^+/MCF).
2. Calculate the [gal C_5^+/bbl oil] by multiplying results of step 1 by G/O, the gas/oil ratio [MCF/bbl oil]. Sum the values for both stages 1 and 2 to give the total gal C_5^+ lost per barrel of oil.
3. Repeat for different values of operating pressure, P_2.
4. Plot the value of gal C_5^+/MCF for both stages 2 and 3 versus P_2, as shown in Figure 22. It is to be noted that the C_5^+ content per MCF in the gas streams decreases with the increase in P_2 for both stages. However, it is higher for stage 3 than stage 2.
5. Plot the total gal C_5^+ lost per barrel of oil versus P_2, as given in Figure 23. An optimum value, $(P_2)_o$ is determined, which corresponds to a minimum loss of C_5^+ in gallons per barrel of oil.
6. This is a double check on the value obtained from Figure 21 [3].

Empirical Equations

Two empirical equations have been developed [7] to calculate $(P_2)_o$ for the second stage in a three-stage separation system, which are functions of

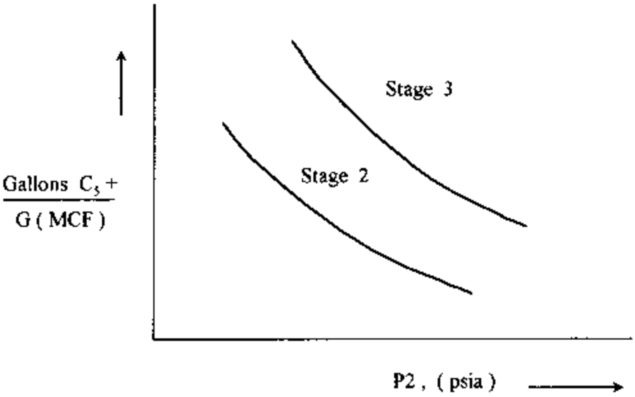


Figure 22 Variation of gasoline contents with P_2.

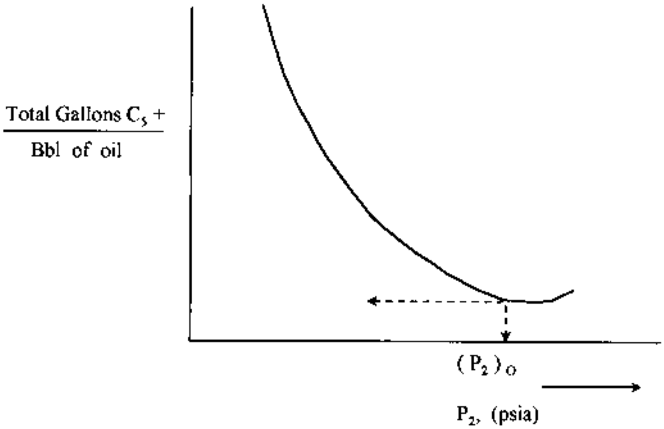


Figure 23 Determination of $(P_2)_o$ experimentally.

the initial-stage pressure and the system composition. Equation (27a) is used when the molecular weight, M, of the well stream relative to air is greater than 1.0:

$$(P_2)_o = AP_1^{0.686} + 0.429 \times 10^2(A - 0.057) \text{ psia} \tag{27a}$$

Equation (27b) is used when M is less than 1.0:

$$(P_2)_o = AP_1^{0.765} + 0.833 \times 10^2(A + 0.028) \text{ psia} \tag{27b}$$

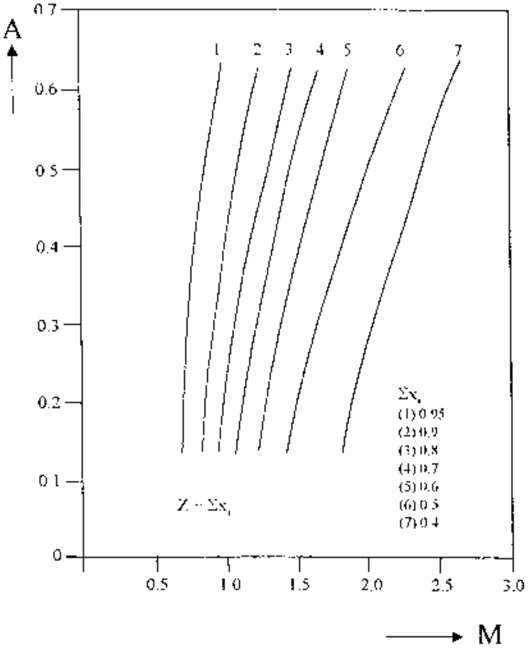


Figure 24 Determination of the *A* factor.

A is a dimensionless constant evaluated from Figure 24 using *M* and a parameter, *Z* (the sum of the mole fractions of the hydrocarbons $C_1 + C_2 + C_3$) found in the feed.

Approximate Formula

The following simplified formula can be used for determining $(P_2)_o$ as a function of P_1 and P_3 [8]:

$$(P_2)_o = (P_1 P_3)^{0.5} \tag{28}$$

Example 3.5

A well fluid with the composition (in mole fraction) $C_1 = 0.4$, $C_2 = 0.2$, $C_3 = 0.1$, $C_4 = 0.1$, $C_5 = 0.1$, $C_6 = 0.05$, and $C_7 = 0.05$ is to undergo a gas–oil separation process using a three-stage GOSP. The last stage is operated at atmospheric pressure. The first stage is fixed at a pressure of 500 psia. Calculate the optimum operating pressure for the second stage using

(i) Equation (27a) or (27b)
(ii) Equation (28)

Solution

(i) The average molecular weight of the well stream, M_{av}, is 39.92. M is determined next:

$$\begin{aligned} M &= \text{Molecular weight of air } M_{av} \\ &= 28.96/39.92 \end{aligned}$$

Because M is greater than 1.0, Eq. (27a) is applied.
To find the value of A, go to Figure 24. Using $Z = 0.7$ and $M = 1.38$, A is found to be $= 0.421$. Substituting the values of A and P_1 in Eq. (27a), the most favorable operating pressure $(P_2)_o$ is calculated and found to be 50 psia.

(ii) Direct substitution of the values of $P_1 = 500$ psia and $P_3 = 14.7$ psia in Eq. (28) yields $(P_2)_o = 85.5$ psia

Equilibrium Flash Vaporization Calculation

Equations describing flash vaporization, as given in Section 3.8, can be applied to calculate the value of optimum operating pressure for the second-stage separator for a three-stage GOSP. An outline of a trial-and-error calculation procedure is presented in Figure 25.

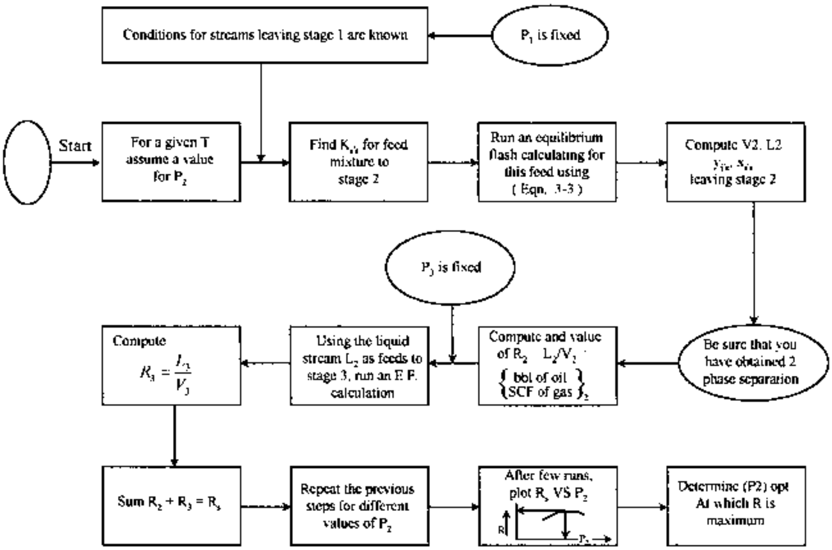


Figure 25 An outline of a proposed procedure for computing $(P_2)_o$ for a three-stage GOSP where P_1 and P_3 are fixed.

3.8 SELECTIONS AND PERFORMANCE OF GAS–OIL SEPARATORS

For the selection of a particular separator, preliminary survey of all conditions and factors prevailing on site in an oil field should be considered. The main factors underlying this survey as well as the basic steps included in the selection of a separator are embodied in one integrated scheme outlined in Figure 26.

The performance of gas–oil separators, on the other hand, are controlled by four factors:

1. *Operating temperature:* A higher temperature will cause more evaporation of the hydrocarbons, diminishing the recovery of the liquid portion. In general, the trend of change in the relative volume of the liquid recovered from a GOSP with temperature is illustrated in Figure 27.
2. *Operating pressure:* A higher pressure will allow more hydrocarbons to condense, increasing liquid recovery. However, after reaching a certain peak, a higher pressure causes liquid to decrease [4]. This is in accordance with the retrograde phenomena explained by Abdel-Aal [3]. In Figure 28, the lower curve V_s represents this case.
3. *Number of stages:* Increasing the number of stages in general will increase the efficiency of separation, resulting in a higher yield of the stable stock tank oil. This is true for a number of stages in the range of two to three. However, if the number increases beyond three, improvement in the recovery diminishes. A four-stage GOSP is not economically attractive, as the recovery of stable stock tank oil increases by 8% by adding one more stage [8].
4. In addition to the above factors, the composition of the well streams has to be considered in evaluating the performance of a gas–oil separator.

In order to judge the performance of a GOSP, a number of tests are commercially carried out to evaluate the efficiency of operation. Most important are the following:

1. *Evaluation of particle size:* The method requires determining the size of liquid particles entrained by the gas stream. The efficiency of gas–oil separator is thus evaluated based on this size. It is accepted that liquid particles with size larger than 10 μm coming in the gas stream is an indication of poor performance.

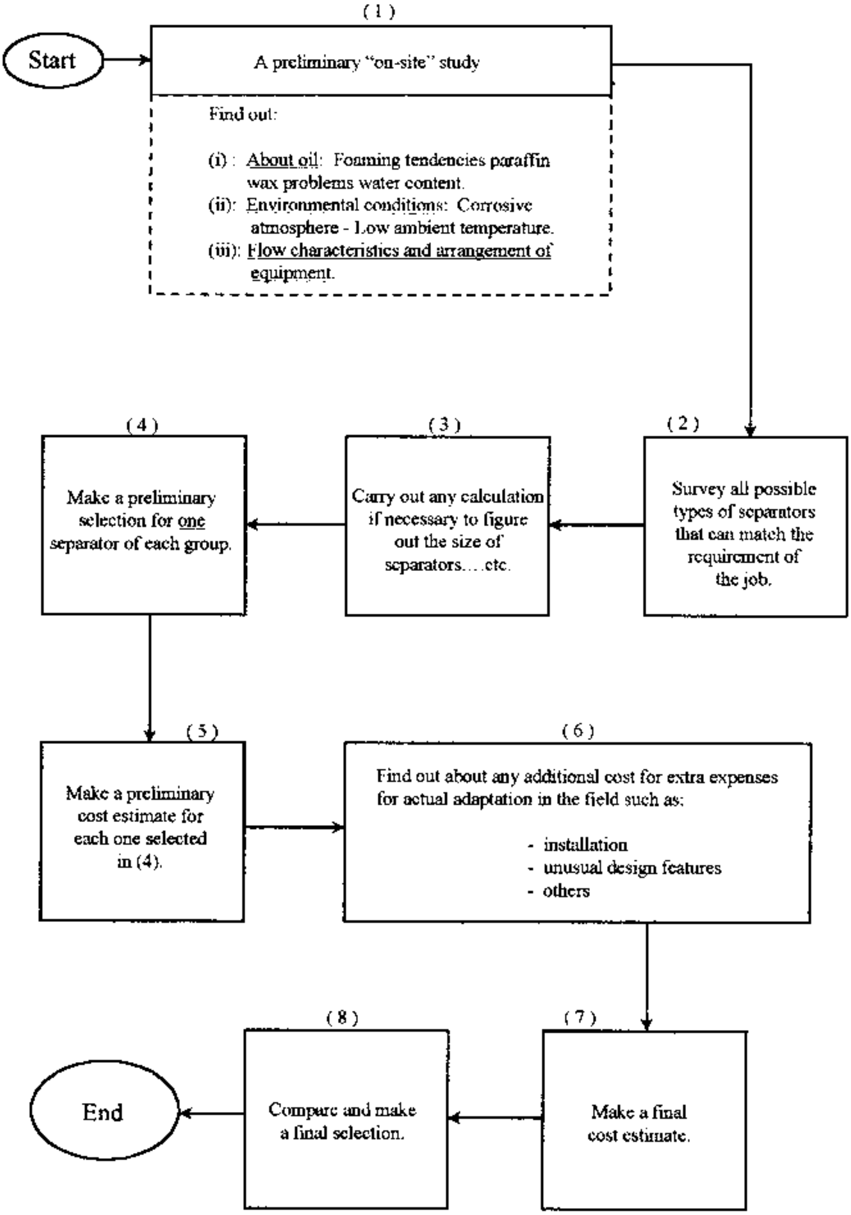


Figure 26 How to select a separator for a particular job.

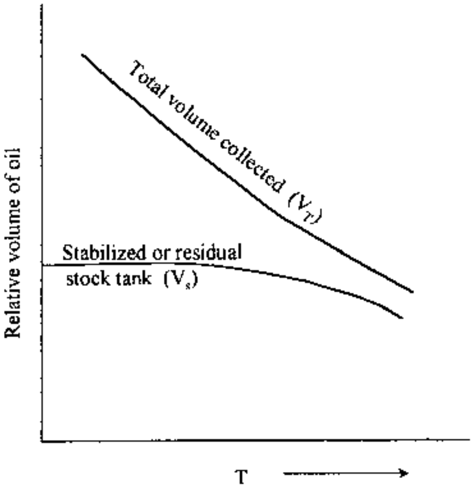


Figure 27 Change of relative volume of oil recovered from GOS with operating *T*.

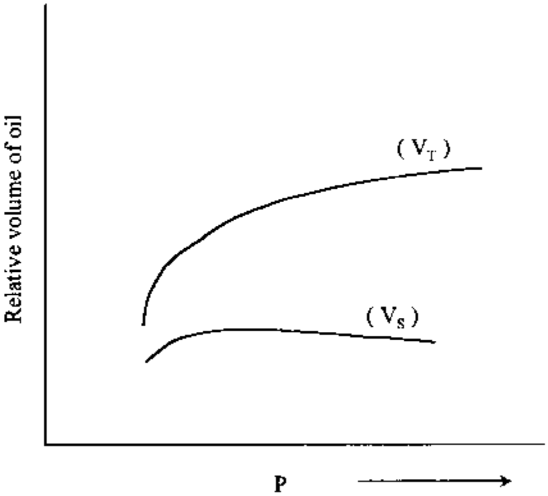


Figure 28 Change of relative volume of oil recovered from a GOS with operating pressure.

2. *Determination of the quantity of liquid carryover:* The method requires determining the volume of liquid entrained or "carried over" by the gas stream. A limit of 0.10 gal/MCF in the exit gas is usually specified (where $M = 1000$)

3. *Stain test:* It is rather an old test (formerly named the "handkerchief" test). It simply consists of holding and exposing a white cloth in the gas stream leaving the separator. The performance of the separator is considered acceptable if a brown strain does not form on the cloth in 1 min.

3.9 FLASH CALCULATIONS

3.9.1 Introduction

The problem of separating the gas from crude oil for well fluids (crude oil mixtures) breaks down to the well-known problem of flashing a partially vaporized feed mixture into two streams: vapor and liquid. In the first case, we use a gas–oil separator; in the second case, we use what we call a flashing column, as shown in Figure 29.

A "flash" is a single-stage distillation in which a feed is partially vaporized to give a vapor that is richer in the more volatile components. This is the case of a feed heated under pressure and flashed adiabatically across a valve to a lower pressure, the vapor being separated from the liquid residue in a flash drum. This is the case of "light liquids." Apart from the gas–oil separation problem addressed here, methods used in practice to produce and hence separate two-phase mixtures are as follows

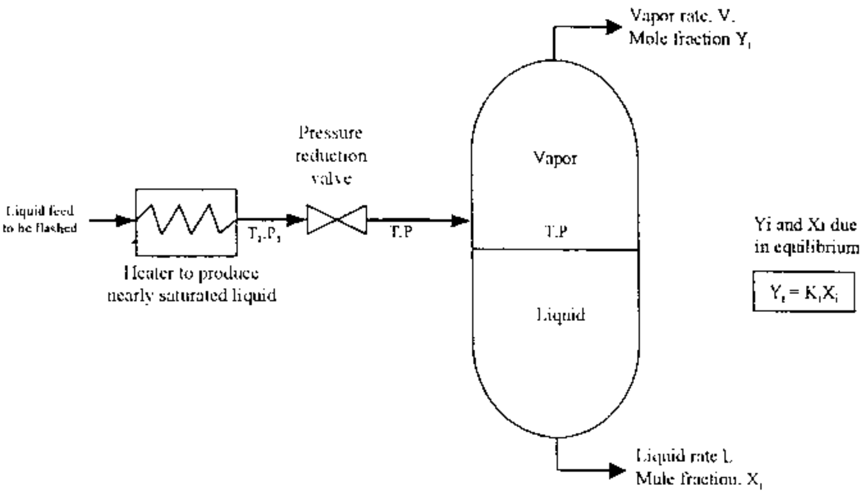


Figure 29 Schematic presentation of equilibrium flash vaporization (adiabatic flash with value; for isothermal flash, no valve is used).

Initial phase	Process and conditions
1. Higher-pressure liquids (light)	Heat under pressure, then flash adiabatically using valve
2. Low-pressure liquids (heavy)	Partial vaporization by heating, flash isothermally (no valve)
3. Gas	Cool-after initial compression
4. Gas	Expand through a valve or engine

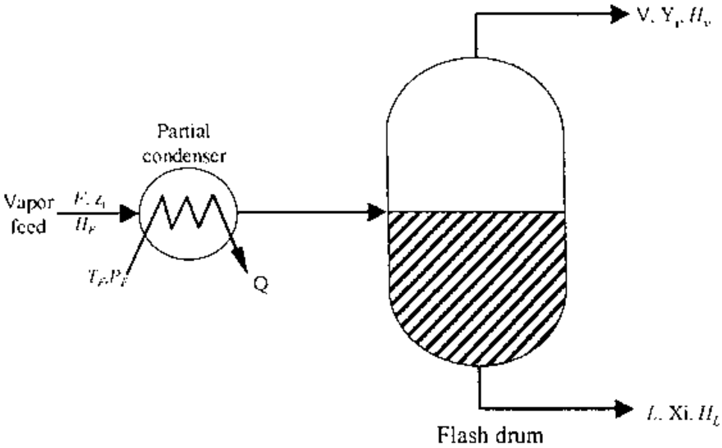


Figure 30 Continuous single-stage equilibrium separation by partial condensation of a gas feed (case similar to isothermal flash for a required T_r).

The diagram shown in Figure 30 illustrates case 3, which is a continuous single-stage separation of a gas feed by partial condensation. The vapor and liquid leaving the flashing drum are in intimate contact, hence, they are in equilibrium.

3.9.2 Conditions Necessary for Flashing

For flashing to take place, the feed has to be two-phase mixture, that is, it satisfies the following:

$T_{\mathrm{BP}} < T_f < T_{\mathrm{DP}}$ (as indicated in Fig. 31)

or

The sum of $[z_i K_i]$ for all components is greater than 1 and
the sum of $[z_i / K_i]$ for all components is less than 1

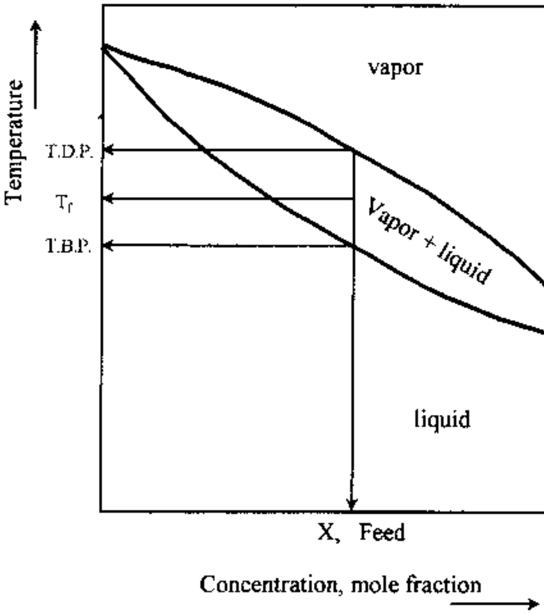


Figure 31 Demonstration of the conditions for flashing for a binary system.

where T_{BP}, T_f, and T_{DP} are the bubble point of the feed mixture, flash temperature, and dew point of the feed mixture, respectively; Z_i and K_i are feed. The composition and equilibrium constant, respectively, for component i.

3.9.3 The Flash Equation

The need to discuss flash calculation arises from the fact that it provides a tool to determine the relative amounts of the separation products V (gas) and L (oil) and their composition Y_i and X_i, respectively.

The flash equation is derived by material balance calculations as presented in Figure 32. Two forms are presented:

Simple form:

$$X_i = \frac{Z_i}{1 - \frac{V}{F}(1 - K_i)} \tag{29}$$

For given conditions of P and T, the solution of the equation to find the value of X_i is obtained by trial and error assuming a value for V/F (take F unity), until the sum of $[X_i] = 1$ is satisfied.

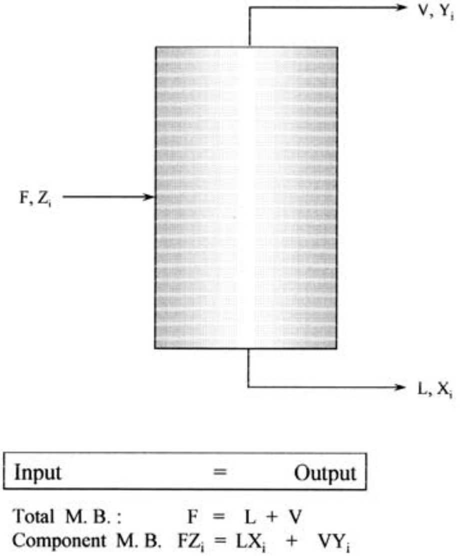


Figure 32 Material balance for flash vessel.

Functional form:

$$f(g) = \sum_{i=1}^{i=c} \left[\frac{Z_i}{1 - g(1 - K_i)} \right] - 1 \tag{30}$$

where $g = V/F$. The desired root to this function, $gr = (V/F)r$, is the value that makes the function $f(g)$ goes to zero or the sum of $[X_i] = 1$. The solution of this flash equation is carried out using computers. Details on using functional technique are fully described in Ref. 9.

The vapor–liquid equilibrium constant, defined as

$$K_i = Y_i/X_i$$

where Y_i is the mole fraction of component i in the vapor phase, X_i is the mole fraction of component i in the liquid phase, and K_i is the equilibrium constant of component i for a given T and P, is considered the key concept used in the computation of phase behavior of hydrocarbon mixtures in oil and gas streams. K is called the *distribution constant* because it predicts the distribution of a component in each phase: vapor and liquid.

K is a function of T, P, and the composition of a given system. The K values most widely used are those developed by the National Gas Processors Association (NGPA) and are presented in Appendix A. Because the K values are designated for a specific *convergence pressure* (defined as the pressure at a given T, where the values of K for all component in a system become or tend to become equal to unity), it is important that the value of the operating pressure of the system should be below the convergence pressure used in predicting the K values.

3.9.4 Some Important Applications for the Flash Equation

In addition to the above, other applications for the flash equation are as follows:

1. *To find the number of stages needed for a given separation:* This simply utilizes the stage-to-stage concept in which another stage is added to each trial. Convergence is established when the liquid composition leaving the last stage equals the specified crude oil.
2. *To determine the optimum pressure for the second stage:* This involves a trial and error procedure [3].

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REVIEW QUESTIONS AND EXERCISE PROBLEMS

1. What is the purpose of each of the following internal components of a separator?
   (a) Inlet diverter
   (b) Mist extractor
   (c) Vortex breaker
   (d) Wave breaker
2. What are the forces acting on a liquid droplet in the gravity settling section of a separator?
3. What is the significance of retention time in separator design?
4. Complete the following statements:
   (a) In the gravity settling of a gas–oil separator, oil droplets down to a size of ______________ should be removed in order to prevent ______________
   (b) In a gas scrubber, removal of 500-μm droplets in the gravity settling section is sufficient because _______________________
5. Which type of separator (vertical or horizontal) is more suitable for the following cases (**justify your selection**):
   (a) Handling solids in the fluid
   (b) Handling large volumes of gas
   (c) Handling liquid surges
   (d) Offshore operations
6. State whether each of the following statements is true (**T**) or false (**F**):
   (a) Carryover is the escape of free liquid with the gas due to vortexes.
   (b) In, the separator design, unless laboratory data are available, the smallest water droplet size that should be removed from oil is taken as 100 μm.
   (c) Blowby is the escape of free gas with the liquid that may result from excessive foaming.
   (d) In long horizontal separators, waves may develop at the interface. This results in level control problems, which could be eliminated by the installation of coalescing plates.
7. Circle the correct answers (note that more than one answer could be correct):
   I. Foamy crude oil may result in
      (a) Improper gas–oil interface level control

  - (b) Carryover
  - (c) Better separation in the gravity settling section
  - (d) Entrapment of foam in the oil

II. Accumulation of produced sand in the separator may

  - (a) Result in incomplete separation of oil and gas
  - (b) Be better handled in horizontal separators
  - (c) Reduce the gas capacity of the separator
  - (d) Reduce the liquid (oil) capacity of the separator

8. Derive the equations for gas capacity and oil capacity constraints for a horizontal separator for the following two cases:

   (a) The gas occupies 75% of the separator effective volume.
   (b) The gas occupies 25% of the separator effective volume.

9. The following data are given for a small oil field:

   Oil rate: 8000 bbl/day
   Gas–oil ratio: 1200 SCF/bbl
   Operating pressure: 500 psia
   Operating temperature: 80°F
   Oil gravity: 35°API
   Gas specific gravity: 0.6
   Oil viscosity: 10 cP
   Oil retention time: 180 s

   (a) Design a horizontal separator to handle the field production.
   (b) If the field is located offshore, design the appropriate separator to handle to field production.
   (c) Determine the actual capacity of a horizontal two-phase separator having a 60 in. diameter and 15 ft seam-to-seam length for the above operating conditions.

10. Determine the actual gas and oil capacities for vertical and horizontal separators having a 48-in. diameter and 16-ft seam-to-seam length, operating at 585 psi and 80°F, given the following data:

    Oil specific gravity: 0.875
    Gas specific gravity: 0.6
    Oil viscosity: 5.0 cP
    Operating pressure: 700 psia
    Operating temperature: 80°F
    Oil retention time: 120 s

C_d: 0.86
Z: 0.89

11. Design a vertical and horizontal *gas scrubbers* for the following conditions:

    Gas rate: 80 MMSCFD
    Gas specific gravity: 0.6
    Operating pressure: 1200 psia
    Operating temperature: 80°F
    Oil rate: Negligible
    Oil gravity: 40°API

4

Three-Phase Oil–Water–Gas Separation

4.1 INTRODUCTION

The concepts, theory, and sizing equations for two-phase gas–liquid separators have been discussed in Chapter 3. The material presented in Chapter 3 applies, in general, to the separation of any gas–liquid system such as gas–oil, gas–water, and gas–condensate systems. In almost all production operations, however, the produced fluid stream consists of three phases: oil, water, and gas.

Generally, water produced with the oil exists partly as free water and partly as water-in-oil emulsion. In some cases, however, when the water–oil ratio is very high, oil-in-water rather than water-in-oil emulsion will form. Free water produced with the oil is defined as the water that will settle and separate from the oil by gravity. To separate the emulsified water, however, heat treatment, chemical treatment, electrostatic treatment, or a combination of these treatments would be necessary in addition to gravity settling. This is discussed in Chapter 5. Therefore, it is advantageous to first separate the free water from the oil to minimize the treatment costs of the emulsion.

Along with the water and oil, gas will always be present and, therefore, must be separated from the liquid. The volume of gas depends largely on the producing and separation conditions. When the volume of gas is relatively small compared to the volume of liquid, the method used to separate free water, oil and gas is called a *free-water knockout*. In such a case, the separation of the water from oil will govern the design of the vessel. When there is a large volume of gas to be separated from the liquid (oil and water), the vessel is called a *three-phase separator* and either the gas capacity requirements or the water–oil separation constraints may govern the vessel design. Free-water knockout and three-phase separators are basically similar in shape and components. Further, the same design concepts and procedures are used for both types of vessel. Therefore, the term *three-phase separator* will be used for both types of vessel throughout the chapter.

Three-phase separators may be either horizontal or vertical pressure vessels similar to the two-phase separators described in Chapter 3. However, three-phase separators will have additional control devices and may have additional internal components. In the following sections, the two types of separator (horizontal and vertical) are described and the basic design equations are developed.

4.2 HORIZONTAL THREE-PHASE SEPARATORS

Three-phase separators differ from two-phase separators in that the liquid collection section of the three-phase separator handles two immiscible liquids (oil and water) rather than one. This section should, therefore, be designed to separate the two liquids, provide means for controlling the level of each liquid, and provide separate outlets for each liquid. Figures 1 and 2 show schematics of two common types of horizontal three-phase separators. The difference between the two types is mainly in the method of controlling the levels of the oil and water phases. In the first type (Fig. 1), an interface controller and a weir provide the control. The

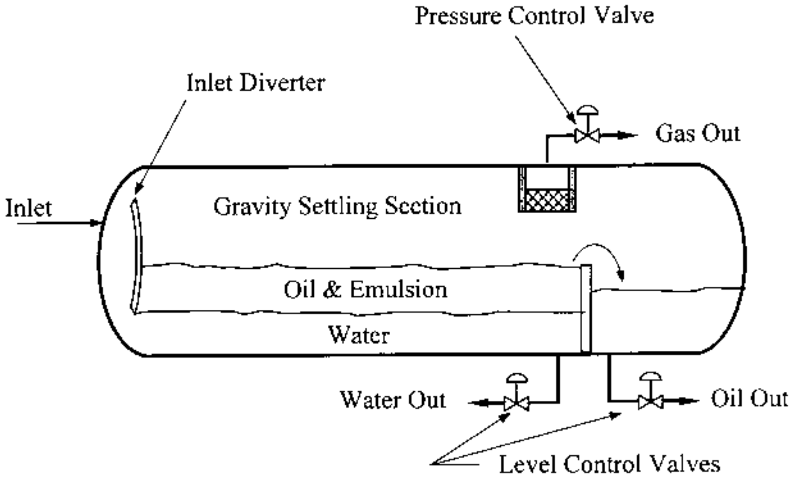

Figure 1 Horizontal three-phase separator schematic of one type.

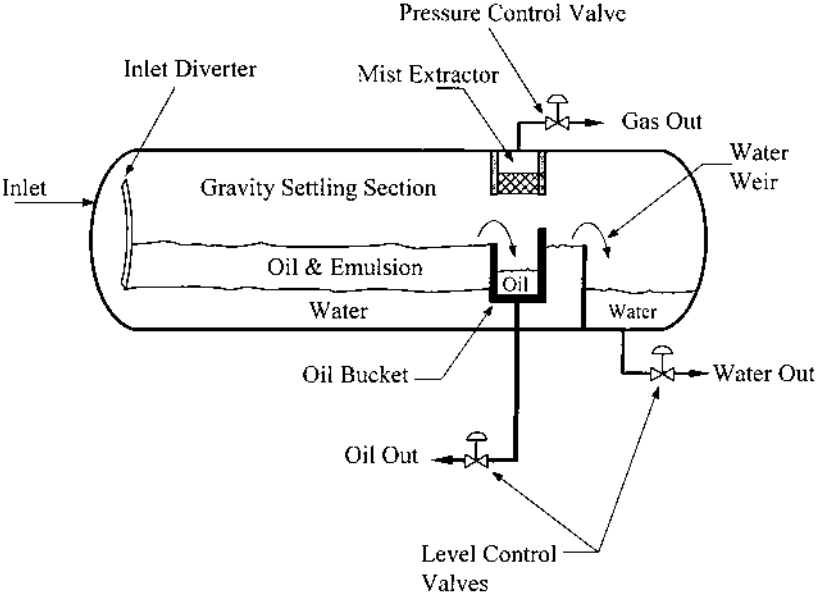

Figure 2 Horizontal three-phase separator; bucket and weir design.

design of the second type (Fig. 2), normally known as the *bucket and weir design*, eliminates the need for an interface controller.

The operation of the separator is, in general, similar to that of the two-phase separator. The produced fluid stream, coming either directly from the producing wells or from a free-water knockout vessel, enters the separator and hits the inlet diverter, where the initial bulk separation of the gas and liquid takes place due to the change in momentum and difference in fluid densities. The gas flows horizontally through the gravity settling section (the top part of the separator) where the entrained liquid droplets, down to a certain minimum size (normally 100 μm), are separated by gravity. The gas then flows through the mist extractor, where smaller entrained liquid droplets are separated, and out of the separator through the pressure control valve, which controls the operating pressure of the separator and maintains it at a constant value. The bulk of liquid, separated at the inlet diverter, flows downward, normally through a downcomer that directs the flow below the oil–water interface. The flow of the liquid through the water layer, called *water washing*, helps in the coalescence and separation of the water droplets suspended in the continuous oil phase. The liquid collection section should have sufficient volume to allow enough time for the separation of the oil and emulsion from the water. The oil and emulsion layer forming on top of the water is called the *oil pad*. The weir controls the level of the oil pad and an interface controller controls the level of the water and operates the water outlet valve. The oil and emulsion flow over the weir and collect in a separate compartment, where its level is controlled by a level controller that operates the oil outlet valve.

The relative volumes occupied by the gas and liquid within the separator depend on the relative volumes of gas and liquid produced. It is a common practice, however, to assume that each of the two phases occupies 50% of the separator volume. In such cases, however, where the produced volume of one phase is much smaller or much larger than the other phase, the volume of the separator should be split accordingly between the phases. For example, if the gas–liquid ratio is relatively low, we may design the separator such that the liquid occupies 75% of the separator volume and the gas occupies the remaining 25% of the volume.

The operation of the other type of horizontal separator (Fig. 2) differs only in the method of controlling the levels of the fluids. The oil and emulsion flow over the oil weir into the oil bucket, where its level is controlled by a simple level controller that operates the oil outlet valve. The water flows through the space below the oil bucket, then over the water weir into the water collection section, where its level is controlled by a level controller that operates the water outlet valve. The level of the liquid in the separator, normally at the center, is controlled by the height of the oil weir. The thickness of the oil pad must be sufficient to provide adequate oil retention time. This is controlled by the height of the water weir relative to that of the oil weir. A simple pressure balance at the bottom of the separator between the water side and the water and oil side can be used to approximately determine the thickness of the oil pad as follows:

$$H_o = \frac{H_{\text{ow}} - H_{\text{ww}}}{1 - (\rho_o/\rho_w)} \tag{1}$$

where H_o is the thickness of the oil pad, H_{ow} is the height of the oil weir, H_{ww} is the height of the water weir, and ρ_o and ρ_w are the oil and water densities, respectively.

Equation (1) gives only an approximate value for the thickness of the oil pad. A more accurate value could be obtained if the density of the oil in Eq. (1) is replaced by the average value of the density of oil and density of emulsion, which depends on the thickness of the oil and emulsion layers within the oil pad. The height of the water weir should not be so small as to avoid the downward growth of the oil pad and the possibility of the oil flowing below the oil bucket, over the water weir, and out with the water.

It is advisable to have the oil bucket as deep as possible and to have either the oil weir, or the water weir, or both to be adjustable to accommodate any unexpected changes in flow rates and/or liquids properties. Such problems are easily accommodated in the interface controller and weir design of Fig. 1, as the interface controller could be easily adjusted. In some cases, however, when the difference in density between the water and oil, or the water and emulsion are small (e.g., in heavy oil operations), the operation of the interface controller becomes unreliable and the bucket design (Fig. 2) will be preferred.

4.3 VERTICAL THREE-PHASE SEPARATORS

As discussed in Chapter 3, the horizontal separators are normally preferred over vertical separators due to the flow geometry that promotes phase separation. However, in certain applications, the engineer may be forced to select a vertical separator instead of a horizontal separator despite the process-related advantages of the later. An example of such applications is found in offshore operations, where the space limitations on the production platform may necessitate the use of a vertical separator.

Figure 3 shows a schematic of a typical three-phase vertical separator. The produced fluid stream enters the separator from the side

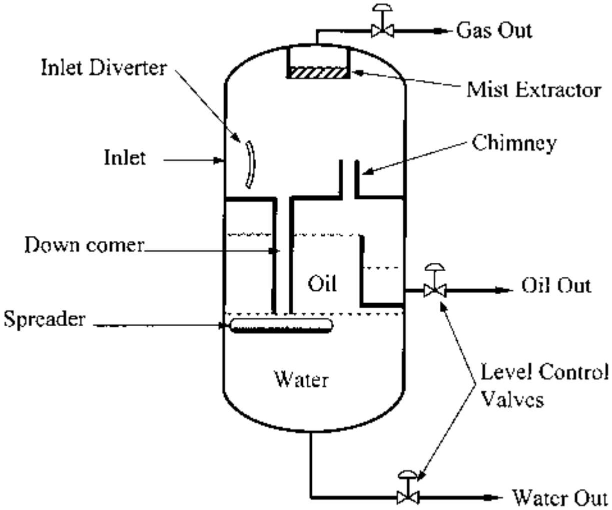

Figure 3 Schematic of a three-phase vertical separator.

and hits the inlet diverter, where the bulk separation of the gas from the liquid takes place. The gas flows upward through the gravity settling sections which are designed to allow separation of liquid droplets down to a certain minimum size (normally 100 μm) from the gas. The gas then flows through the mist extractor, where the smaller liquid droplets are removed. The gas leaves the separator at the top through a pressure control valve that controls the separator pressure and maintains it at a constant value.

The liquid flows downward through a downcomer and a flow spreader that is located at the oil–water interface. As the liquid comes out of the spreader, the oil rises to the oil pad and the water droplets entrapped in the oil settle down and flow, countercurrent to the rising oil phase, to collect in the water collection section at the bottom of the separator. The oil flows over a weir into an oil chamber and out of the separator through the oil outlet valve. A level controller controls the oil level in the chamber and operates the oil outlet valve. Similarly, the water out of the spreader flows downward into the water collection section, whereas the oil droplets entrapped in the water rise, countercurrent to the water flow, into the oil pad. An interface controller that operates the water outlet valve controls the water level. In the design shown in Figure 3, a chimney must be provided, as shown in the figure, to allow the gas liberated from the oil to rise and join the rest of the separated gas and, thus, avoid overpressurizing the liquid section of the separator. The use of the oil weir and chamber in this design provides good separation of water from oil, as the oil has to rise to the full height of the weir before leaving the separator. The oil chamber, however, presents some problems. First, it takes up space and reduces the separator volume needed for the retention times of oil and water. It also provides a place for sediments and solids to collect, which creates cleaning problems and may hinder the flow of oil out of the vessel. In addition, it adds to the cost of the separator.

Other methods of level control are also available. Figure 4 shows a schematic of a separator where an oil–water-interface controller and a gas–oil-interface controller control the water and oil levels, respectively. Figure 5 shows yet another method of level control. In this design, an external water column equipped with adjustable weir is connected to the water section of the separator. The column is also piped to the gas section of the separator to establish pressure equilibrium between the water column and separator. A simple level controller controls the height of the water in the column, which, in turn, controls the height of water in the separator. This eliminates the need for an oil–water interface controller and avoids the potential problems associated with such controllers. This design, however, takes additional space and adds additional significant cost.

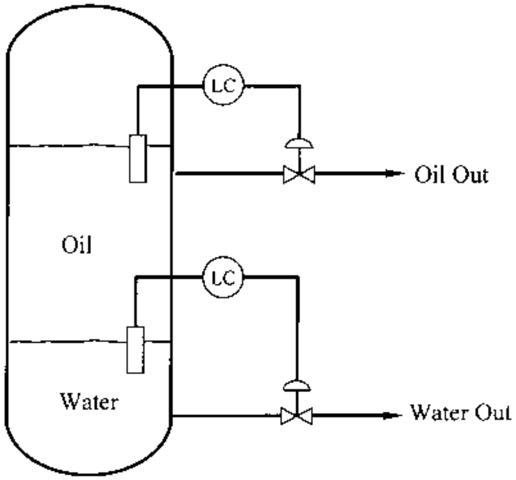

Figure 4 Interface level control.

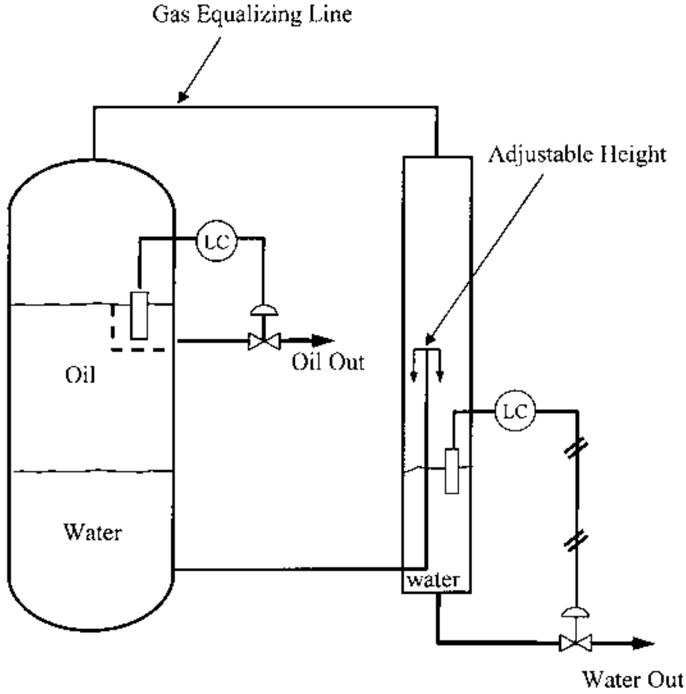

Figure 5 Water leg with or without oil chamber.

Liquid–liquid interface controllers will function effectively as long as there is an appreciable difference between the densities of the two liquids. In most three-phase separator applications, water–oil emulsion forms and a water–emulsion interface will be present in the separator instead of a water–oil interface. The density of the emulsion is higher than that of the oil and may be too close to that of the water. Therefore, the smaller density difference at the water–emulsion interface will adversely affect the operation of the interface controller. The presence of emulsion in the separator takes up space that otherwise would be available for the oil and/or the water. This reduces the retention time of the oil and/or water and, thus results in a less efficient oil–water separation. In most operations where the presence of emulsion is problematic, chemicals known as deemulsifying agents are injected into the fluid stream to mix with the liquid phase. These chemicals help in breaking the emulsion, as will be described in Chapter 5. Another method that is also used for the same purpose is the addition of heat to the liquid within the separator. In both cases, however, the economics of the operations have to be weighted against the technical constraints.

4.4 SEPARATION THEORY

The basic separation concepts and settling equations developed for two-phase separators in Chapter 3 are, in general, valid for three-phase separators. In particular, the equations developed for separation of liquid droplets from the gas phase, which determined the gas capacity constraint, are exactly the same for three-phase separators.

Treatment of the liquid phase for three-phase separators is, however, different from that used for two-phase separators. The liquid retention time constraint was the only criterion used for determining the liquid capacity of two-phase separators. For three-phase separators, however, the settling and separation of the oil droplets from water and of the water droplets from oil must be considered in addition to the retention time constraint. Further, the retention time for both water and oil, which might be different, must also be considered.

In separating oil droplets from water, or water droplets from oil, a relative motion exists between the droplet and the surrounding continuous phase. An oil droplet, being smaller in density than the water, tends to move vertically upward under the gravitational or buoyant force, F_g. The continuous phase (water), on the other hand, exerts a drag force, F_d, on the oil droplet in the opposite direction. The oil droplet will accelerate until the fractional resistance of the fluid drag force, F_d, approaches and balances F_g; thereafter, the oil droplet continues to rise at a constant velocity known as the *settling* or *terminal velocity*. Similarly, a water droplet, being higher in density than the oil, tends to move vertically downward under the gravitational or buoyant force, F_g. The continuous phase (oil), on the other hand, exerts a drag force, F_d, on the water droplet in the opposite direction. The water droplet will accelerate until the frictional resistance of the fluid drag force, F_d, approaches and balances F_g; thereafter, the water droplet continues to rise at a constant velocity known as the *settling* or *terminal velocity*. Upward settling of oil droplets in water and downward settling of water droplets in oil follow Stokes' law and the terminal settling velocity can be obtained as follows. The drag force, F_d, is proportional to the droplet surface area perpendicular to the direction of flow, and its kinetic energy per unit volume; Hence,

$$F_d = C_d \frac{\pi}{4} d^2 \frac{\rho_c u^2}{2g} \tag{2}$$

whereas F_g is given by

$$F_g = \frac{\pi}{6} d^3 (\Delta\rho) \tag{3}$$

where d is the diameter of the droplet (ft), u is the settling velocity of the droplet (ft/s), ρ_c is the density of the continuous phase (lb/ft^3), g is gravitational acceleration (ft/s), and C_d is the drag coefficient. For a low Reynolds number, Re, flow, the drag coefficient is given by

$$C_d = \frac{24}{\mathrm{Re}} = \frac{24\mu' g}{\rho d u} \tag{4}$$

where μ' is the viscosity of the continuous phase (lb-s/ft^2).
Substituting for C_d from Eq. (4) into Eq. (2) yields

$$F_d = 3\pi\mu' d u \tag{5}$$

The settling terminal velocity, u, is reached when $F_d = F_g$. Therefore, equating Eqs. (3) and (5) and solving for u, the droplet settling velocity, we obtain

$$u = \frac{(\Delta\rho) d^2}{18\mu'}$$

The typical units used for droplet diameter and viscosity are the micrometers and centipoise, respectively. Letting μ be the viscosity in centipoise and d_m be the droplet diameter in micrometers, the above equation becomes

$$u = \frac{(\Delta\rho)(3.281 \times 10^{-6} d_m)^2}{18(2.088 \times 10^{-5}\mu)}$$

$$u = 2.864 \times 10^{-8} \frac{(\Delta\rho){d_m}^2}{\mu} \qquad \frac{ft}{s} \tag{6}$$

or

$$u = 1.787 \times 10^{-6} \frac{(\Delta\gamma){d_m}^2}{\mu} \qquad \frac{ft}{s} \tag{7}$$

where

$$\Delta\gamma = \gamma_w - \gamma_o$$

γ_o and γ_w are the specific gravity of oil and water, respectively, u is the terminal settling velocity of the droplet (ft/s), d_m is the diameter of the droplet (μm), and μ is the viscosity of the continuous phase (cP).

Equation 6 [or Eq. (7)] shows that the droplet settling velocity is inversely proportional to the viscosity of the continuous phase. Oil viscosity is several magnitudes higher than the water viscosity. Therefore, the settling velocity of water droplets in oil is much smaller than the settling velocity of oil droplets in water. The time needed for a droplet to settle out of one continuous phase and reach the interface between the two phases depends on the settling velocity and the distance traveled by the droplet. In operations where the thickness of the oil pad is larger than the thickness of the water layer, water droplets would travel a longer distance to reach the water–oil interface than that traveled by the oil droplets. This, combined with the much slower settling velocity of the water droplets, makes the time needed for separation of water from oil longer than the time needed for separation of oil from water. Even in operations with a very high water–oil ratio, which might result in having a water layer that is thicker than the oil pad, the ratio of the thickness of the water layer to that of the oil pad would not offset the effect of viscosity. Therefore, the separation of water droplets from the continuous oil phase would always be taken as the design criterion for three-phase separators.

The minimum size of the water droplet that must be removed from the oil and the minimum size of the oil droplet that must be removed from the water to achieve a certain oil and water quality at the separator exit depend largely on the operating conditions and fluid properties. Results obtained from laboratory tests conducted under simulated field conditions provide the best data for design. The next best source of data could be obtained from nearby fields. If such data are not available, the minimum water droplet size to be removed from the oil is taken as 500 μm. Separators design with this criterion have produced oil and emulsion containing between 5% and 10% water. Such produced oil and emulsion could be treated easily in the oil dehydration facility, as discussed in Chapter 5. Experience has also shown that three-phase separators designed based on the 500-μm water droplet removal produces water with a suspended oil content that is below 2000 mg/L. This produced water must be treated before it is disposed of, as described in Chapter 7.

Another important aspect of separator design is the retention time, which determines the required liquid volumes within the separator. The oil phase needs to be retained within the separator for a period of time that is sufficient for the oil to reach equilibrium and liberates the dissolved gas. The retention time should also be sufficient for appreciable coalescence of the water droplets suspended in the oil to promote effective settling and separation. Similarly, the water phase needs to be retained within the separator for a period of time that is sufficient for coalescence of the suspended oil droplets. The retention times for oil and water are best determined from laboratory tests; they usually range from 3 to 30 min, based on operating conditions and fluid properties. If such laboratory data are not available, it is a common practice to use a retention time of 10 min for both oil and water.

4.5 SEPARATOR SIZING EQUATIONS AND RULES

In this section, the equations and rules used for determining the dimensions of horizontal and vertical three-phase separators are developed and presented. It should be realized that these equations are generally used for preliminary sizing of the separators. Other important aspects of the design should not be ignored. The changes in operating conditions, such as production rates, gas–liquid ratio, water–oil ratio, fluid properties, pressure, and temperature, over the life of the field should be incorporated in the design. For new-field development, there is always some degree of uncertainty in the available data and information. This should be an integral part of the facility design. Also, cost, availability, and space limitation could affect the design and selection of equipment.

The sizing procedure is generally similar to that for two-phase separator with the exception that the separation of water from oil, and oil from water are additional constraints for three-phase separators.

4.5.1 Sizing Equations for Horizontal Separators

As with two-phase separators, consideration of the gas capacity constraint and the liquid retention time constraint results in developing two equations; each relates the vessel diameter to its length. Analysis of the two equations determines the equation that governs the design and that should be used to determine possible combinations of diameters and lengths. For three-phase horizontal separators, consideration of the settling of water droplets in oil results in a third equation that determines the maximum diameter of the separator. Therefore, in determining the vessel's diameter–length combinations, the diameters selected must be equal to or less than the determined maximum diameter.

Water Droplet Settling Constraint

In comparison to two-phase separators, the additional constraint in the design of three-phase horizontal separators is that the oil retention time should be sufficient for the water droplets of certain minimum size to settle out of the oil. To be on the conservative side, we shall assume that the water droplets to be separated are at the top of the oil pad. Therefore, such droplets have to travel a distance equal to the thickness of the oil pad before they reach the water–oil interface. This constraint can be translated into a useful relationship by equating the time needed for the water droplets to travel through the oil pad thickness to the oil retention time.

The time needed for the water droplets to travel through the oil pad, t_{wd} min, is obtained by dividing the oil pad thickness, H_o (in.), by the water settling velocity given by Eq. (7); therefore,

$$t_{\mathrm{wd}} = \left(\frac{1}{60}\right)\frac{(H_o/12)}{1.787 \times 10^{-6}(\Delta\gamma)d_m^2/\mu_o} \qquad \text{min} \tag{8}$$

Equating Eq. (8) to the oil retention time, t_o, and solving for H_o, we obtain the maximum allowable oil pad thickness, $H_{o,\max}$ expressed as follows:

$$H_{o,\max} = \frac{1.28 \times 10^{-3} t_o(\Delta\gamma)d_m^2}{\mu_o} \qquad \text{in.} \tag{9}$$

The minimum water droplet diameter to be removed, d_m, is determined, as discussed earlier, from laboratory tests. In case such data are not available, d_m may be assigned the value of 500 μm.

The oil and water flow rates and retention times and the vessel diameter control the height of the oil pad. Considering a separator that is half full of liquid, the following geometrical relation is easily derived:

$$\frac{A}{A_w} = \left(\frac{1}{\pi}\right)\left[\cos^{-1}\left(\frac{2H_o}{D}\right) - \left(\frac{2H_o}{D}\right)\left(1 - \frac{4H_o^2}{D^2}\right)^{-0.5}\right] \tag{10}$$

where A_w and A are the cross-sectional area of the separator occupied by water and the total cross-sectional area of the separator, respectively, and D is the diameter of the vessel. For a given oil and water flow rates and retention times, the ratio A_w/A can be determined as follows. For a separator that is half full of liquid, the total cross-sectional area of the separator, A, is equal to twice the area occupied by the liquid, which is equal to the area occupied by water, A_w, and the area occupied by oil, A_o; therefore,

$$A = 2(A_o + A_w)$$

It follows that

$$\frac{A_w}{A} = 0.5\frac{A_w}{A_o + A_w}$$

Because the volume occupied by each phase is the product of the cross-sectional area and the effective length, the cross-sectional area is directly proportional to the volume. Further, the volume occupied by any phase is also determined as the product of the flow rate and retention time. Therefore,

$$\frac{A_w}{A} = 0.5\frac{Q_w t_w}{Q_o t_o + Q_w t_w} \tag{11}$$

Therefore, once the ratio A_w/A is determined from Eq. (11), Eq. (10) can be solved to determine the ratio H_o/D. This is then used with the value of $H_{o,\max}$ determined from Eq. (9) to determine the maximum vessel diameter associated with the maximum oil pad height according to Eq. (12):

$$D_{\max} = \frac{H_{o,\max}}{H_o/D} \tag{12}$$

This, therefore, sets the upper limit for the separator diameter. To obtain the value of H_o/D from Eq. (9) it is convenient to use the graphical solution of Eq. (10) (given in Fig. 6).

Equations similar to Eqs. (10) and (11) could be derived for other cases where the liquid may occupy more or less than half the volume of the separator.

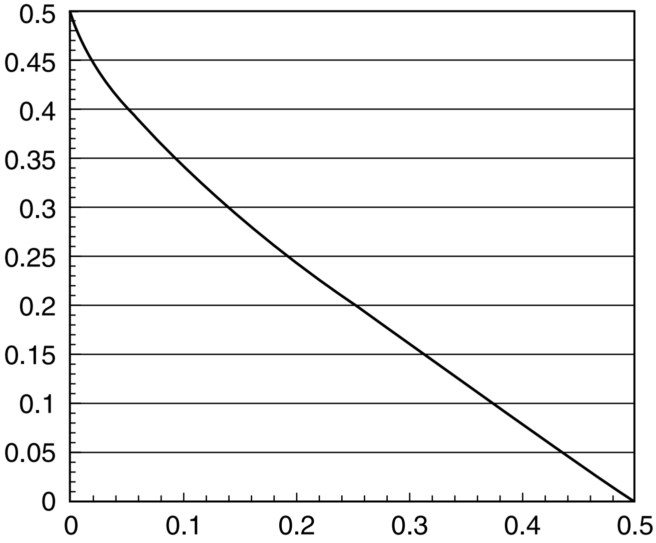

Figure 6 H_o/D as a function of A_w/A.

Gas Capacity Constraint

The gas capacity constraint equation developed for two-phase horizontal separator in Chapter 3 is also valid for three-phase horizontal separators. The equation provides a relationship between the separator diameter and effective length as follows:

$$LD = 422\left(\frac{Q_g TZ}{P}\right)\left[\left(\frac{\rho_g}{\rho_o - \rho_g}\right)\left(\frac{C_d}{d_m}\right)\right]^{1/2} \tag{13}$$

where D is the separator internal diameter (in.), L is the effective length of the separator (ft), T is the operating temperature (°R), Z is the gas compressibility at operating pressure and temperature, P is the operating pressure, ρ_g and ρ_o are the gas and oil densities, respectively (lb/ft^3), C_d is the drag coefficient, and d_m is the minimum oil droplet size to be separated from gas (μm). As discussed in Chapter 3, d_m is normally taken as 100 μm and C_d is determined by the iterative procedure described there.

Using diameters smaller than the maximum diameter determined from the above water droplet settling constraint, Eq. (13) is used to determine possible diameter and length combinations that satisfy the gas capacity constraint.

Retention Time Constraint

The separator size must provide sufficient space for the oil and water such that each phase is retained within the separator for the desired retention time. The assumption that the liquid will occupy half of the separator volume, which was used for two-phase separators, is also used here. However, in the present case, both oil and water occupy that volume. Therefore, the volume occupied by the liquid phase (both oil and water), V_l, in a separator having a diameter, D (in.), and effective length, L (ft), is given by

$$V_l = 0.5\left(\frac{\pi}{4}\right)\left(\frac{D}{12}\right)^2 L \qquad \text{ft}^3$$

Because 1 barrel (bbl) = 5.61 ft^3,

$$V_l = 4.859 \times 10^{-4} D^2 L \qquad \text{bbl} \tag{14}$$

The volume of separator occupied by oil, V_o, is the product of the oil flow rate, Q_o, and the oil retention time, t_o. If Q_o is in barrels per day (BPD) and t_o is in minutes, then

$$V_o = \frac{t_o Q_o}{24 \times 60} \qquad \text{bbl} \tag{15}$$

Similarly, the volume of separator occupied by water, V_w, is the product of the water flow rate, Q_w, and water retention time, t_w:

$$V_w = \frac{t_w Q_\text{w}}{24 \times 60} \qquad \text{bbl} \tag{16}$$

Because $V_l = V_o + V_w$,

$$4.859 \times 10^{-4} D^2 L = \frac{Q_o t_o + Q_w t_w}{24 \times 60}$$

Therefore,

$$D^2 L = 1.429(Q_o t_o + Q_w t_w) \qquad \text{in.}^2\ \text{ft} \tag{17}$$

Again, using diameters smaller than the maximum diameter determined from the above water droplet settling constraint, Eq. (12) is used to determine possible diameter and length combinations that satisfy the retention time constraint.

The procedure for determining the diameter and length of a three-phase horizontal separator can, therefore, be summarized in the following steps:

1. Determine the value of A_w/A from Eq. (11).
2. Use Figure 6 to determine the value of H_o/D for the calculated value of A_w/A.

3. Determine the maximum oil pad thickness, $H_{o,\max}$ from Eq. (9) with d_m equal to 500 μm.
4. Determine $D_{\max}$ from Eq. (12).
5. For diameters smaller than $D_{\max}$, determine the combinations of D and L that satisfy the gas capacity constraint from Eq. (13), substituting 100 μm for d_m.
6. For diameters smaller than $D_{\max}$, determine the combinations of D and L that satisfy the retention time constraint from Eq. (17).
7. Compare the results obtained in steps 5 and 6 and determine whether the gas capacity or retention time (liquid capacity) governs the separator design.
8. If the gas capacity governs the design, determine the seam-to-seam length of the separator, L_s, from

$$L_s = L + \frac{D}{12} \tag{18}$$

If the liquid retention time (liquid capacity) governs the design, determine L_s from

$$L_s = 4\frac{L}{3} \tag{19}$$

9. Recommend a reasonable diameter and length with a slenderness ratio in the range of 3–5. In making the final selection, considerations such as cost and availability will be important. It should be mentioned that, in some cases, the slenderness ratio might be different from the range of 3–5. In such cases, especially when the slenderness ratio is larger than 5, internal baffles should be installed to act as wave breakers in order to stabilize the gas–liquid interface.

Example 1

Determine the diameter and seam-to-seam length of a three-phase horizontal separator for the following operating conditions:

Oil production rate:	8000 BPD
Water production rate:	3000 BPD
Gas–oil ratio:	1000 SCF/bbl
Oil viscosity:	20 cP
Oil specific gravity:	0.89
Water specific gravity:	1.04
Gas specific gravity:	0.65
Gas compressibility:	0.89
Operating pressure:	250 psia
Operating temperature:	95°F

Oil retention time: 15 min
Water retention time: 10 min

Solution

Using Eq. (9), determine $H_{o,\max}$:

$$H_{o,\max} = \frac{1.28 \times 10^{-3} t_o(\gamma_o - \gamma_w)d_m^2}{\mu_o} \qquad \text{in.}$$

$$H_{o,\max} = \frac{1.28 \times 10^{-3}(15)(1.04 - 0.89)(500)^2}{20} = 36 \qquad \text{in.}$$

Use Eq. (11) to determine the ratio A_w/A:

$$\frac{A_w}{A} = \frac{0.5\,Q_w t_w}{Q_o t_o + Q_w t_w}$$

$$\frac{A_w}{A} = \frac{0.5(3000 \times 10)}{(3000 \times 10) + (8000 \times 15)} = 0.1$$

From Figure 6, determine the ratio H_o/D for $A_w/A = 0.1$:

$$\frac{H_o}{D} = 0.338$$

Therefore,

$$D_{\max} = \frac{H_{o,\max}}{H_o D} = \frac{36}{0.338} = 106.5 \qquad \text{in.}$$

This is the maximum allowable vessel diameter.

The gas capacity constraint, Eq. (13), yields

$$DL = 420\left(\frac{TZQ_g}{P}\right)\left(\frac{\rho_g C_d}{d_m(\rho_o - \rho_g)}\right)^{0.5}$$

$$DL = 420\left(555 \times 0.89 \times \frac{8}{250}\right)\left(\frac{\rho_g C_d}{100(\rho_l - \rho_g)^1}\right)^{0.5}$$

Determine the gas and oil densities and substitute $C_d = 0.65$ in the above equation:

$$\rho_g = \frac{2.7\gamma_g p}{TZ} = \frac{2.7 \times 0.65 \times 250}{555 \times 0.89} = 0.888 \qquad \text{lb/ft}^3$$

$$\rho_l = \rho_w \gamma_1 = 62.4 \times 0.89 = 55.54 \qquad \text{lb/ft}^3$$

Therefore, the gas capacity constraint is expressed by

$$DL = 68.22 \tag{E1}$$

Use Eq. (17) to check the liquid capacity (retention time) constraint:

$$D^2L = 1.429(3000 \times 10 + 8000 \times 15) = 214{,}350 \tag{E2}$$

Select diameters smaller than the determined maximum diameter and determine the corresponding effective length from Eqs. (E1) and (E2) for the gas capacity and liquid capacity constraints, respectively. Investigation of Eq. (E1), however, shows that for any selected diameter, the effective length is too small compared to that calculated from Eq. (E2). Therefore, the gas capacity does not govern the design. For the liquid capacity constraints, the results are tabulated as follows:

D (in.)	L (ft) [Eq. (E2)]	$L_s(=4L/3)$ (ft)	$L/(d/12)$
66	49.21	65.61	11.93
72	41.35	55.13	9.19
78	35.23	46.98	7.23
84	30.38	40.50	5.786
90	26.46	35.28	4.71
96	23.26	31.01	3.88
102	20.65	27.47	3.23

Because the most common slenderness ratio is between 3 and 5, the last three diameter and length combinations in the above table will be suitable selections. Therefore, the recommended separator size can be either 90 in. by 36 ft, or 96 in. by 31 ft, or 102 in. by 28 ft based on cost and availability. Normally, the smaller diameter and longer separator is less expensive than the larger diameter and shorter separator.

The selected separator will be able to handle a much higher gas flow rate. The actual separator gas capacity can be calculated from Eq. (8) by substituting the values of d and L and calculating the value of Q_g. For a 96-in. by 31-ft separator ($L = 3L_s/4 = 23.26$), the gas capacity is 263 MMSCFD. This is much larger than the production rate of 8 MMSCFD. This indicates that designing the separator on the basis of being half full of liquid is not efficient. The size of the separator could be made smaller by allowing the liquid to occupy more than half the volume of the separator.

4.5.2 Sizing Equation for Vertical Separators

Sizing of a vertical three-phase separator is done in a similar manner to that used in sizing vertical two-phase separators (see Chapter 3); that is, the gas capacity constraint is used to determine the minimum diameter of the vessel and the liquid retention time constraint is used to determine the height of the vessel. For three-phase separators, however, a third constraint is added. This is the requirement to settle water droplets of a certain minimum size out of the oil pad. This results, as shown in the following subsection, in a second value for the minimum diameter of the separator. Therefore, in selecting the diameter of the vessel, the larger of the minimum diameters determined from the gas capacity constraint and water settling constraint should be considered as the minimum acceptable vessel diameter.

Water Droplets Settling Constraint

The condition for the settling and separation of water droplets from the oil is established by equating the average upward velocity of the oil phase, u_o, to the downward settling velocity of the water droplets of a given size, u_w. The average velocity of the oil is obtained by dividing the oil flow rate by the cross-sectional area of flow. If Q_o is the oil rate (in BPD), then

$$u_o = \frac{Q_o \times 5.61}{24 \times 3600}\ \frac{\text{ft}^3}{\text{s}} \left(\frac{4 \times 144}{\pi D^2}\right) \text{ft}^2$$

$$u_o = 0.0119 \frac{Q_o}{D^2} \qquad \frac{\text{ft}}{\text{s}} \tag{20}$$

Equation (7) gives the water droplet settling velocity:

$$u_w = 1.787 \times 10^{-6} \frac{(\Delta\gamma) d_m^2}{\mu_o} \qquad \frac{\text{ft}}{\text{s}} \tag{21}$$

For water droplets to settle out of the oil, u_w must be larger than u_o. Equating u_w to u_o would result, therefore, in determining the minimum diameter of the separator, D_{min}, that satisfies the water settling constraint. From Eqs. (20) and (21), it follows that

$$D_{\text{min}}^2 = 6686 \frac{Q_o \mu_o}{(\Delta\gamma) d_m^2} \qquad \text{in.}^2 \tag{22}$$

where D is the separator internal diameter (in.), Q_o is the oil flow rate (BPD), μ_o is the oil viscosity (cP), $\Delta\gamma$ is the difference in specific gravity of oil and water, and d_m is the minimum water droplet size to be separated from gas (μm) microns.

Any diameter, larger than the minimum diameter determined from Eq. (22) yields a lower average oil velocity and, thus, ensures water separation.

Gas Capacity Constraint

As described in Chapter 3, the gas capacity constraint for a vertical separator yields an expression for the minimum vessel diameter as follows:

$$D_{\min}^2 = 5058 Q_g \left(\frac{TZ}{P}\right)\left(\frac{\rho_g}{\rho_o - \rho_g}\frac{C_d}{d_m}\right)^{1/2} \qquad \text{in}^2 \tag{23}$$

where D is the separator internal diameter (in.), T is the operating temperature (°R), Z is the gas compressibility at operating pressure and temperature, P is the operating pressure, ρ_g and ρ_o are the gas and oil densities (1b/ft^3), C_d is the drag coefficient, and d_m is the minimum oil droplet size to be separated from gas (μm).

Any diameter that is larger than the minimum diameter determined from Eq. (23) results in a lower gas velocity and, therefore, ensures settling and separation of liquid droplets of diameters equal to and larger than d_m out of the gas.

Liquid Retention Time (Capacity) Constraint

The separator must provide sufficient volume for the oil and water to be retained within the separator for the required retention times. The retention times are determined to allow separation of the entrained water droplets from the oil, separation of the entrained oil droplets from the water, and for the oil to reach equilibrium with the gas. As explained earlier, retention times are best determined from laboratory tests and they normally range from 3 to 30 min based on fluid properties and operating conditions. In the absence of laboratory data, a retention time of 10 min may be used for both oil and water.

Let H_o and H_w be the heights of the oil and water (in in.), respectively. Therefore, the volume of each phase within the separator is given by

$$V_o = \left(\frac{1}{12}\right)^3 \left(\frac{\pi}{4}\right) D^2 H_o \qquad \text{ft}^3$$

and

$$V_w = \left(\frac{1}{12}\right)^3 \left(\frac{\pi}{4}\right) D^2 H_w \qquad \text{ft}^3$$

Therefore,

$$V_o + V_w = 4.543 \times 10^{-4} D^2 (H_o + H_w) \qquad \text{ft}^3 \tag{24}$$

The volume is also calculated by multiplying the volumetric flow rate by the retention time; therefore,

$$V_o = Q_o \frac{5.61}{24 \times 60} \frac{\text{ft}^3}{\text{min}} \times t_o \,\text{min} \qquad \text{ft}^3$$

$$V_w = Q_w \frac{5.61}{24 \times 60} \frac{\text{ft}^3}{\text{min}} \times t_w \,\text{min} \qquad \text{ft}^3$$

Then,

$$V_o + V_w = 3.896 \times 10^{-3}(Q_o t_o + Q_w t_w) \tag{25}$$

From Eqs. (24) and (25), we obtain

$$(H_o + H_w)D^2 = 8.576(Q_o t_o + Q_w t_w) \qquad \text{in.}^3 \tag{26}$$

In summary then, the diameter and seam-to-seam length of vertical three-phase separators are determined as follows:

1. Determine the minimum diameter that satisfies the water droplets settling constraint from Eq. (22).
2. Determine the minimum diameter that satisfies the gas capacity constraint from Eq. (23).
3. The larger of the two minimum diameters determined in steps 1 and 2 is then considered as the minimum allowable vessel diameter.
4. For various values of diameter larger than the minimum allowable vessel diameter, use Eq. (26) to determine combinations of diameters and liquid heights.
5. For each combination, determine the seam-to-seam length from the following:

For $D > 36$ in.

$$L_s = \frac{1}{12}(H_o + H_w + D + 40) \qquad \text{ft} \tag{27}$$

For $D < 36$ in.

$$L_s = \frac{1}{12}(H_o + H_w + 76) \qquad \text{ft} \tag{28}$$

Example 2

Determine the diameter and seam-to-seam length of a three-phase vertical separator for the following operating conditions:

Oil production rate:	6000 BPD
Water production rate:	3000 BPD

Gas production rate: 8.0 MMSCFD
Oil viscosity: 10 cP
Oil specific gravity: 0.87
Water specific gravity: 1.07
Gas specific gravity: 0.6
Gas compressibility: 0.88
Operating pressure: 500 psia
Operating temperature: 90°F
Oil retention time: 10 min
Water retention time: 10 min
c_d: 0.64

Solution

Determine ρ_g and ρ_o:

$$\rho_g = \frac{2.7 \times 0.6 \times 500}{550 \times 0.88} = 1.674 \qquad \text{lb/ft}^3$$

$$\rho_o = 0.87 \times 62.4 = 54.288 \qquad \text{lb/ft}^3$$

Use Eq. (22) with d_m equal to 500 m to determine the minimum diameter for water droplet settling:

$$D^2_{\min} = \frac{6686 Q_o \mu_o}{(\gamma_o - \gamma_w) d_m^2} \qquad \text{in.}^2$$

$$D^2_{\min} = \frac{6686 \times 6000 \times 10}{0.2 \times 500^2} = 8023.2 \qquad \text{in.}^2$$

$$D^2_{\min} = 89.57 \text{ in.}$$

Use Equation (23) with d_m equal to 100 m to determine the minimum diameter for gas capacity constraint:

$$D^2_{\min} = 5058 Q_g \left(\frac{TZ}{P}\right)\left(\frac{\rho_g}{\rho_o - \rho_g}\frac{C_d}{d_m}\right)^{1/2}$$

$$D^2_{\min} = 5058 \times 8 \left(\frac{550 \times 0.88}{500}\right)\left(\frac{1.674}{52.614}\frac{0.64}{100}\right)^{1/2}$$

$$= 835.86 \qquad \text{in.}^2$$

$$D_{\min} = 28.911 \qquad \text{in.}$$

Therefore, the minimum allowable vessel diameter is taken as the larger of the two calculated above; that is, 89.57 in.

Use Eq. (26) to determine the relation between the diameter and liquid height for retention time constraint:

$$
\begin{aligned}
(H_o + H_w)D^2 &= 8.576(Q_o t_o + Q_w t_w) \qquad \text{in.}^3 \\
&= 8.576 \times (6000 \times 10 + 3000 \times 10) \\
&= 771{,}840 \qquad \text{in.}^3
\end{aligned}
\tag{E3}
$$

D (in.)	$H_o + H_w$ (in.)	L_s (ft)	$12L_s/d$
90	95.289	18.774	2.50
96	83.750	18.312	2.29
102	74.187	18.016	2.12
108	66.173	17.848	1.98
114	59.391	17.783	1.87
120	53.600	17.800	1.78
132	44.298	18.025	1.64
144	37.222	18.4351	1.54

For diameters larger than 89.57 in., use Eq. (E3) to determine the value of $H_o + H_w$; then, calculate the corresponding values of the seam-to-seam length from Eq. (27) and, finally, calculate the corresponding slenderness ratio. The results are summarized in the preceding table:

All combinations of diameter and seam-to-seam length in the above table are acceptable because the slenderness ratio falls between 1.5 and 3 for all of them. The final selection would, therefore, depend on cost and availability. A 96-in. by 19-ft separator is probably the best choice.

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REVIEW QUESTIONS AND EXERCISE PROBLEMS

1. State whether each of the following statements is True (**T**) or False (**F**):
    - i. The knockout drum is used to separate the bulk of free water and free gas from the produced stream.
    - ii. Separation of oil droplets from water is easier and faster than separation water droplets from oil because the density of oil is lower than that of water.
    - iii. In separator design, unless laboratory data are available, the smallest water droplet size that should be removed from oil is taken as 100 μm.
2. Circle the correct answers (note that more than one answer could be correct):
    - i. In three-phase separators, the produced fluid is separated into the following:
        - (a) Gas, oil, and water + emulsion
        - (b) Gas oil + emulsion, and water
        - (c) Gas, oil, emulsion, and water
    - ii. The formulation of emulsion in the separators may cause
        - (a) Improper oil–gas interface level control
        - (b) Reduction of oil and water effective retention times
        - (c) Blowby
        - (d) Plugging of mist extractor
    - iii. Separation of water droplets from oil compared to separation of oil droplets from water is
        - (a) More difficult because oil density is less than that of water
        - (b) Easier because water viscosity is lower than oil viscosity
        - (c) More difficult because oil viscosity is higher than water viscosity
        - (d) A more important criterion is designing the separator

3. The following data are given for a small oil field:

   Oil rate: 8000 BPD
   Water rate: 3000 BPD
   Gas–oil ratio: 2000 SCF/bbl
   Operating pressure: 500 psia
   Operating temperature: 80°F
   Oil gravity: 35°API
   Water specific gravity: 1.07
   Gas specific gravity: 0.6
   Oil viscosity: 10 cP
   Oil retention time: 600 s
   Water retention time: 360 s

   (a) Design a horizontal separator to handle the field production.
   (b) Design a vertical separator to handle the field production.

4. Determine the actual gas, oil, and water capacities of a horizontal three-phase separator having a 60-in. diameter and 15-ft seam-to-seam length for the above operating conditions.
5. Determine the actual gas, oil, and water capacities for a three-phase horizontal separator having a 48-in. diameter and a 16-ft seam-to-seam length, operating at 585 psi and 80°F, given the following data:

   Gas specific gravity: 0.6
   Water specific gravity: 1.07
   Oil specific gravity: 0.875
   Water viscosity: 1.12 cP
   Oil viscosity: 5.0 cP
   Water–oil ratio: 1 : 1.5
   Oil retention time: 600 s
   Water retention time: 480 s
   K: 0.26
   Z: 0.89

6. Determine the actual gas, oil, and water capacities of a vertical three-phase separator for the same conditions given in Problem 5. Compare the results for the two separators.

5

Emulsion Treatment and Dehydration of Crude Oil

5.1 INTRODUCTION

The fluid produced at the wellhead consists usually of gas, oil, free water, and emulsified water (water–oil emulsion). Before oil treatment begins, we must first remove the gas and free water from the well stream. This is essential in order to reduce the size of the oil–treating equipment.

As presented in Chapters 3 and 4, the gas and most of the free water in the well stream are removed using separators. Gas, which leaves the separator, is known as ''primary gas.'' Additional gas will be liberated during the oil treatment processes because of the reduction in pressure and the application of heat. Again, this gas, which is known as ''secondary gas,'' has to be removed. The free water removed in separators is limited normally to water droplets of 500 μm and larger. Therefore, the oil stream leaving the separator would normally contain free water droplets that are 500 μm and smaller in addition to water emulsified in the oil. This oil has yet to go through various treatment processes (dehydration, desalting, and stabilization) before it can be sent to refineries or shipping facilities.

This chapter deals with the dehydration stage of treatment. The objective of this treatment is first to remove free water and then break the oil emulsions to reduce the remaining emulsified water in the oil. Depending on the original water content of the oil as well as its salinity and the process of dehydration used, oil-field treatment can produce oil with a remnant water content of between 0.2 and 0.5 of 1%. The remnant water is normally called the *bottom sediments and water* (B.S.&W.). The treatment process and facilities should be carefully selected and designed to meet the contract requirement for B.S.&W. Care should be taken not to exceed the target oil dryness. Removal of more remnant water than allowed by contract costs more money while generating less income because the volume of oil sold will be based on the contract value of the B.S.&W.

The basic principles for the treating process are as follows

1. Breaking the emulsion, which could be achieved by either any, or a combination of the addition of heat, the addition of chemicals, and the application of electrostatic field
2. Coalescence of smaller water droplets into larger droplets
3. Settling, by gravity, and removal of free water

The economic impact of these treating processes is emphasized by Abdel-Aal *et al.* [8].

5.2 OIL EMULSIONS

Rarely does oil production takes place without water accompanying the oil. Salt water is thus produced with oil in different forms as illustrated in Figure 1. Apart from free water, emulsified water (water-in-oil emulsion) is the one form that poses all of the concerns in the dehydration of crude oil.

Oil emulsions are mixtures of oil and water. In general, an emulsion can be defined as a mixture of two immiscible liquids, one of which is dispersed as droplets in the other (the continuous phase), and is stabilized by an emulsifying agent. In the oil field, crude oil and water are encountered as the two immiscible phases together. They normally form water-in-oil emulsion (W/O emulsion), in which water is dispersed as fine droplets in the bulk of oil. This is identified as type C in Figure 2. However, as the water cut increases,

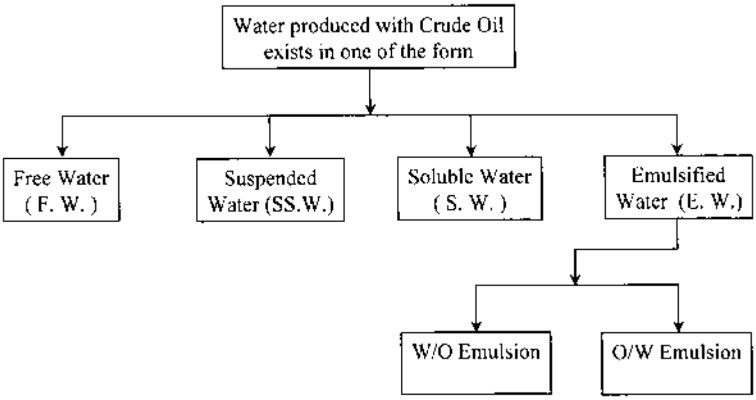


Figure 1 Forms of saline water produced with crude oil.

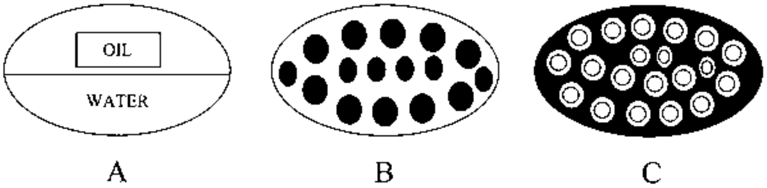


Figure 2 Schematic representation of (A) a non-dispersed system, (B) an O/W emulsion, and (C) a W/O emulsion.

the possibility of forming reverse emulsions (oil-in-water, or O/W emulsion) increases. This is type B in Figure 2.

For two liquids to form a stable emulsion, three conditions must exist:

1. The two liquids must be immiscible.
2. There must be sufficient energy of agitation to disperse one phase into the other.
3. There must be the presence of an emulsifying agent.

Conditions 2 and 3 are discussed in the following subsections.

5.2.1 Energy of Agitation

Emulsions normally do not exist in the producing formation, but are formed because of the agitation that occurs throughout the oil production system. Starting within the producing formation, the oil and water migrate through the porous rock formation, making their way into the wellbore, up the well tubing, through the wellhead choke, and through the manifold into the surface separators. Throughout this journey, the fluids are subjected to agitation due to the turbulent flow. This energy of agitation, which forces the water drops in the bulk of oil, functions in the following pattern:

First, energy is spent to overcome the viscous force between the liquid layers, leading to their separation into thin sheets or parts. This is what we call "shearing energy" and is mathematically approximated by the formula

$$\mathrm{SE} = \tau \mathrm{AD_o} \tag{1}$$

where SE is the shearing energy, A is the shear surface area, D_o is the characteristic length, and τ is the shearing force per unit area, which is defined by Eq. (2) as follows:

$$\tau = \frac{C_d \rho u^2}{2g_c} \tag{2}$$

where C_d is the drag coefficient, ρ is the density of the fluid, v is the velocity of flow, and g_c is a conversion factor (lbmft/lbfsec2) (lbm)(ft)/(lbf)(sec^2).

Second, energy is used in the formation of "surface energy," which occurs as a result of the separation of the molecules at the plane of cleavage. This surface energy is related to the surface tension, which involves the creation of an enormous area of interface with attendant free-surface energy. Energy contained per unit area is referred to as "surface tension," having the units of dynes/cm.

The drops attain the spherical shape, which involves the least energy contained for a given volume. This is in accordance with the fact that all energetic systems tend to seek the lowest level of free energy [4]. Because the surface tension is defined as "the physical property due to molecular forces existing in the surface film of the liquid," this will cause the volume of a liquid to be contracted or reduced to a shape or a form with the least surface area. This is the same force that causes raindrops to assume a spherical shape.

A schematic presentation of energy utilization in emulsion formation is given in Figure 3. A crucial question that can be asked now is the following: Can the plant designer prevent emulsion formation? Well, the best he can do is to reduce its extent of formation based on the fact that the liquids initially are not emulsified. From the design point of view, primarily reducing the flowing velocity of the fluid and minimizing the restrictions and sudden changes in flow direction could minimize formation of emulsion.

5.2.2 Emulsifying Agents

If an oil emulsion is viewed through a microscope, many tiny spheres or droplets of water will be seen dispersed through the bulk of oil, as depicted

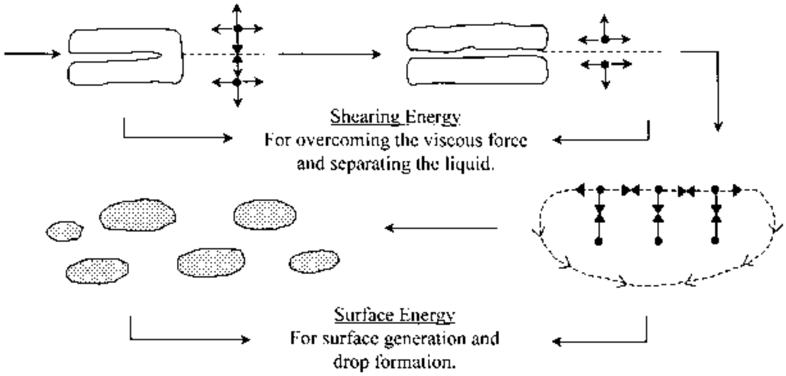


Figure 3 Forms of energy participating in emulsification.

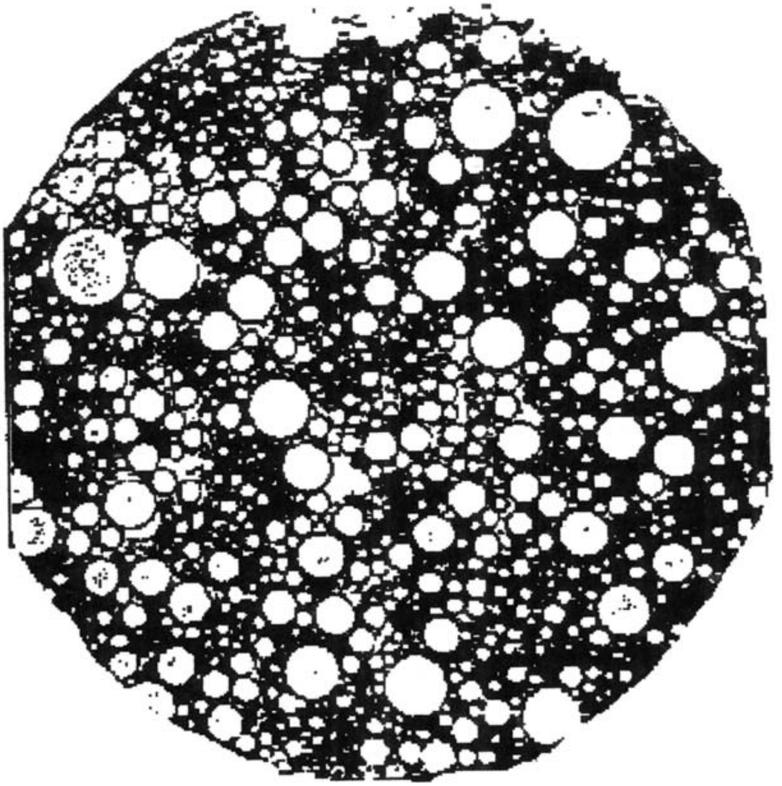


Figure 4 Photomicrograph of loose emulsion containing about 30% emulsified water in the form of droplets ranging in diameter from about 60 μm downward.

in Figure 4. A tough film surrounds these droplets; this is called a stabilizing film. Emulsifying agents, which are commonly found in crude oil or water in the natural state or introduced in the system as contaminants during drilling and/or maintenance operations, create this film. Some of the common emulsifiers are as follows:

1. Asphaltic materials
2. Resinous substances
3. Oil-soluble organic acids
4. Finely dispersed solid materials such as sand, carbon, calcium, silica, iron, zinc, aluminum sulfate, iron sulfide, and so on.

These emulsifying agents support the film formation encasing the water droplets, hence the stability of an emulsion.

The stability of oil–water emulsions could be viewed through the following analysis. The relative difficulty of separating an emulsion into two phases is a measure of its stability. A very stable emulsion is known as a "tight" emulsion and its degree of stability is influenced by many factors. Accordingly, we can best understand the resolution problem and, hence, the treatment procedure if we consider the following factors:

1. *Viscosity of oil*: Separation is easier for a less viscous oil phase.
2. *Density or gravity difference between oil and water phases*: Better separation is obtained for a larger difference.
3. *Interfacial tension between the two phases* (which is related to the type of emulsifying agent): Separation is promoted if this force is lowered (i.e., decreasing the interfacial tension).
4. *Size of dispersed water droplets*: The larger the size of water drops, the faster is the separation. This could be readily concluded from Figure 5, which relates the velocity of settling of emulsified water drops to the diameter for different temperatures. The size of dispersed water droplets is an important factor in emulsion stability. A typical droplet size distribution for emulsion samples was determined by using a special computer scanning program. Results reported in Figure 6 [2] indicate that most of the droplets found in oil emulsions are below 50 µm.

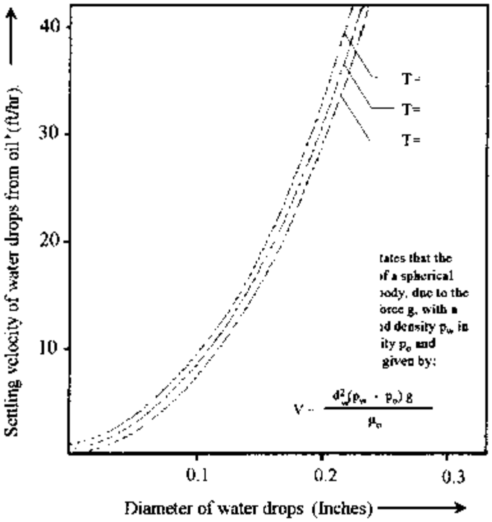


Figure 5 The change of settling velocity with water drop size for different operating temperatures (40°API oil).

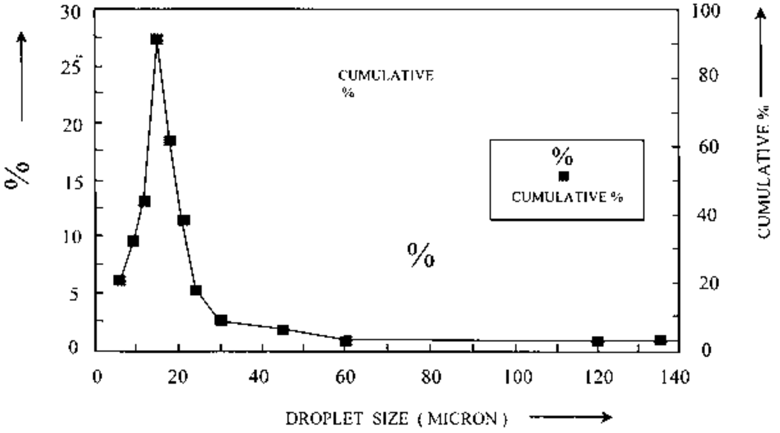


Figure 6 Droplet size distribution for an emulsion sample.

5. *Percentage of dispersed water*: The presence of a small percentage of water in oil under turbulence conditions could lead to highly emulsified mixture. Water droplets are finely divided and scattered with very little chance of agglomerating to larger particles.
6. *Salinity of emulsified water*: Highly saline water will lead to a faster separation because of a higher density difference between the oil and the water phases.

5.3 DEHYDRATION/TREATING PROCESSES

The method of treating “wet” crude oil for the separation of water associated with it varies according to the form(s) in which water is found with the crude. Free-water removal comes first in the treating process, followed by the separation of “combined” or emulsified water along with any foreign matter such as sand and other sediments [1]. The basic approaches of handling “wet” crude oils are illustrated in Figure 7.

Again, from an economic point of view, removal of free water at the beginning will reduce the size of the treating system, hence its cost. The same applies for the separation of associated natural gas from oil in the gas–oil separator plant (GOSP).

A dehydration system in general comprises various types of equipment. Most common are the following:

Free-water knockout vessel
Wash tank

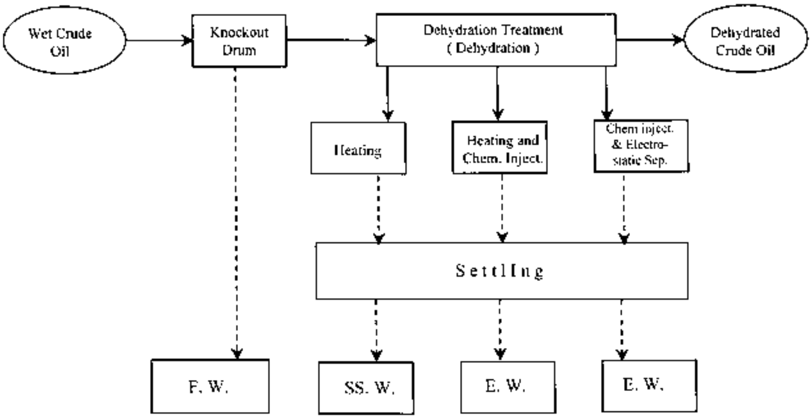


Figure 7 Basic approach of handling wet crude oil. (F.W. = fresh water, SS.W. = suspended water, E.W. = emulsified water.)

Gun barrel
Flow treater (heater/treater)
Chemical/Injector
Electrostatic dehydrator

5.3.1 Removal of Free Water

Free water is simply defined as that water produced with crude oil and will settle out of the oil phase if given little time. There are several good reasons for separating the free water first:

1. Reduction of the size of flow pipes and treating equipment
2. Reduction of heat input when heating the emulsion (water takes about twice as much heat as oil)
3. Minimization of corrosion because free water comes into direct contact with the metal surface, whereas emulsified water does not.

Free water, on the other hand, has its distinctive benefits. Free water found in the reservoir fluid will carry twice as much heat as oil and take it up the tubing to the surface. Eventually, it will help in breaking oil emulsions. It is to be observed that a well producing salt water (free water) will be much warmer than a well producing oil only. Further, free water contributes to what is called “water wash,” which is the action of the salt water to break the oil emulsions (as will be discussed in Chapter 6).

Free water removal takes place using a knockout vessel, which could be an individual piece of equipment or incorporated in a flow treater.

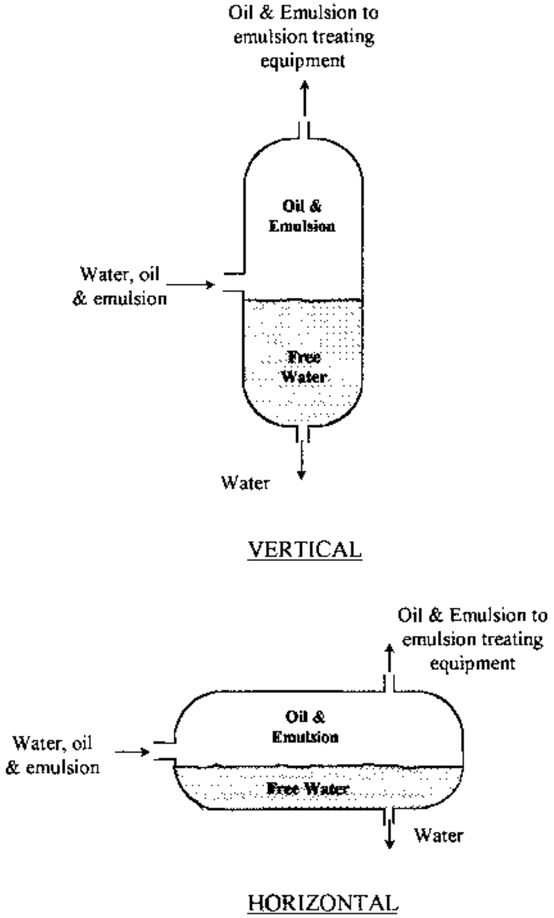


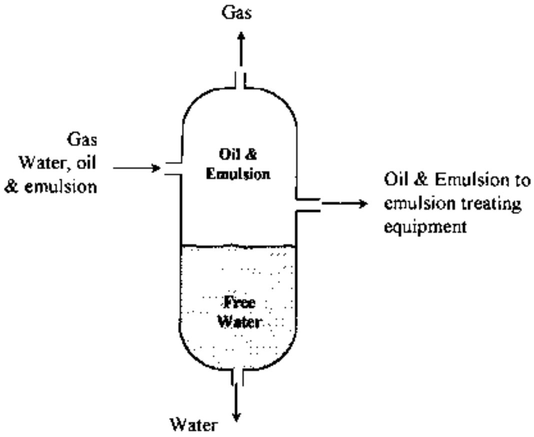
Figure 8 Two-phase free-water knockouts.

Figures 8 and 9 show some of the common types of two-phase and three-phase free-water knockout drums, respectively.

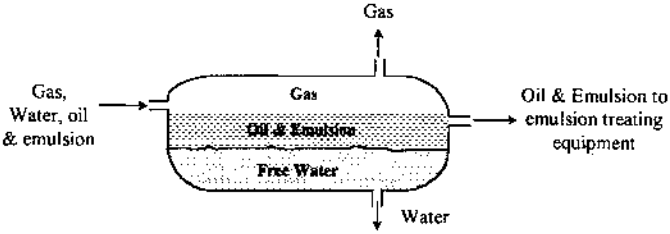
5.3.2 Resolution of Emulsified Oil

This is the heart of the dehydration process, which consists of three consecutive steps:

1. *Breaking the emulsion*: This requires weakening and rupturing the stabilizing film surrounding the dispersed water droplets.



VERTICAL



HORIZONTAL

Figure 9 Three-phase free-water knockouts.

This is a destabilization process and is affected by using what is called an "aid," such chemicals and heat.

2. *Coalescence*: This involves the combination of water particles that became free after breaking the emulsion, forming larger drops. Coalescence is a strong function of time and is enhanced by applying an electrostatic field, impingement on a solid surface area, and water washing.
3. *Gravitational settling and separation of water drops*: The larger water droplets resulting from the coalescence step will settle out of the oil by gravity and be collected and removed.

Because these steps are in series, the slowest one is the most controlling. Out of these, coalescence is the slowest step. In other words, using either heat or chemicals followed by gravitational settling can break some emulsions, but the process is dependent on the time spent in coalescence. This time is the element that determines the equipment size, hence its capital cost.

5.3.3 Treating the Emulsion

As explained earlier, using chemicals followed by settling can break some emulsions. Other emulsions require heating and allowing the water to settle out of the bulk of oil. More difficult (tight) emulsions require, however, both chemicals and heat, followed by coalescence and gravitational settling.

Basically, a dehydration process that utilizes any or a combination of two or more of the treatment aids mentioned earlier (heating, adding chemicals, and an applying electrical field) is used to resolve water–oil emulsions [6,7]. The role of each of these aids is discussed next in detail.

5.4 HEATING

Heating is the most common way of treating water–oil emulsions. To understand how heating aids in the resolution of water–oil emulsions and separation of the water droplets from the bulk of oil, reference is made to the droplet settling velocity equation derived in Chapter 4 [Eq. (7)]:

$$u = 1.787 \times 10^{-6} \frac{(\Delta\gamma)d_m}{\mu_o} \text{ ft/sec} \tag{3}$$

where u is the water droplet settling velocity, $\Delta\gamma$ is the difference between water and oil specific gravities, d_m is the diameter of the water droplet (in μm) and μ_o is the oil viscosity.

5.4.1 Benefits and Drawbacks of Heating

Heating of water–oil emulsions aids in the resolution of the emulsion and the separation of the emulsified water in several ways. The most significant effect is the reduction of oil viscosity with temperature. The viscosity of all types of crude oil drops rapidly with temperature. From Eq. (1), such reduction in viscosity results in increasing the water droplet settling velocity and, thus, speeds and promotes the separation of water from the oil.

As the water and oil mixture is heated, the density (specific gravity) of both water and oil is reduced. However, the effect of temperature on oil density is more pronounced than on water density. The result is that the difference in density (or specific gravity) increases as the emulsion is heated. For example, if oil and water are heated from 60°F to 150°F, the following change in their relative specific gravity takes place:

	at 60°F	at 150°F
Oil specific gravity	0.83	0.79
Water specific gravity	1.05	1.03
Difference in specific gravity	0.22	0.24

With reference to Eq. (1) an increase in $\Delta\gamma$ increases the settling velocity and, therefore, promotes the separation of water droplets from the bulk of oil. The change in the specific gravity difference is, however, small. Therefore, this effect is not as significant as the effect of viscosity. In fact, we may completely ignore the effect of specific gravity on the process up to a temperature of 200°F. For some specific crude oils, increased temperature may cause a reverse effect on the difference in specific gravity. For some heavy oils, the specific gravity of the oil and water will be equal at certain temperature. This situation must be avoided, as it will stop the separation process completely. Therefore, care should be exercised when determining the treating temperature for a specific crude oil.

Another beneficial effect of heating is that the increased temperature promotes movements of the small water droplets, which upon collision with one and the other, may form larger size droplets. The increased droplet size significantly speeds the settling process, as indicated by Eq. (1). Heat will also help to destabilize (weakening) the emulsifying film, hence breaking the emulsion. Further, heating will dissolve the small paraffin and asphaltenes crystals and, thus, neutralize their potential effect as emulsifiers.

Despite of all of the above-discussed benefits of heating, there are some drawbacks associated with this method of treatment. Heating of the oil can result in significant loss of the lighter hydrocarbon ends and thus results in loss of the oil volume. For example, heating 35°API oil from 100°F to 150°F results in losing more than 1% of the oil volume. Of course, the evaporates (light ends) could be collected and sold with the gas. This, however, will not make up for the loss of revenue resulting from oil losses. In addition to oil losses, evaporation of the light ends leaves the treated oil with lower API gravity (i.e., lower quality), which will be sold at a lower price. Finally, heating requires additional investment for heating equipment and additional operating cost for the fuel gas and maintenance.

Because of the above drawbacks, it is generally recommended to avoid using heating as a treatment process if at all possible. Otherwise, some of the benefits of heating may be realized with the minimum amount of heating.

5.4.2 Methods of Heating Oil Emulsions

The fuel used to supply heat in oil-treating operations is practically natural gas. Under some special conditions, crude oil may be used.

Heaters are generally of two basic types:

1. *Direct heaters*, in which oil is passed through a coil exposed to the hot flue gases of the burned fuel or to introduce the emulsion into a vessel heated using a fire tube heater.
2. *Indirect heaters*, in which heat is transferred from the hot flue gases to the emulsion via water as a transfer medium. The emulsion passes through tubes immersed in a hot water bath.

In general, the amount of free water in the oil emulsion will be a factor in determining which method is to be used. If free water is found to be 1–2%, then use an indirect heater. If the free-water content is more enough to hold a level around the fire tube, then use a direct heater. Both types are shown in Figure 10.

5.4.3 Heat Requirement

The amount of heat transferred to, or gained by, a fluid is generally given by

$$q = mc\Delta T \qquad \text{Btu/h} \tag{4}$$

where q is the rate of heat transferred/gained (Btu/h), m is the mass flow rate of fluid (lb/h), c is the specific heat of fluid (Btu/lb-°F), and ΔT is the increase in temperature due to heat transferred (°F).

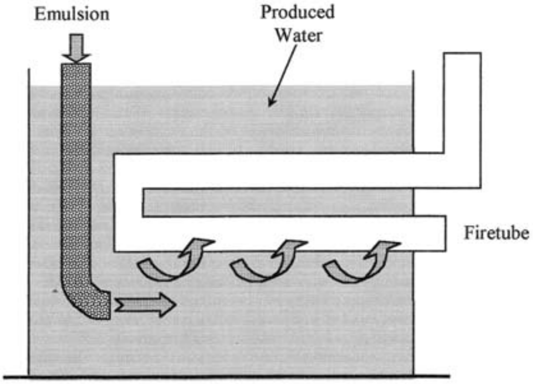
The mass flow rate, m, could be calculated from the volumetric flow rate, Q, that is normally given in BPD (bbl/day) as follows:

$$m = Q\left(\frac{\text{bbl}}{\text{day}}\right)\left(\frac{1}{24}\right)\left(\frac{\text{day}}{\text{h}}\right)5.61\left(\frac{\text{ft}}{\text{bbl}}^{3}\right)62.4\gamma\left(\frac{\text{ft}^3}{\text{lb}}\right)\ \frac{\text{lb}}{\text{h}}$$

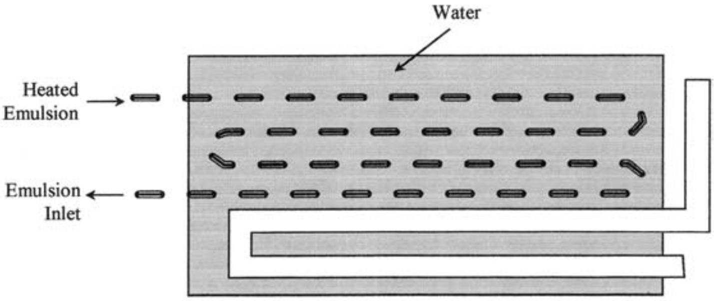
Therefore,

$$m = 14.59\gamma Q \cong 15\gamma Q \ \ \text{lb/h} \tag{5}$$

where γ is the specific gravity of the liquid.



Direct Heating



Indirect Heating

Figure 10 Methods of heating the emulsion.

When heating is used for emulsion treatment, the bulk of the free water must be removed prior to the application of heat (as described in Chapter 4). The remaining water would be mostly emulsified water and some free water that was not separated (water droplets of 500 μm and smaller). In estimating the total amount of heat required for the treatment (q), we should account for the heat transferred to the water (q_w) along with that transferred to the oil (q_o). Further, we should also account for the heat losses (q_l). These could be estimated using Eqs. (2) and (3) as follows:

$$q_o = 15\gamma_o Q_o c_o (\Delta T) \tag{6}$$

Normally, the amount of water is given as a percentage of the oil volume (i.e., $Q_w = wQ_o$, where w is the water percentage). Therefore,

$$q_w = 15\gamma_w w Q_o c_w(\Delta T) \tag{7}$$

The heat loss is normally expressed as a percentage of the total heat input. Let l be the percent of total heat lost; then,

$$q_l = lq \tag{8}$$

Because $q = q_o + q_w + q_l$, adding Eqs. (6)–(8) and solving for q, we obtain

$$q = \frac{1}{1-l} 15 Q_o(\Delta T)(\gamma_o c_o + w\gamma_w c_w)\ \frac{\text{Btu}}{\text{h}} \tag{9}$$

where, l is the percent of heat loss (fraction), Q_o is the flow rate of oil (bbl/day), ΔT is the increase in temperature due to heating (°F), γ_o and γ_w are the specific gravity of oil and water, respectively, c_o and c_w are the specific heats of oil and water, respectively, and w is the percent, by volume, of water in the oil (fraction). Equation (9) is used to estimate the required heat duty of the burner.

5.4.4 Types of Heater Treater

The various types of field heater and heater treater and gunbarrel treaters are presented in Figures 11 and 12, respectively.

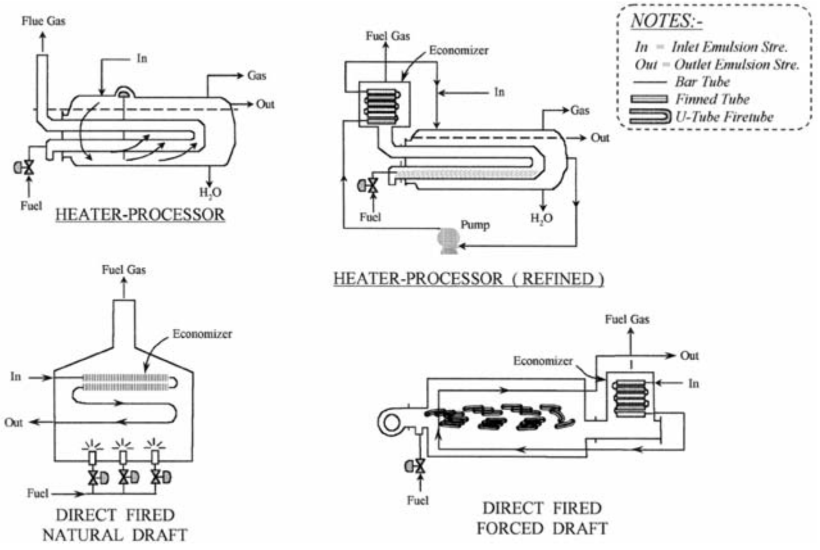
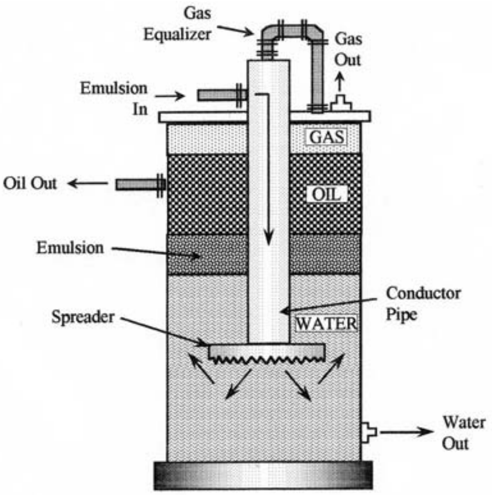
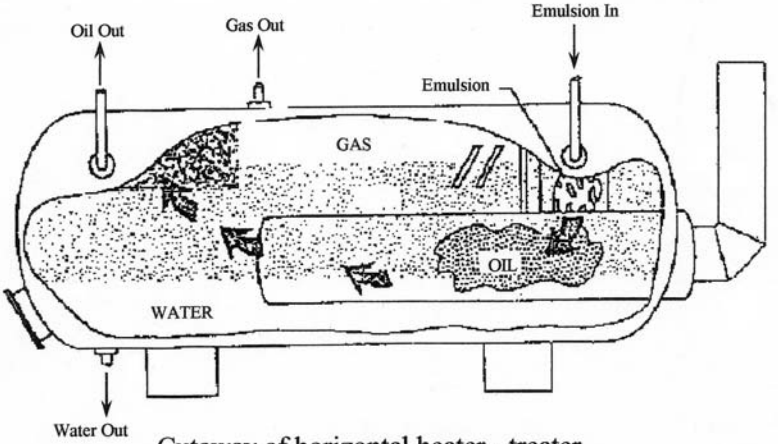


Figure 11 Types of field heater.



Gun barrel with internal conductor pipe and spreader.



Cutaway of horizontal heater - treater.

Figure 12 Details of heater treater.

Vertical Treaters

Vertical treaters are commonly used as single-well treaters. The oil + emulsion stream enters the treater from the side at the top section of the vessel where gas, if any, separates and leaves the vessel at the top through a mist extractor. The liquid flows downward through a pipe called the downcomer and exits through a flow spreader located slightly below the water–oil interface to *water wash* the oil–emulsion stream. Water washing helps in coalescing the small water droplets suspended in the oil. The oil and emulsion flow upward, exchanging heat with the heater fire tubes, then through the coalescing section. The coalescing section, normally packed with porous material such as hay, is sized to provide sufficient time for the coalescence of the water droplets and their settling out of the oil. The treated oil is then collected from the treater.

Horizontal Treaters

This type of treater is normally used in centralized multiwell-treating facilities (GOSP). The oil and emulsion stream is introduced to heating section of the treater near the top where gas is flashed, separated, and exits the vessel at the top through a mist extractor. The liquid is made to flow tangent to the inside surface of the vessel and falls below the water–oil interface, where it is water washed. Water washing causes coalescence and separation of free water. The oil + emulsion rises up, exchanging heat with the fire tubes, and flows over a weir into an oil surge chamber. The hot oil + emulsion leaves the oil surge chamber near the bottom of the vessel and enters the coalescing section of the treater through a flow spreader, which ensures that the oil flows evenly throughout the length of the coalescing section. The oil flows upward, where it is withdrawn from the vessel through a collector. The spreader–collector system allows the oil flow to be vertical. This section of the treater is sized to allow sufficient retention time for the coalescence and settling of the water out of the oil. The separated water is removed from the treater at two locations: one at the bottom of the heating section and the other at the bottom of the coalescing section. Interface level controllers control both outlet valves.

Gunbarrel Settling Tanks

Gunbarrel tanks are large-diameter vertical tanks operating mostly at atmospheric pressure. They are generally used for small fields where no or minimum heating is required for separation of the emulsion. When heating is needed, the most common way is to preheat the oil and emulsion stream before it enters the tank.

The oil + emulsion stream enters the tank at the top (where gas is flashed and separated) into a downcomer. It leaves the downcomer through a spreader located below the water–oil interface and rises vertically upward, flowing through the large cross-sectional area of the tank. As the oil + emulsion rises, it is first water washed to coalesce the water droplets. Then, it is retained for a sufficient time in the settling section to allow for the separation of the water droplets, which flow countercurrent to the oil flow and collects at the bottom section of the tank.

5.5 CHEMICAL TREATMENT

As mentioned earlier, some oil emulsions will readily break upon heating with no chemicals added; others will respond to chemical treatment without heat. A combination of both “aids” will certainly expedite the emulsion-breaking process [5]. Chemical additives, recognized as the second “aid,” are special surface-active agents comprising relatively high-molecular-weight polymers. These chemicals (deemulsifiers), once adsorbed to the water–oil interface, can rupture the stabilizing film and/or displace the stabilizing agent due to the reduction in surface tension on the inside of the film (i.e., on the water side of the droplet). In other words, when the deemulsifiers are added to the oil, they tend to migrate to the oil–water interface and rupture the stabilizing film, as depicted in Figure 13. A deemulsifier, as it reaches to oil–water interface, functions in the following pattern: flocculation, then film rupture, followed by coalescence. The faster the deemulsifier reaches the oil–water interface, the better job it achieves. Figure 14 illustrates these steps.

5.5.1 Selection and Injection of Chemicals (Deemulsifiers)

The very first step for selecting the proper chemical for oil treating is testing an oil sample. The representative sample is measured into a number

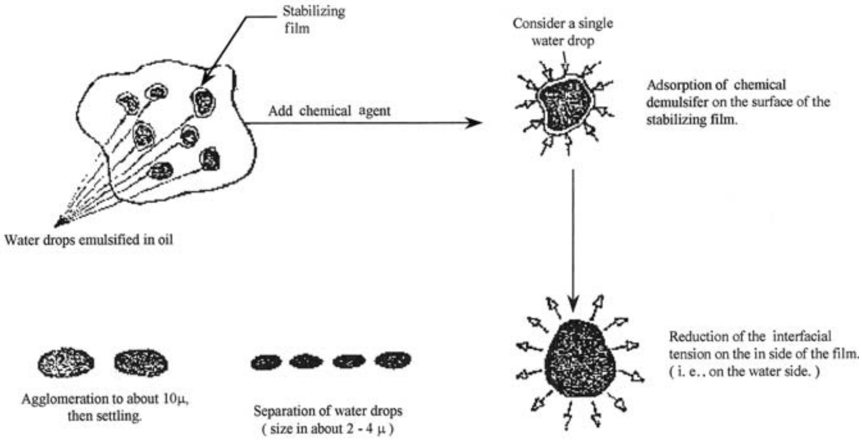


Figure 13 Action of chemical agents in deemulsification of water drops.

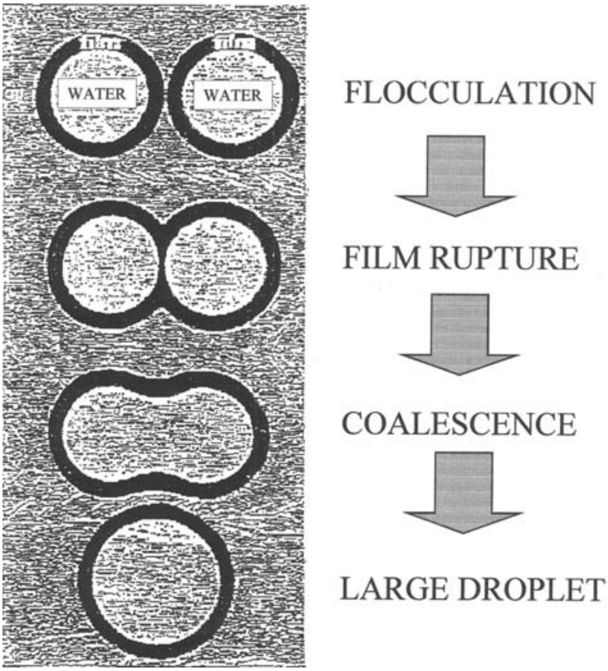


Figure 14 Steps leading to large drop formation.

of bottles (12 or more). To each bottle, a few drops of different chemicals are added, followed by shaking to ensure good mixing between the emulsion and the chemical. Heat could be applied if needed. Final selection of the right chemical will be based on testing a sample of the oil to find out how complete the water removal was.

From the practical point of view, most oil deemulsifiers are oil soluble rather than water soluble. Because such small amounts are used in treating and to ensure thorough mixing, it is recommended to dilute the chemical with a solvent to have a larger volume of the solution to inject.

The point of injection of deemulsifiers will depend largely on the type used. For the case of water-soluble deemulsifiers, injection is carried out after free water has been removed; otherwise, most of the chemical is lost down the drain. Three points of injection are recommended:

1. Upstream of the choke, where violent agitation takes place in the choke as the pressure is lowered from wellhead to that corresponding to the gas/oil separator. It is considered the ideal injection point. This is illustrated in Figure 15.

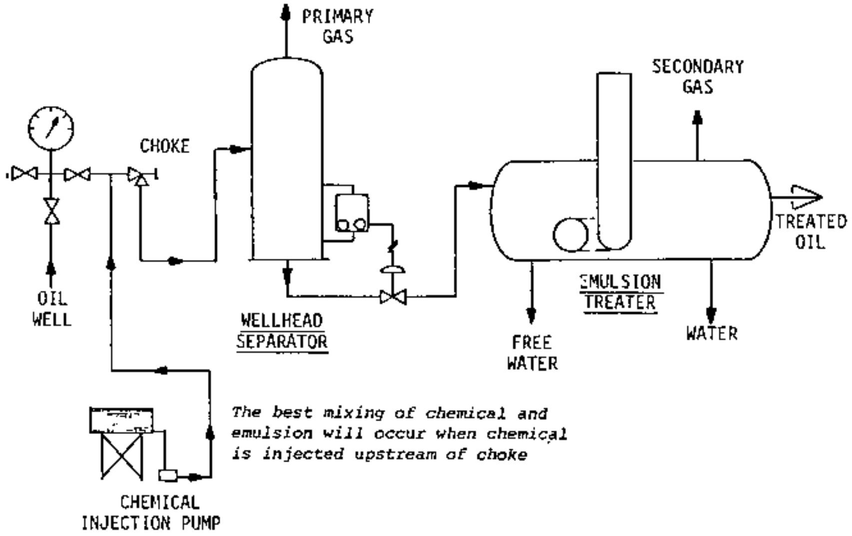


Figure 15 Chemical injection: upstream of the choke.

2. Upstream of the level control valve on the separator, where agitation occurs in the valve as the pressure is lowered. This is illustrated in Figure 16.
3. For the case in which the treating system does not include a gas–oil separator, the injection point is placed 200–250 ft from the emulsion treater.

Chemicals are applied and injected using a small chemical pump. The pump is of the displacement plunger type. The chemical pump should be able to deliver small quantities of the deemulsifier into the oil line. At normal treating conditions, 1 L of chemical is used for each 15–20 m^3 of oil (or about 1 qt per 100 bbl of oil). As was stated earlier, dilution of the chemical with proper solvents is necessary. Based on the type of oil, the required concentration of the chemical ranges between 10 and 60 ppm (parts per million) [7]. Chemical deemulsifiers (emulsion breakers) are complex organic compounds with surface-active characteristics. A combination of nonionic, cationic and anionic materials contributes the surface-active properties. Some of the deemulsifiers are sulfonates, polyglycol esters, polyamine compounds, and many others.

A final and important word is that excessive amounts of chemicals can do harm. Too much chemicals is usually called overtreating. In addition to the unnecessary additional operating cost, excessive treatment

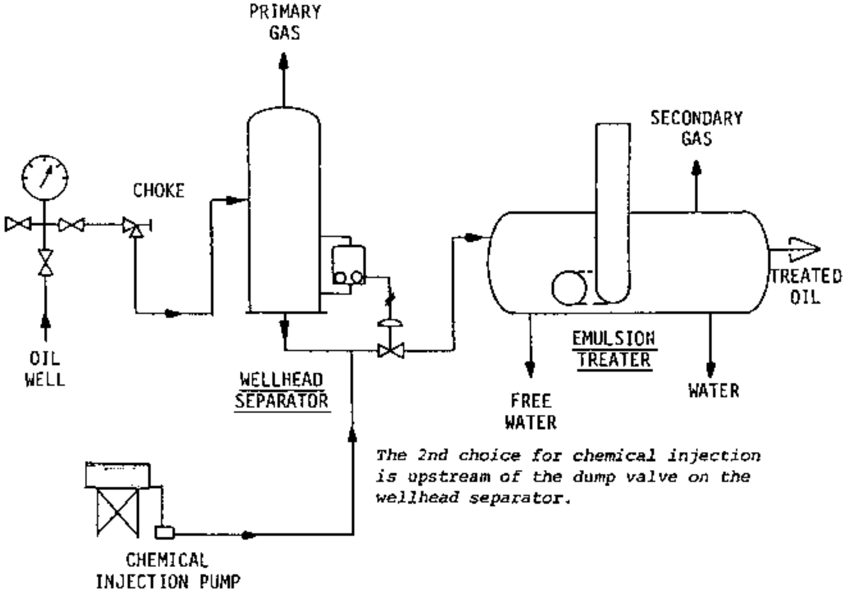


Figure 16 Chemical injection: upstream of the level control valve of the gas–oil separator.

would lead to what is known as "burning of the emulsion" (i.e., unbreakable or tight emulsion).

5.6 ELECTRICAL AID

This is the third aid of emulsion treating in crude oil dehydration. However, it should be realized that both heating and chemical treating work in order to "break the emulsion", whereas electrical emulsion treating is aimed at speeding up "coalescence," hence settling. In other words, electric dehydration does not break the emulsion electrically.

Looking at the three consecutive steps involved in the dehydration of emulsified crude oils (breaking the emulsion, coalescence of water droplets, and settling and separation) and assuming that the first and third steps are fast compared to the second step, it can be concluded that coalescence is the controlling step. In other words, coalescence, which is a function of time, influences settling. Consequently, in the design of dehydrators, some means should be implemented to reduce the coalescence time, hence the settling time. Some of these means are (1) installing a coalescing medium in the settling section to speed up the buildup and the formation of water drops, (2) applying centrifugal force to the emulsion that can promote separation, and (3) applying an electrical field in the settling section of the treater.

The principle of breaking oil–water emulsions using electrical current is known as electrostatic separation. Ionization of these emulsions with the aid of electric field was introduced in 1930 for crude oil desalting in oil refineries. High-voltage field (10,000 to 15,000 v) is used to help dehydration according to the following steps:

1. The water droplet is made up of polar molecules, because the oxygen atom has a negative end, and the hydrogen atoms have positive charges. These polar forces are magnetized and respond to an external electrical force field. Therefore, a dipole attraction between the water droplets in the emulsion is established, leading to coalescence, hence settling and separation (see Fig. 17).
2. As a result of the high-voltage field, the water droplets vibrate rapidly, causing the stabilizing film to weaken and break.
3. The surface of the water droplets expand (their shapes change into ellipsoids); thus attracted to each other, they collide and then coalesce, as depicted in Figure 18.
4. As the water droplets combine, they grow in size until they become heavy enough to separate by settling to the bottom of the treater.

5.7 CHEMIELECTRIC DEHYDRATROS (EMULSION TREATERS)

It is normal practice to call emulsion treaters "heater treaters." However, when other or additional treating aids are used, the name of the treater would be made to reflect such aids of treatment. Consequently, a name such as *chemielectric dehydrator* is used to indicate that both chemical and electrical aids are used (in addition to heating) in the treatment.

Figure 19 is a sketch illustrating a typical chemielectrical treater. The vessel is basically the same as the horizontal heater treater described earlier. Once the oil is heated, it flows to the settling section. Free water, which separates from the emulsion (under the effect of both heat and chemicals), settles to the bottom. The oil on the other hand moves slowly upward, passing across the electric grid in the settling section, where

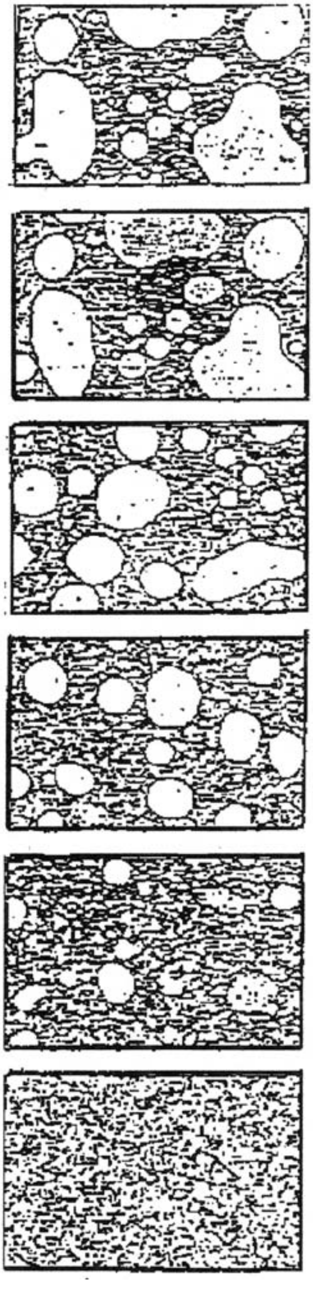


Figure 17 Coalecence of water droplets in W/O emulsion.

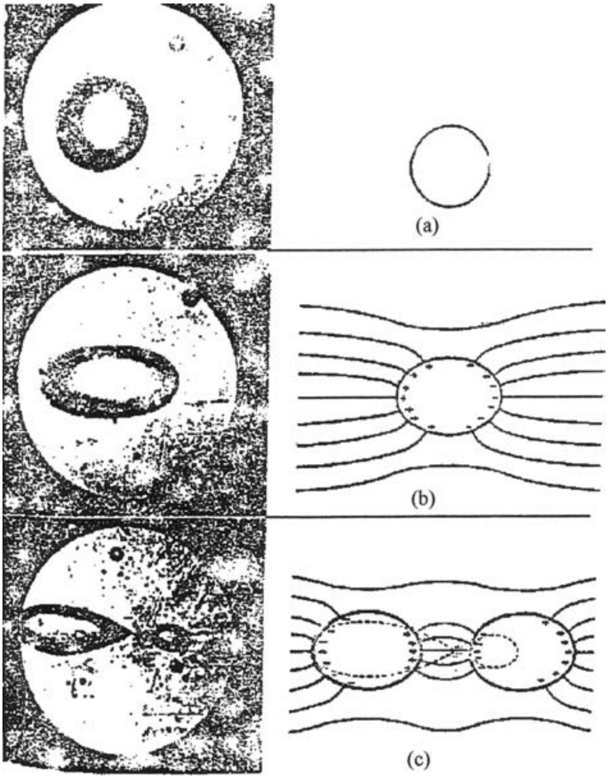


Figure 18 Emulsion breaking with the aid of an electric current.

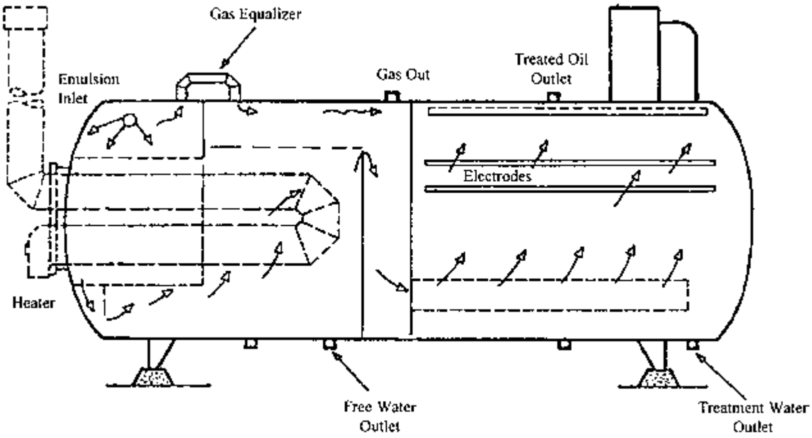


Figure 19 Chemelectric dehydrator.

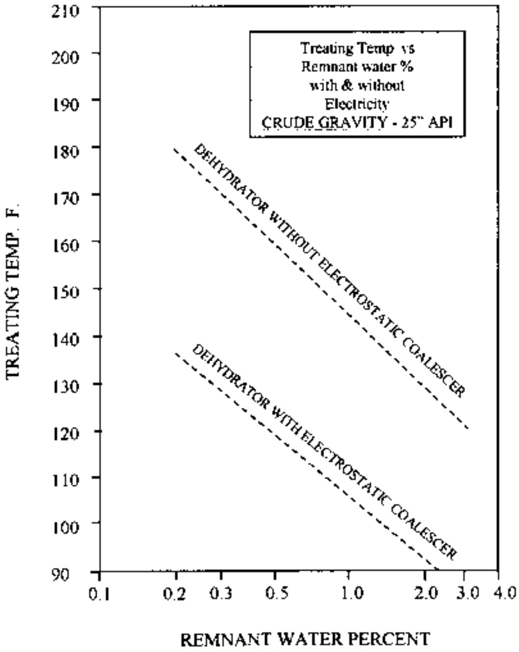


Figure 20 Effect of an electrical field on coalescence.

remnant emulsified water is separated as explained earlier. Finally, clean oil flows to the top of the treater.

It should be made clear that most of the emulsified water is removed by the dual action of both heat and chemicals before the oil passes to the electric grid. The water content in the oil could be reduced to 1–0.5% before it gets to the grid.

The application of the electrical field has a significant impact on the treater's performance. This is exemplified by Figure 20, which clearly illustrates the improvement of a treater when an electric field is applied.

5.8 DESIGN OF TREATERS

In the present discussion, we shall limit our design consideration to determination of the dimensions of the coalescing section and the heat input requirement.

In order to achieve a certain quality for the treated oil (i.e., certain maximum water content), we must determine the smallest water droplet size that must be removed. The temperature of the emulsion, the retention time, and the viscosity of the oil affect the water droplet size that must be removed.

Heating the emulsion results in a state of excitation where water droplets collide and coalesce into larger droplets. Therefore, by increasing the temperature of the emulsion, the water droplet size that must be removed increases.

Increasing the retention time allows water droplets to grow larger. However, after a certain initial retention time, increasing the time does not significantly affect the water droplet size. For practically sized treaters, the retention time is normally kept between 10 and 30 min. The higher retention time is usually associated with heavier oils.

The viscosity of the oil is the most influencing factor on the water droplet size that can be removed. The more viscous the oil, the larger the droplet size that can be removed (settled) within a reasonable time. In a recent article by Thro and Arnold [3], the following equations relating the water droplet diameter, d_m (that must be removed to achieve 1% water content in the treated oil), to oil viscosity were developed using field data:

For $\mu_o < 80$ cP

$$d_m = 200\mu_o \ \mu\text{m} \tag{10}$$

For electrostatic treaters, $3 \text{ cP} < \mu_o < 80 \text{ cP}$

$$d_m = 170\mu_o \ \mu\text{m} \tag{11}$$

The above equations are useful in the absence of actual laboratory measurements, which are very difficult to obtain.

With the water droplet size that must be removed determined from the above equations, the dimensions of the coalescing/settling section of the treater must be sufficient to allow settling of such water droplets and also allow for the required retention time. These two conditions could be used to develop two equations that govern the dimensions of the treater as detailed in the following subsections for horizontal and vertical treaters.

5.8.1 Sizing Horizontal Treaters

Water Droplets Settling Constraint

We start with the settling equation [Eq. (3)]

$$u = 1.787 \times 10^{-6} \frac{(\Delta\gamma) d_m}{\mu_o} \ \frac{\text{ft}}{\text{s}} \tag{3}$$

This equation gives the terminal settling velocity of the water droplet. For settling to occur, the upward average velocity of the oil must not exceed the water settling velocity. The average oil velocity, u_o, is obtained by dividing the oil volumetric flow rate, Q_o, by the flow cross-sectional area, A.

Let D be the treater inside diameter in inches and L be the effective length of the settling/coalescing section in feet. Therefore,

$$u_o = \frac{Q_o(\text{bbl/day}) \times 5.61(\text{ft}^3/\text{bbl}) \times 1.1574 \times 10^{-5}(\text{day/s})}{(D/12)L}$$

$$u_o = 7.792 \times 10^{-4} \frac{Q_o}{DL} \frac{\text{ft}}{\text{s}}, \tag{12}$$

Equating u_o [Eq. (12)] to u [Eq. (3)], we obtain

$$DL = 436 \frac{Q_o \mu_o}{(\Delta\gamma) d_m} \text{ in. ft} \tag{13}$$

Retention Time Constraint

The retention time, t, can be obtained by dividing the volume of the settling/coalescing section occupied by oil, V_o, by the oil volumetric flow rate, Q_o. Assuming that the oil occupies only 75% of the coalescing/settling section,

$$t = \frac{0.75(\pi D^2 L/4 \times 144)}{5.61 Q_o \times 6.944 \times 10^{-4}(\text{day/min})} \text{ min}.$$

Solving for D^2L, we get

$$D^2 L = \frac{Q_o t}{1.05} \text{ in.}^2 \text{ ft} \tag{14}$$

Sizing Procedure

The following procedure is mostly aimed at determining the minimum size of the coalescing/settling section of the treater and the rating of the burner. Such information will be very useful in preparing equipment specifications for vendors and for evaluating the quotations received from the vendors. The vendors would provide the detailed design and dimensions of the treater.

1. The first step is to decide on a treating temperature. This is best determined from laboratory tests. The optimum treating temperature must provide a minimum loss of oil volume and quality along with a practical treater size. If laboratory data are not available, the treating temperature may be determined based on experience. In

such cases, however, the design (following steps) may be executed for different assumed treating temperature and a final decision is made based on analysis of the design results.

2. Determine the diameter of the water droplet that must be removed [from Eq. (10) or (11)].
3. Use Eq. (13) to obtain the relation between *D* and *L* that satisfies the settling constraint. Assume various values of *D* and determine the corresponding values of *L* from this relation.
4. Use Eq. (14) to obtain another relation between *D* and *L* that satisfies the retention time constraint. For the same values of *D* assumed in step 3, determine corresponding values of *L* from this relation.
5. Compare the results obtained from the above two steps and select a combination of *D* and *L* that satisfies both settling and retention time constraints.
6. Use Eq. (9) to determine the heat requirement for the selected treating temperature.

Example 1

Determine the heat requirement and the size of the settling/coalescing section of a horizontal heater treater for the following conditions:

Oil flow rate:	7000 BPD
Inlet B.S.&W.:	15%
Outlet B.S.&W.:	1%
Oil specific gravity:	0.86
Oil viscosity:	45 cP at 85°F
	20 cP at 105°F
	10 cP at 125°F
Water specific gravity:	1.06
Specific heat of oil:	0.5 Btu/lb °F
Specific heat of water:	1.1 Btu/lb °F
Inlet temperature:	85°F
Retention time:	20 min
Treating temperature:	Examine 105°F, 125°F, and no heating

Solution

Use Eq. (10) to determine the water droplet diameter for each treating temperature:

$$\text{For } T = 125^\circ\text{F}: \quad d_m = 200\mu_o = 200(10)^{0.25} = 356\,\mu\text{m}$$

For $T = 105°\text{F}$: $d_m = 200\mu_o = 200(20)^{0.25} = 423\,\mu\text{m}$

For $T = 85°\text{F}$: $d_m = 200\mu_o = 200(45)^{0.25} = 518\,\mu\text{m}$

Ignoring the effect of temperature on specific gravity, use Eq. (13) to determine the settling constraint for each treating temperature:

For $T = 125°\text{F}$

$$DL = 436\frac{Q_o\mu_o}{(\Delta\gamma)d_m} = 436\frac{7000 \times 10}{0.2 \times 356^2} = 1204 \text{ in. ft} \tag{E1}$$

For $T = 105°\text{F}$

$$DL = 436\frac{Q_o\mu_o}{(\Delta\gamma)d_m} = 436\frac{7000 \times 20}{0.2 \times 423^2} = 1706 \text{ in. ft} \tag{E2}$$

For $T = 85°\text{F}$

$$DL = 436\frac{Q_o\mu_o}{(\Delta\gamma)d_m} = 436\frac{7000 \times 45}{0.2 \times 518^2} = 2559 \text{ in. ft} \tag{E3}$$

Use Eq. (14) to determine the relationship for retention time constraints:

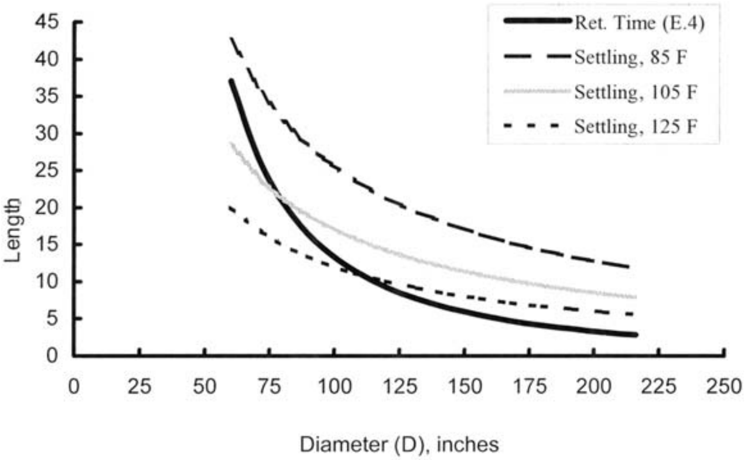
$$D^2L = \frac{Q_ot}{1.05} = \frac{7000 \times 20}{1.05} = 133{,}333 \text{ in.}^2\text{ft} \tag{E4}$$

Assume different values for D and determine the corresponding values of L from Eqs. (E1)–(E4). The results are summarized in the following table and are plotted for comparison.

D (in.)	L (ft) [Eq. (E4)]	L (ft) [Eq. (E3)]	L (ft) [Eq. (E2)]	L (ft) [Eq. (E1)]
60	37.04	**42.65**	28.43	20.07
72	25.72	**35.54**	23.69	16.72
84	18.90	**30.46**	**20.31**	14.33
96	14.47	**26.66**	**17.77**	12.54
108	11.43	**23.69**	**15.80**	11.15
120	9.26	**21.33**	**14.22**	**10.03**
132	7.65	**19.39**	**12.92**	**9.12**
144	6.43	**17.77**	**11.85**	**8.36**
156	5.48	**16.40**	**10.94**	**7.72**

Analyzing the tabulated/plotted results yields the following conclusions:

1. Any combination of D and L that exists in the plot area below the retention time curve is not acceptable.



2. For the treater diameters selected in the table, only the values of L shown in bold are acceptable, as they satisfy both settling and retention time constraints.
3. As the treating temperature increases, the size of the coalescing/settling section decreases.
4. There is no need to treat the emulsion at 125°F, as the reduction is treater size is not significant, and the increased temperature would negatively affect the volume and quality of the treated oil.
5. There is a good potential of treating this oil without any heating aid, as the treater size required seems to be practical.
6. A practical and economical selection would be an 84-in.-diameter by 21-ft-long coalescing section with a burner that can provide a treating temperature of 105°F.

Now use Eq. (9) to calculate the heat requirement, assuming 10% heat losses:

$$q = \frac{1}{1-l} 15 Q_o (\Delta T)(\gamma_o c_o + w \gamma_w c_w)$$

$$q = \frac{1}{1-0.1} 15 \times 7000(105 - 85)(0.86 \times 0.5 + 0.15 \times 1.06 \times 1.1)$$

$$= 1,411{,}433 \text{ Btu/h}$$

Therefore, a burner rated at 1.5 MM Btu/h would be a good selection.

It should be noted that after the installation of the treater in the field, the operator will run the treater at various settings while inspecting samples of the treated oil until he determines the optimum operating conditions that provide the required treatment with minimum heating. Optimization of the operating conditions is an important activity and should be conducted as often as possible to adopt with changing field conditions.

5.8.2 Sizing Vertical Treaters

Water Droplets Settling Constraint

Similar to the analysis performed for the horizontal treater, we shall start with the settling equation, [Eq. (3)]:

$$u = 1.787 \times 10^{-6} \frac{(\Delta\gamma) d_m}{\mu_o} \ \frac{\text{ft}}{\text{s}} \tag{3}$$

This equation gives the terminal settling velocity of the water droplet. For settling to occur, the upward average velocity of the oil must not exceed the water settling velocity. The average oil velocity, u_0, is obtained by dividing the oil volumetric flow rate, Q_o, by the flow cross-sectional area, A, which is the cross-sectional area of the theater: $A = (\pi/4)(D/12)^2$. Therefore,

$$u_o = \frac{Q_o(\text{bbl/day}) \times 5.61(\text{ft}^3/\text{bbl}) \times 1.1574 \times 10^{-5}(\text{day/s})}{(\pi/4)(D/12)^2}$$

$$u_o = 1.19 \times 10^{-2} \frac{Q_o}{D^2} \ \frac{\text{ft}}{\text{s}} \tag{15}$$

Applying equation u_o [Eq. (15)] to u [Eq. (3)], we obtain

$$D^2 = 6665 \frac{Q_o \mu_o}{(\Delta\gamma) d_m^2} \ \text{in.ft} \tag{16}$$

Retention Time Constraint

The reaction time, t, can be obtained by dividing the volume of the settling/coalescing section occupied by oil, V_o, by the oil volumetric flow rate, Q_o. Let H be the height of the coalescing/settling section (in in.); then,

$$t = \frac{(\pi D^2/4 \times 144)(H/12)}{5.61 Q_o \times 6.944 \times 10^{-4}(\text{day/min})} \ \text{min.}$$

Solving for D^2L, we get

$$D^2 H = 8.575 Q_o t \ \text{in.}^2\text{ft} \tag{17}$$

Equations (16) and (17) can be used for gunbarrel tanks as well. However, when the diameter of the tank is larger than 48 in., the equations must be multiplied by a factor (greater than 1.0) to account for short-circuiting.

Sizing Procedure

Similar to horizontal treaters, the following producer is primarily aimed at determining the minimum size of the coalescing/settling section of the treater and the rating of the burner.

1. Determine the optimum treating temperature that provides the minimum lose of oil volume and quality along with a practical treater size. If this is not available, the design (following steps) may be executed for different assumed treating temperature and a final decision is made based on analysis of the design results.
2. Determine the diameter of the water droplet that must be removed [from Eq. (10) or (11)].
3. Use Eq. (16) to obtain the minimum treater diameter *D* that satisfied the settling constraint.
4. Repeat the above steps for different assumed treating temperatures and determine the values of *D* for each treating temperature.
5. Use Eq. (17) to obtain a relation between *D* and *H* that satisfies the retention time constraint. Then, assume different values of *D* and determine corresponding value of *H* from this relation.
6. Analyze the results to determine the combinations of *D* and *H*, for each treating temperature, that satisfy both settling and retention time constraints.
7. Use Eq. (9) to determine the heat requirement for the selected treating temperature.

Example 2

Determine the heat requirement and the size of the settling/coalescing section of a single-well vertical heater treater for the same conditions of Example 1 given that the well flow rate is 1200 BPD.

Solution

Use Eq. (16) to determine the minimum diameter at the three treating temperatures:

For $T = 85°\mathrm{F}$

$$D^2 = 6665 \frac{Q_o \mu_o}{(\Delta\gamma) d_m} = 6665 \frac{1200 \times 45}{0.2(518)^2} = 6707 \text{ in.}^2$$

$$D_{85} = 81.89 \text{ in.}$$

For $T = 105°\mathrm{F}$

$$D^2 = 6665\frac{Q_o\mu_o}{(\Delta\gamma)d_m} = 6665\frac{1200 \times 20}{0.2(423)^2} = 4476 \text{ in.}^2$$

$$D_{105} = 66.86 \textit{ in.}$$

For $T = 85°\mathrm{F}$

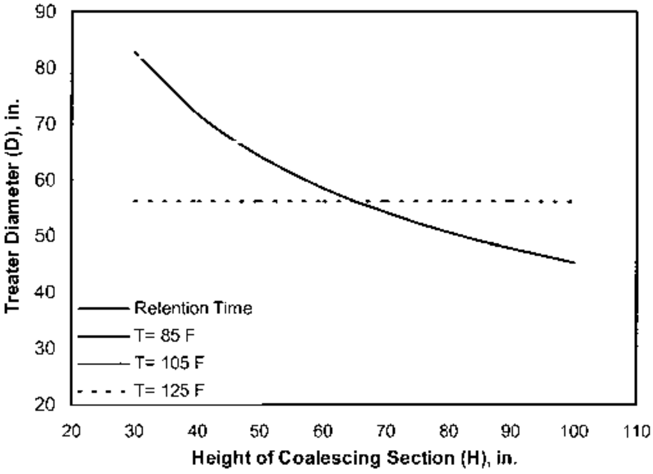
$$D^2 = 6665\frac{Q_o\mu_o}{(\Delta\gamma)d_m} = 6665\frac{1200 \times 10}{0.2(356)^2} = 3155 \text{ in.}^2$$

$$D_{125} = 56.17 \text{ in.}$$

Now, use Eq. (17) for the retention time constraint:

$$D^2H = 8.575Q_ot = 8.575 \times 1200 \times 20 = 205{,}800 \text{ in.}^3$$

Assume different values for D and determine corresponding values of H from the above relation. The results are plotted as follows:



From the figure, all diameters and heights that fall below the retention time curve are not acceptable. For the three treating temperatures, a coalescing section height equal to the value at the intersection with the retention time curve, or larger, will satisfy both retention time and settling constraints.

A reasonable selection will be a treater with a 66-in. diameter and a 60-in. coalescing section height. The treating temperature will be 105°F, with the possibility of treatment being at lower temperatures in reality.

The burner rating in determined from Eq. (9):

$$q = \frac{1}{1-l} 15 Q_o (\Delta T)(\gamma_o c_o + w \gamma_w c_w)$$

$$q = \frac{1}{1-0.1} 15 \times 1200(105-85)(0.86 \times 0.5 + 0.15 \times 1.06 \times 1.1)$$

$$= 241{,}960 \ \frac{\text{Btu}}{\text{h}}$$

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REVIEW QUESTIONS AND EXERCISE PROBLEMS

1. Check the correct answers for the following:

    (i) A normal goal in dehydration of crude oil is to reduce the water content to below:

    - (a) 0.5% by volume
    - (b) 5% by volume
    - (c) 10% by volume
    - (d) None of the above

    (ii) The objective of oil dehydration is to

    - (a) Separate free water only

(b) Reduce the emulsified water in the oil

(c) Separate free water (if found) and reduce the emulsified water

(iii) Conditions leading to stable water–oil emulsions during oil production are

(a) Heat and energy of agitation
(b) Heat, energy of agitation, and emulsifying agents
(c) Energy of agitation and presence of emulsifying agents

(iv) An emulsion can be broken by applying

(a) Heat and settling time
(b) Heat and chemicals
(c) Pressure and settling time
(d) Chemicals and settling time

(v) Heat consumed to raise a given weight of salt water to a certain temperature compared to that consumed for the same weight of crude oil is

(a) Twice as much
(b) One-half as much
(c) The same amount of heat

(vi) Free water found in crude oil is considered more corrosive than an emulsion because

(a) Emulsions are in direct contact with the steel
(b) Salt water comes is direct contact with the steel
(c) Only oil in emulsion contacts the steel

(vii) Gas separation—when oil is heated—is necessary because

(a) It causes a build-up of pressure in the treater
(b) Separated hot gas can sell for a value higher than oil
(c) It can cause an interference with water settling from emulsion

2. Match the items in column A with the correct statements given in column B.

Column A	Column B
Set #1:	
1. Water–oil emulsion	a. Less than 1% B.S.&W.
2. To break emulsion	b. Water droplets in oil
3. Heating emulsion	c. Destroy film surrounding water drops

4. Pipeline quality crude	d. Reduce oil viscosity

Set#2:

1. Primary gas	a. Reduces volume of oil
2. Free water	b. Wellhead gas separator
3. Secondary gas	c. Knockout drum
4. Emulsion heater	d. Forms in treating equipment
5. Chemical selection	e. Emulsion testing

Set #3:

1. Primary gas	a. Gas, water, emulsion
2. Secondary gas	b. Free water knockout
3. Three-phase separator	c. Release gas
4. Reduce oil pressure	d. Wellhead GOSP

2. Several factors affect the oil dehydration process. For the following factors, indicate the desirable level or magnitude that would accomplish easier or faster separation of water from oil (check one for each factor).

	Desirable level magnitude		
Factor	High/large	Low/small	Insignificant
(a) Viscosity of oil			
(b) Density difference			
(c) Size of dispersed water droplets			
(d) Water/oil volume ratio			
(e) Salinity of water			

3. Discuss briefly the effects of heating, deemulsifiers, and electrostatic field on the oil dehydration process.
4. Discuss and compare the effects of temperature, oil viscosity, and retention time on the water droplet size that can be removed from the oil to achieve certain dehydration goal.
5. Four thousand barrels per day of 20°API oil containing 15% by volume emulsified water of specific gravity 1.05 need to be dehydrated. A horizontal heater treater has been selected for this operation. The fluid entering the heater at 90°F must be heated up to 160°F and retained in the coalescing/settling section for 25 min.

(a) Determine the required diameter and effective length of the coalescing/settling section of the treater.

(b) Given that the specific heats of oil and water are 0.5 and 1.0 BTU/lb°F, respectively, determine the required heat duty of the heater in MMBTU/h assuming 15% heat loss to the surroundings.

(c) If the heater fire tube is made of 6 in. pipe, determine the length of the fire tube assuming a mean temperature difference of 130°F and an overall heat transfer coefficient of 2000 BTU/h ft^{2}°F.

(d) What would you suggest to reduce the retention time.

6. Determine the diameter and height of the coalescing section of a vertical heater treater and the required heat duty, given the following data:

Oil rate: 10,000 BPD (specific gravity = 0.875, $C_p = 0.5$ BTU/lb°F)
Water content: 10% by volume (specific gravity = 1.04, $C_p = 1.1$ BTU/lb°F)
Intel temperature: 90°F
Treating temperature: 160°F
Retention time: 20 min

6

Desalting of Crude Oil

6.1 INTRODUCTION

The removal of salt from crude oil for refinery feed stocks has been and still is a mandatory step. This is particularly true if the salt content exceeds 20 PTB (pounds of salt, expressed as equivalent sodium chloride, per thousand barrels of oil).

The most economical place for the desalting process is usually in the refinery. However, when marketing or pipeline requirements are imposed, field plants are needed for processing the salty oil prior to shipping. The principles involved are the same whether desalting takes place at the refinery or in the field. Salt in crude oil is, in most cases, found dissolved in the remnant brine within the oil.

The remnant brine is that part of the salty water that cannot be further reduced by any of the dehydration methods described in the previous chapter. It is commonly reported as basic sediments and water (B.S.&W.). It is understood that this remnant water exists in the crude oil as a dispersion of very fine droplets highly emulsified in the bulk of oil. The mineral salts of this brine consist mainly of chlorides of sodium, calcium and magnesium. A summary of the properties of crude oil as received at the refinery is given in Table 1. Nelson [1] compiled the data given in Table 2 on the amount of salts found in oils for various regions in the world.

The amount of salt in the crude oil is a function of the amount of the brine that remains in the oil W_R (% B.S.&W.) and of its salinity S_R in parts per million (ppm). In other words, this relationship could be written in the following functional form (after Manning and Thompson [2]):

$$\mathrm{PTB} = 350\gamma_{\mathrm{Brine}}\left(\frac{1000W_R}{100 - W_R}\right)\left(\frac{S_R}{10^6}\right) \tag{1}$$

Table 1 Properties of Crude Oils Shipped to Refineries

	Range	Average
Water in crude, % by volume of crude	0.1–2.0	0.3–0.5
Salt content in crude, PTB	10–250	60–130
Salt concentration in brine, wt%	0.4–25	—
Salt concentration in brine, ppm	4,000–250,000	—

Source: Ref. 1.

Table 2 Average Values for the PTB for Some Typical Crude Oils

Source of oil	Avg. salt content (PTB)
Middle East	8
Venezuela	11
United States	
Pennsylvania	1
Wyoming	5
East Texas	28
Gulf Coast	35
Oklahoma and Kansas	78
West Texas	261
Canada	200

The method of reducing the PTB by lowering the quantity of remnant water W_R is usually referred to as the treating process of oil dehydration. This was the main theme of the last chapter. The other alternative of reducing the PTB is to substantially decrease the dissolved salt content of the remnant water (i.e., its concentration, S_R). This practice is the one we are dealing with in this chapter and is known as desalting. Desalting of crude oil will eliminate or minimize problems resulting from the presence of mineral salts in crude oil. These salts often deposit chlorides on the heat transfer equipment of the distillation units and cause fouling effects. In addition, some chlorides will decompose under high temperature, forming corrosive hydrochloric acid [3]:

$$\mathrm{MgCl_2(aq.)} \xrightarrow[\mathrm{H_20}]{\text{High temp.}} \mathrm{Mg(OH)_2} + \mathrm{HCl} \tag{2}$$

The removal of these salts is aimed at providing an economical operating cycle in the refining process of crude oil. The reduction of salt content down to 5 PTB is feasible. Even with this low salt content, it has been reported that the processing of 25,000 bbl/day of crude oil could result in an amount of HCl equal to 65 lb/day [4].

6.2 RELATIONSHIP AMONG VOLUME OF REMNANT WATER, ITS SALINITY AND SALT CONTENT

Taking one-tenth of 1% by volume of water to remain in the crude as a basis for our calculation, the relationship between the salt content of the remnant water expressed in pounds of salt per thousand barrels of oil (PTB) and its concentration or salinity expressed in ppm is presented graphically in Figure 1. For other volumes, simple multiples of the numbers given by this graph are used. The following example illustrates the use of this relationship.

Example 1

Find the PTB of a crude oil having 10% by volume remnant water if its concentration is estimated to be 40,000 ppm at 25°C.

Solution

The example is solved using two approaches: One is based on Figure 1 and the other approach utilizes basic calculations.

1. Using Figure 1, the PTB of crude oil having 0.1% remnant water with 40,000 ppm salinity is found to be 14 PTB. For crude oil containing 10% remnant water, the value of PTB obtained from the figure should be multiplied by 100; therefore, the given crude contains 1400 PTB.
2. Take a basis of 1000 bbl of wet oil; the B.S.&W. = 10%, and the saline water concentration = 40,000 ppm = 4%. Then,

$$\text{Quantity of water in oil} = (1000)0.1$$
$$= (100\ \text{bbl})(5.6\text{ft}^3/\text{bbl}) = 560\ \text{ft}^3$$

Now, the density of the saline water is estimated using Table 3. For 4% concentration and at 25°C, the density is 1.0253 g/cm^3, or 63.3787 lb/ft^3. Hence,

$$\text{Mass of water} = (560\ \text{ft}^3)63.3787\ \text{lb/ft}^3$$
$$= 35{,}828\ \text{lb}$$

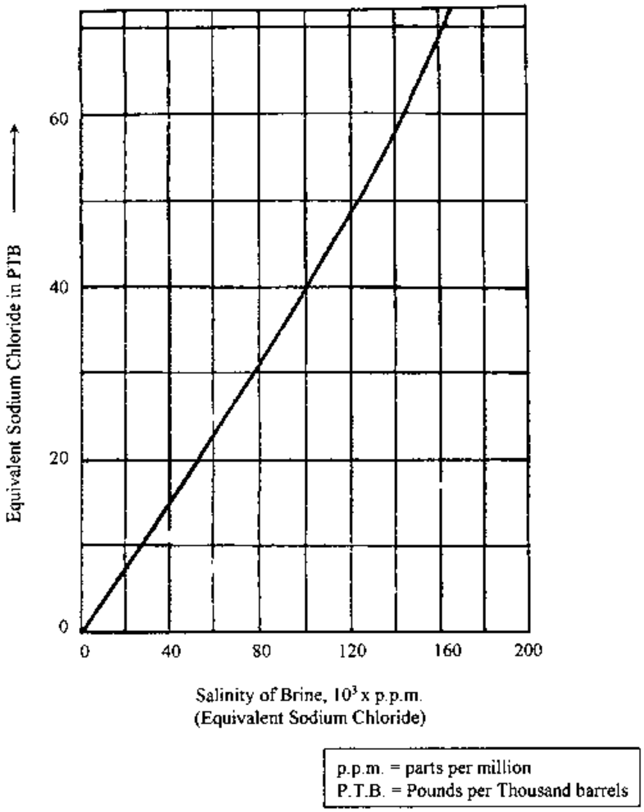

Figure 1 Salt content of crude oil (PTB) as a function of salinity of its remnant – water 0.1% (1/1000) by volume remnant water.

Table 3 Densities of Aqueous Inorganic Solutions [Sodium Chloride (NaCl)]

%	0°C	10°C	25°C	40°C	60°C	80°C	100°C
1	1.00747	1.00707	1.00409	0.99908	0.9900	0.9785	0.9651
2	1.01509	1.01442	1.01112	1.00593	0.9967	0.9852	0.9719
4	1.03038	1.02920	1.02530	1.01977	1.0103	0.9988	0.9855
8	1.06121	1.05907	1.05412	1.04798	1.0381	1.0264	1.0134
12	1.09244	1.08946	1.08365	1.07699	1.0667	1.0549	1.0420
16	1.12419	1.12056	1.11401	1.10688	1.0962	1.0842	1.0713
20	1.15663	1.15254	1.14533	1.13774	1.1268	1.1146	1.1017
24	1.18999	1.18557	1.17776	1.16971	1.1584	1.1463	1.1331
26	1.20709	1.20254	1.19443	1.18614	1.1747	1.1626	1.1492

The quantity of NaCl salt found in this mass of water is $(35{,}828)(40{,}000)/10^6 = 1433$ lb. Since our basis is 1000 bbl of oil, the salt content is 1433 PTB.

3. Using Eq. (1) we get

$$\text{PTB} = 350\gamma_{\text{Brine}} \frac{1000 W_R}{100 - W_R}\left(\frac{S_R}{10^6}\right)$$

$$\text{PTB} = (350)\,(1.0253)\frac{1000(10)}{100 - 10}\left(\frac{40000}{10^6}\right)$$

$$\text{PTB} = 1595$$

Example 2

Rework Example 2 of Chapter 5, assuming that the water salinity is 40,000 ppm instead of 20,000 ppm and dehydration is done to two-tenths of 1%. Calculate the PTB of the oil.

Solution

With reference to Figure 1, the PTB corresponding to one-tenth of 1% is found to be 14. Hence, the PTB of two-tenths of 1% is $(14)(0.2/0.1) = 28$. This salt content is above the maximum limit of 20 PTB. Therefore, desalting is recommended as a second stage following dehydration in order to reduce the PTB.

6.3 DESCRIPTION OF THE DESALTING PROCESS

It is clear from Example 2 that we cannot economically achieve a satisfactory salt content in oil by using dehydration only (single stage). This is particularly true if the salinity of the water produced with oil is much greater than 20,000 ppm (formation water has a concentration of 50,000–250,000 mg/L). Accordingly, a two-stage system (a dehydration stage and a desalting stage) as shown in Figure 2a. Under certain conditions, however, a three-stage system may be used which consists of a dehydration stage and two consecutive desalting units as shown in Figure 2b.

As shown in Figure 2, *wash water*, also called *dilution water*, is mixed with the crude oil coming from the dehydration stage. The wash water, which could be either fresh water, or water with lower salinity than the remnant water, mixes with the remnant water, thus diluting its salt concentration. The mixing results in the formation of water–oil emulsion.

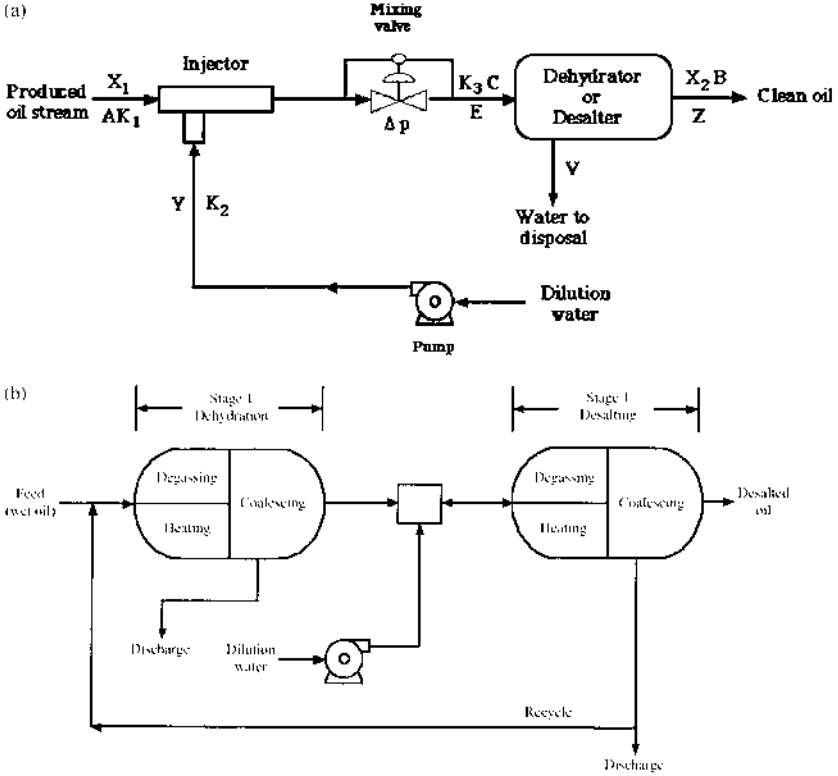

Figure 2 (a) Single-stage desalting system (from Ref. 5). (b) A two-stage desalting system.

The oil (and emulsion) is then dehydrated in a manner similar to that described in Chapter 5. The separated water is disposed of through the field-produced water treatment and disposal system. In the two-stage desalting system, dilution water is added in the second stage and all, or part, of the disposed water in the second stage is recycled and used as the dilution water for the first desalting stage. Two-stage desalting systems are normally used to minimize the wash water requirements.

The mixing step in the desalting of crude oil is normally accomplished by pumping the crude oil (which is the continuous phase) and wash water (which is the dispersed phase) separately through a mixing device. The usual mixing device is simply a throttling valve. The degree of mixing can be enhanced if the interfacial area generated upon mixing is increased. A useful device for such a purpose is the application of multiple-orifice-plate mixers (MOMs) shown in Figure 3. It is of

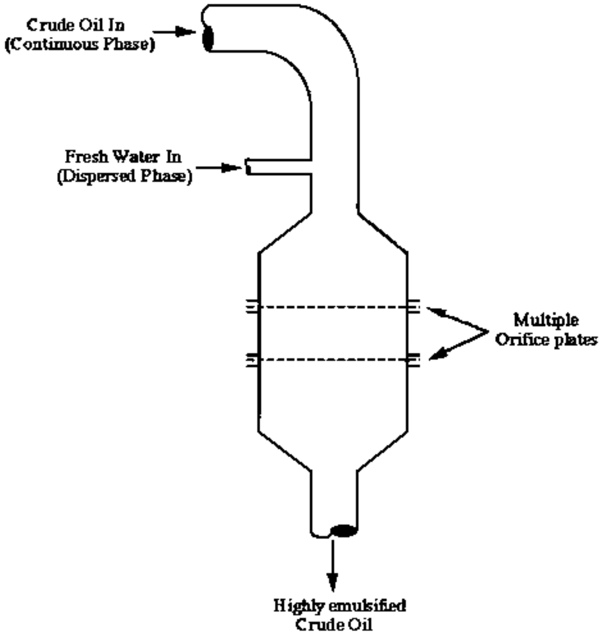

Figure 3 Details of multiple-orifice plate mixers (MOM).

importance to point out that although the theory of dilution of remnant water with fresh water is sound in principle, it can become impossible to implement in actual application. It all depends on the intimate mixing of remnant water with dilution water.

In the emulsion-treating step, a heating, chemical, or electrical demulsifying aid (or a combination of them) is commonly used. The chemical desalting process involves adding chemical agents and wash water to the preheated oil, followed by settling, as shown in Figure 4. The settling time varies from a few minutes to 2 h. Some of the commonly used chemical agents are sulfonates, long-chain alcohols, and fatty acids.

6.4 ELECTROSTATIC DESALTING

In this case, an external electric field is applied to coalesce the small water droplets and thus promote settling of the water droplets out of the oil. The electric field may be applied in any of the following manners:

1. **ac field devices for water-rich emulsions**. Alternating current (ac) is applied, which alternates the polar water molecule arrangements

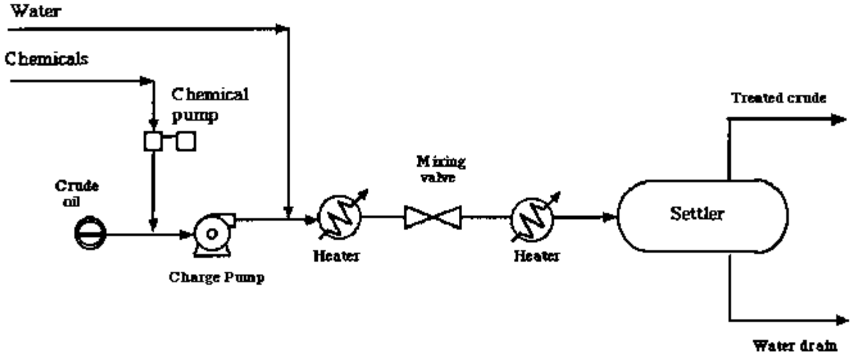

Figure 4 Chemical desalting.

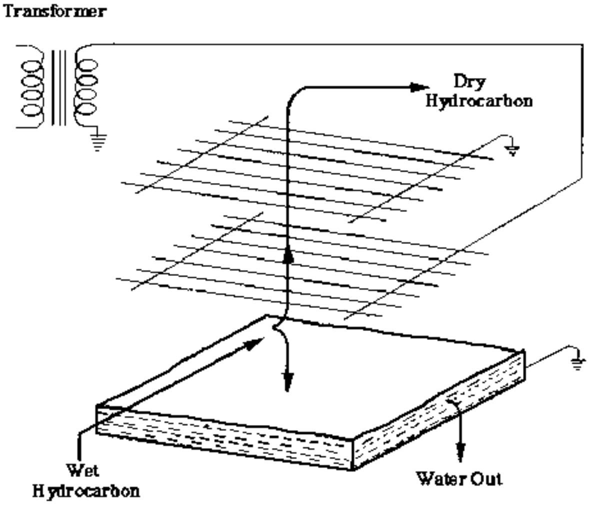

Figure 5 ac electrostatic coalescer.

leading to better coalescence. A schematic diagram of ac electrostatic coalescence is shown in Figure 5.

2. **ac/dc field for maximum dehydration**. A combination of ac and dc (direct current) is used in this case. The basic configuration of this process is shown in Figure 6. The ac is produced in the zone beneath the electrodes, whereas the dc field is produced between adjacent electrodes. This arrangement achieves maximum water removal.

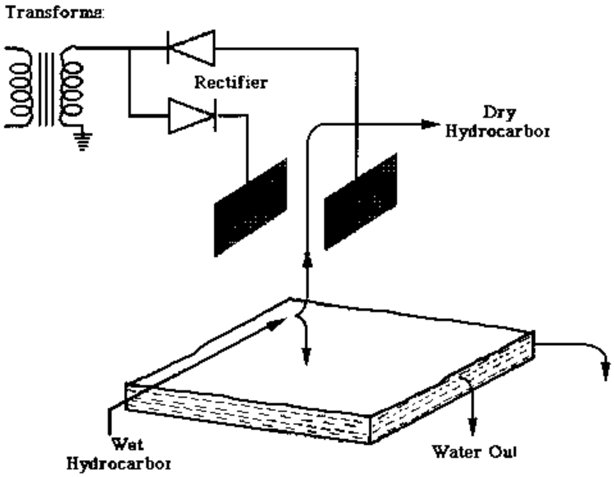

Figure 6 Dual polarity ac/dc field.

3. **Variable gradient field for maximum salt reduction**. If the field gradient is increased beyond a certain limit (E_c), this will shatter the drops; it is expressed as

$$E_c \leq k\left(\frac{\gamma}{d}\right)^{1/2}$$

where k is the dielectric constant, γ is the interfacial tension, and d is the drop diameter. Thus, the drop size can be controlled by the field gradient. The electric field can be used both to mix and separate the drops.

By cycling the field strength, the process can be repeated many times during the retention time of the drops within the electric field. Voltage modulation will create mixing, coalescence, and settling, as shown in Figure 7. The electrostatic field causes the dipolar water molecules to align with the field.

In the electrical desalting process, a high potential field (16,500–33,000 V) is applied across the settling vessels to help coalescence, as shown in Figure 8. The chemielectric concept utilizing both chemical agents and electrical field is schematically illustrated in Figure 9.

In the desalting process, it is a common practice to apply enough pressure to suppress any loss of hydrocarbon due to vaporization of the oil. The pressure normally used in a desalting process is in the range 50–250 psi.

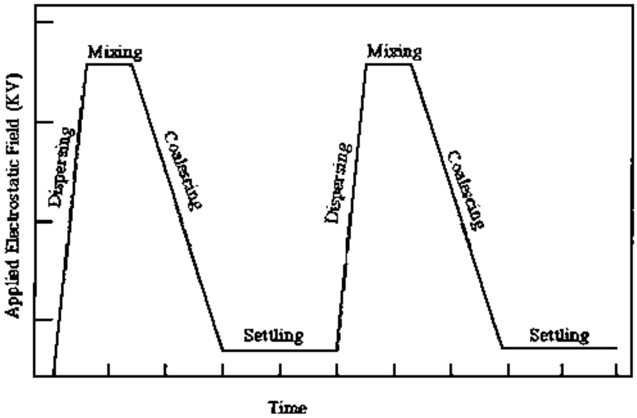

Figure 7 Voltage modulation for electrostatic mixing coalescing.

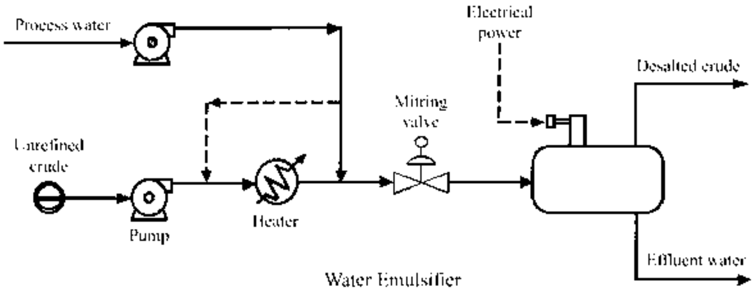

Figure 8 Electrical desalting.

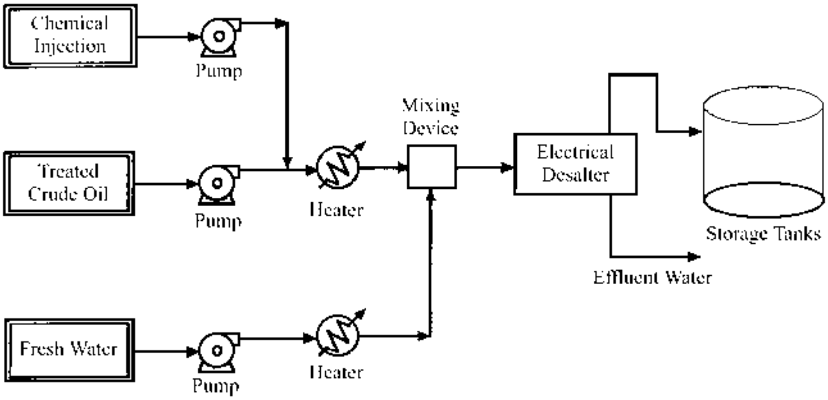

Figure 9 A typical desalting system utilizing chemelectric approach.

6.5 DETERMINING DILUTION WATER REQUIREMENT

From the operational point of view, it has been reported that the amount of water of dilution W_D added in the desalting of crude oils is in the range 5–10% by volume, based on the amount of remnant water and its salinity.

The following correlation has been developed [3] for determining W_D as a function of W_R, S_R, S_D, (salinity of the dilution water) and the efficiency of mixing between the two phases, E.

$$W_D = \frac{2.5 \times 10^3 (W_R)^{0.01533}}{(S_D)^{0.2606} (S_R)^{0.0758} E^{0.6305}} \tag{3}$$

where S_D and S_R are in parts per million. On the other hand, the following analytical relationship could set the acceptable limits on the salt content in crude oil. A component material balance for the salt gives

$$EW_D S_D + W_R S_R = S_B (W_D + W_R)$$

or

$$S_B = \frac{EW_D S_D + W_R S_R}{W_D + W_R} \tag{4}$$

where S_B refers to some average salinity in the bulk of the homogeneous phase as a result of mixing the remnant water with the fresh water.

6.6 EFFECT OF OPERATING PARAMETERS

The efficiency of desalting is dependent on the following parameters [6,7]:

1. **Water–crude interface level**. This level should be kept constant; any changes will change electrical field and perturbs electrical coalescence.
2. **Desalting temperature**. Temperature affects water droplet settling through its effect on oil viscosity; therefore, heavier crude oils require higher desalting temperatures.
3. **Wash water ratio**. Heavy crudes require a high wash water ratio to increase electrical coalescence. A high wash ratio acts similarly to raise temperatures, as illustrated in Table 4.
4. **Pressure drop in the mixing valve**. A high-pressure-drop operation results in the formation of a fine stable emulsion and better washing. However, if the pressure drop is excessive, the emulsion might be difficult to break. The optimum pressure drop is 1.5 bar for light crudes and 0.5 bar for heavy crudes.

Table 4 Average Desalting Conditions

Crude gravity (°API)	Desalting temperature (°C)	Minimum water ratio (vol%)
> 40	110	2–4
30–40	110	4–8
	120	4–7
< 30	130	8–10
	140	>10

Table 5 Problems, Causes, and Solutions

Problems	Causes	Solutions
1. A high salt content in the desalted crude oil	(a) Feed salt content high (b) Wash water injection low (c) Crude oil flow rate exceeds the design flow rate (d) Insufficient mixing of the crude oil and wash water	(a) Increase the wash water rate (b) Reduce the crude oil flow rate (c) Increase the mix value pressure drop
2. Oil in the desalter effluent water	(a) "Interface" level too low (b) Wide emulsion band at the "interface" (c) Excessive crude oil wash water mixing (d) Poor wash water quality (e) Crude temperature too low	(a) Increase the interface level (b) Inject a chemical or dump the emulsion (c) Reduce the mix valve pressure drop (d) Check for any waste in the wash water source
3. High water carry over in desalted crude oil	(a) Wash water flow rate too high (b) Excessive formation water in the crude oil (c) Interface level too high (d) Disturbance in the desalter vessel	(a) Reduce the wash water flow rate and commence or increase chemical injection (b) Reduce the interface level and check the effluent water valve (c) Check for an excessive cause and allow the unit to settle down

5. **Type of demulsifiers**. Demulsifiers are added to aid in complete electrostatic coalescence and desalting. They are quite important when heavy crudes are handled. Levels ranging between 3 and 10 ppm of the crude are used.

6.7 DESIGN CONSIDERATION

The following major parameters are considered when designing the desalting system:

1. Flow scheme arrangements (conventional one-stage or countercurrent contact desalters)
2. Number of desalting stages
3. Dehydration levels achieved
4. Salinity of the brine in the crude
5. Efficiency of valve mixing
6. Salinity of dilution water
7. Target PTB specification

6.8 TROUBLE SHOOTING

Table 5 lists some “tips” that are helpful in solving some of the operating problems or troubles that are of significance to the desalting process.

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REVIEW QUESTIONS AND EXERCISE PROBLEMS

1. Crude oil needs to be desalted for the following two main reasons:

   (a)
   (b)

2. Crude oil should be desalted if its salt content exceeds __________ PBT.
3. Define the following terms:

   (i) Remnant water:
   (ii) Dilution water:
   (iii) BS&W:

4. What is the main advantage of a two-stage desalting process over a single-stage desalting process?
5. Explain why desalting takes place after and not during the dehydration (emulsion treatment) process.
6. Sketch and briefly describe a two-stage desalting process.
7. An oil field produces 200,000 bbl/day of net oil with a salt content of 10 PTB. The oil out of the emulsion treater contains 0.3% water (salinity = 250,000 ppm and specific gravity = 1.07). In the desalting–dehydration process, the oil is mixed with dilution water (3000 ppm salinity and 1.02 specific gravity) and dehydrated down to 0.1% BS&W. Assuming that the mixing efficiency is 80%, determine the following:

   (a) The salt content in PTB of the oil out of the emulsion treater
   (b) The amount of dilution water required (in bbl/day)
   (c) The amount of disposed water (in bbl/day) and its salinity (in ppm)

8. A dehydration–desalting system consists of the following:

   (a) A first-stage dehydration only (no wash water added) which receives the fed containing 10% by volume water having a salinity of 250,000 ppm and a specific gravity of 1.06. The dehydration is carried out to produce oil with 0.3% water.
   (b) A second-stage desalting, which receives the effluent of the first stage and mix it with 10% by volume of wash water containing 3000 ppm salt. Then, dehydration is carried out to produce oil with 0.2% BS&W.

   Determine the salt content in PTB of the following:

   (i) The oil entering the first stage

(ii) The oil leaving the first stage
(iii) The oil leaving the second stage

9. A two-stage desalting unit is used to treat 100,000 BPD of net oil. The oil entering the first stage contains 3% water having 280,000 ppm salt and 1.07 specific gravity. The oil is mixed with recycled water from the second stage (mixing efficiency = 80%) and dehydrated down to 1% water content. In the second stage, the oil is mixed with 7000 bbl of water having 3000 ppm salt and a specific gravity of 1.02 (mixing efficiency = 80%) and is then dehydrated down to 0.1% water content and salt content of 10 PTB.

   (a) Draw a flow diagram of the described process.
   (b) Determine the salinity [in lb of salt/bbl of water ($\mathrm{lb_s/bbl_w}$)] at all points.
   (c) Determine the salt content of the oil (in PTB) at all points.
   (d) Determine the amounts or recycled water and disposed water for the two stages.
   (e) If only a one-stage desalting unit was used, determine the amount of dilution water needed to produce the same final oil quality as for the two-stage unit.

7

Crude Oil Stabilization and Sweetening

7.1 INTRODUCTION

Once degassed and dehydrated–desalted, crude oil is pumped to gathering facilities to be stored in storage tanks. However, if there are any dissolved gases that belong to the light or the intermediate hydrocarbon groups (as was explained in Chap. 3), it will be necessary to remove these gases along with hydrogen sulfide (if present in the crude) before oil can be stored. This process is described as a “dual process” of both stabilizing and sweetening a crude oil.

In stabilization, adjusting the pentanes and lighter fractions retained in the stock tank liquid can change the crude oil gravity. The economic value of the crude oil is accordingly influenced by stabilization. First, liquids can be stored and transported to the market more profitably than gas. Second, it is advantageous to minimize gas losses from light crude oil when stored.

This chapter deals with methods for stabilizing the crude oil to maximize the volume of production as well as its API gravity, against two important constraints imposed by its vapor pressure and the allowable hydrogen sulfide content.

To illustrate the impact of stabilization and sweetening on the quality of crude oil, the properties of oil before and after treatment are compared as follows:

(a) Before treatment

Water content: up to 3% of crude in the form of emulsions and from 3% to 30% of crude as free water
Salt content: 50,000–250,000 mg/L formation water

Gas: dissolved gases in varying amounts depending on the gas–oil ratio (GOR)
Hydrogen Sulfide: up to 1000 ppm by weight

(b) After treatment (dual-purpose operation): Sour wet crude must be treated to make it safe and environmentally acceptable for storage, processing, and export. Therefore, removing water and salt, as in Chapters 5 and 6, respectively, is mandatory to avoid corrosion; separation of gases and H_2S will make crude oil safe and environmentally acceptable to handle.

Water content (B.S.&W.): 0.3% by volume, maximum
Salt content: 10–20 lbs salt (NaCl) per 1000 barrels oil (PTB)
Vapor pressure: 5–20 psia RVP (Reid vapor pressure)
H_2S: 10–100 ppmw

Crude oil is considered "sweet" if the dangerous acidic gases are removed from it. On the other hand, it is classified as "sour" if it contains as much as 0.05 ft^3 of dissolved H_2S in 100 gal of oil. Hydrogen sulfide gas is a poison hazard because 0.1% in air is toxically fatal in 30 min. Additional processing is mandatory—via this dual operation—in order to release any residual associated gases along with H_2S present in the crude. Prior to stabilization, crude oil is usually directed to a spheroid for storage in order to reduce its pressure to very near atmospheric, as shown in Figure 1.

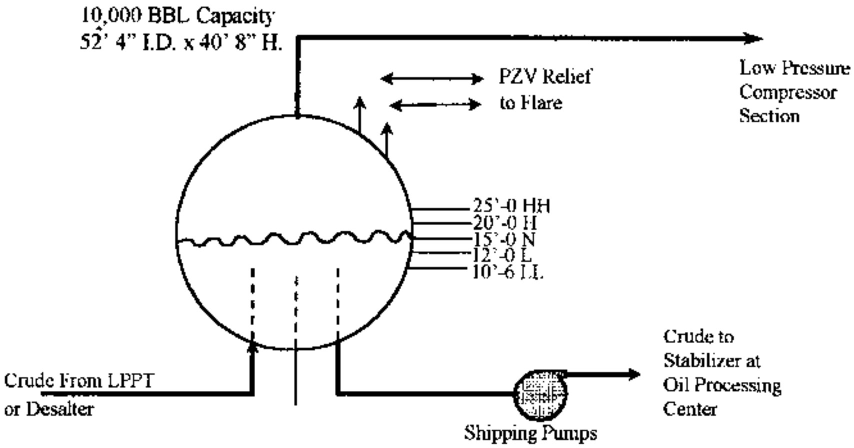

Figure 1 Typical spheroid for oil storage prior to stabilization.

7.2 STABILIZATION OPERATIONS

As was presented in Chapter 3, the traditional process for separating the crude oil–gas mixture to recover oil consists of a series of flash vessels [gas–oil separation plant (GOSP)] operating over a pressure range from roughly wellhead pressure to nearly atmospheric pressure. The crude oil discharged from the last stage in a GOSP or the desalter has a vapor pressure equal to the total pressure in the last stage. Usually, operation of this system could lead to a crude product with a RVP in the range of 4 to 12 psia. Most of the partial pressure of a crude comes from the low-boiling compounds, which might be present only in small quantities—in particular hydrogen sulfide and low-molecular-weight hydrocarbons such as methane and ethane.

Now, stabilization is directed to remove these low-boiling compounds without losing the more valuable components. This is particularly true for hydrocarbons lost due to vent losses during storage. In addition, high vapor pressure exerted by low-boiling-point hydrocarbons imposes a safety hazard. Gases evolved from an unstable crude are heavier than air and difficult to disperse with a greater risk of explosion.

The stabilization mechanism is based on removing the more volatile components by (a) flashing using stage separation and (b) stripping operations.

As stated earlier, the two major specifications set for stabilized oil are as follows:

- The Reid vapor pressure (RVP)
- Hydrogen sulfide content

Based on these specifications, different cases are encountered:

Case 1: Sweet oil (no hydrogen sulfide); no stabilization is needed. For this case and assuming that there is a gasoline plant existing in the facilities (i.e., a plant designed to recover pentane plus), stabilization could be eliminated, allowing the stock tank vapors to be collected [via the vapor recovery unit (VRU)] and sent directly to the gasoline plant, as shown in Figure 2.

Case 2: Sour crude; stabilization is a must. For this case, it is assumed that the field facilities do not include a gasoline plant. Stabilization of the crude oil could be carried out using one of the approaches outlined in Figure 3. Basically, either flashing or stripping stabilization is used.

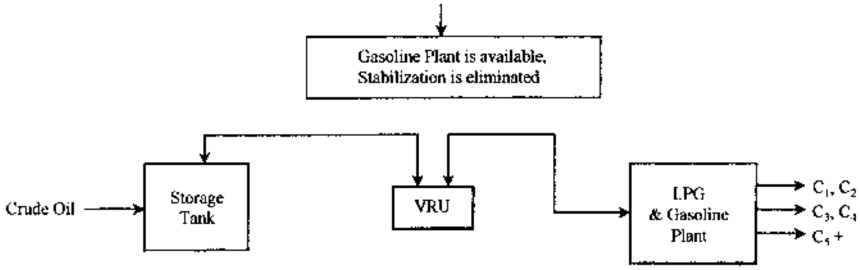

Figure 2 Field operation with no stabilization.

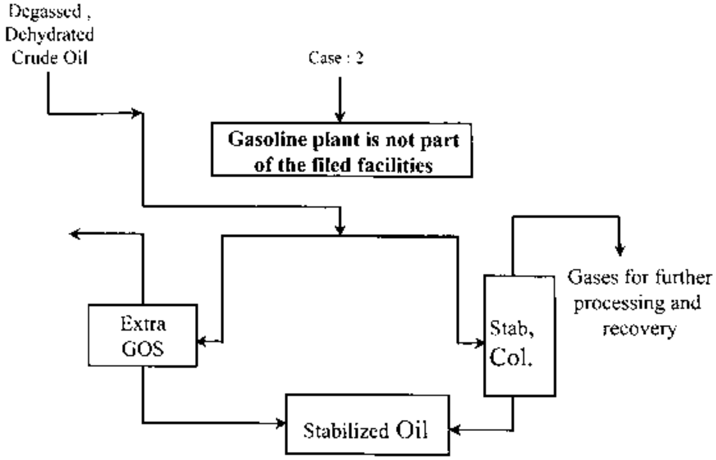

Figure 3 Alternatives for stabilizing crude oil.

It can be concluded from the above that the hydrogen sulfide content in the well stream can have a bearing effect on the method of stabilization. Therefore, the recovery of liquid hydrocarbon can be reduced when the stripping requirement to meet the H_2S specifications is more stringent than that to meet the RVP specified. Accordingly, for a given production facility, product specifications must be individually determined for maximum economic return on any investment.

7.2.1 Stabilization by Flashing (Additional Gas–Oil Separator)

The method utilizes an inexpensive small vessel to be located above the storage tank. The vessel is operated at atmospheric pressure. Vapors separated from the separator are collected using a VRU. This approach is recommended for small-size oil leases handling small volume of fluids to be processed. The principles underlying the stabilization process are the same as for gas–oil separation covered in Chapter 3.

7.2.2 Stabilization by Stripping

The stripping operation employs a stripping agent, which could be either energy or mass, to drive the undesirable components (low-boiling-point hydrocarbons and hydrogen sulfide gas) out of the bulk of crude oil. This approach is economically justified when handling large quantities of fluid and in the absence of a VRU. It is also recommended for dual-purpose operations for stabilizing sour crude oil, where stripping gas is used for stabilization. Stabilizer-column installations are used for the stripping operations.

7.3 TYPES OF STABILIZER EMPLOYING ENERGY AS A STRIPPING AGENT

Two basic types of trayed stabilizer are commonly used:

- Conventional reflux types normally operate from 150 to 300 psia. This type of stabilizer is not common in field installations. It is more suitable for large central field processing plants.
- Nonrefluxed stabilizers generally operate between 55 and 85 psia. These are known as "cold feed" stabilizers. They have some limitations, but they are commonly used in field installations because of their simplicity in design and operation.

7.4 NONREFLUXED STABILIZERS

7.4.1 Introduction

When hydrocarbon liquids are removed from the separators, the liquid is at its vapor pressure or bubble point. With each subsequent pressure reduction, additional vapors are liberated. Therefore, if the liquids were removed directly from a high-pressure separator into a storage tank, vapors generated would cause loss of lighter as well as heavier ones. This explains the need for many stages in a GOSP. Nevertheless, regardless of the number of stages used, some valuable hydrocarbons are lost with the overhead vapor leaving the last stage of separation or the stock tank.

A maximum volume of hydrocarbon liquid could be obtained under stock tank conditions with a minimum loss of solution vapors by fractionating the last-stage separator liquid. This implies using a simple fractionating column, where the vapors liberated by increasing the bottom temperature are counterflowed with the cool feed introduced from the top. Interaction takes place on each tray in the column. The vapors act as a stripping agent and the process is described as stabilization.

7.4.2 Equipment and Operation

In general, a conventional fractionating column would require main auxiliaries such as reflux, pumps, condensers, cooling water, and utilities frequently not available on site in oil fields. Stabilizers or stripping columns, on the other hand, can be operated with a minimum of these auxiliaries. Figure 4 depicts a stabilizer in its simplest form.

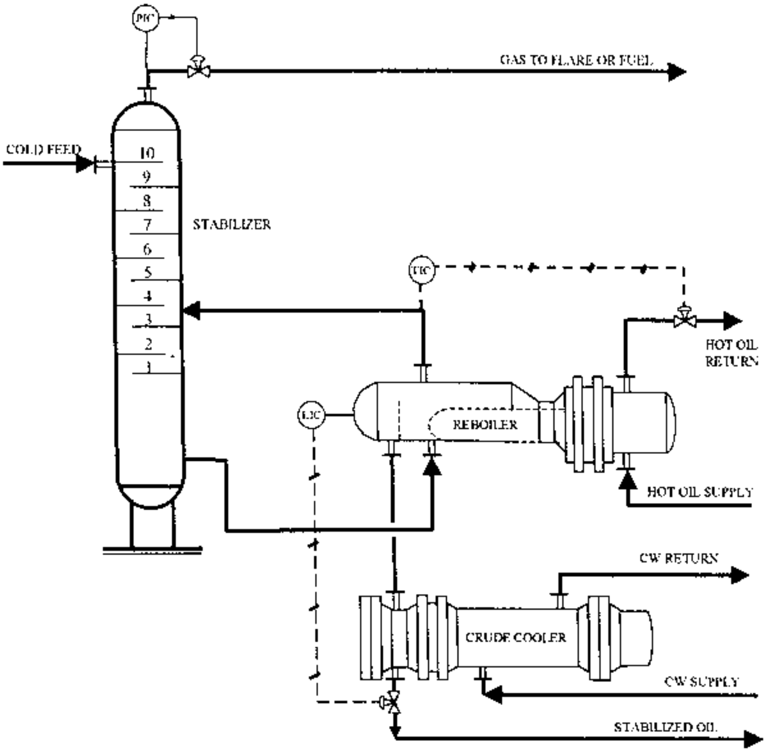

Figure 4 Typical trayed stabilizer.

A cold feed stabilizer normally operates with a fixed top and bottom temperatures. The former is kept as low as possible to maximize recovery, whereas the latter is controlled to maintain the product bottom pressure. It is of interest to mention that the overhead gas temperature is identical to the liquid feed temperature because the ratio of masses of vapor leaving the column to liquid feed entering is rather small.

Most stabilizers operate above 200 psia and consist of 20 bubble trays. High-pressure stabilizers have more trays because of the higher temperature gradient between the top and the bottom trays. More trays allow the column to operate closer to equilibrium. Columns less than 20 inches in diameter generally use packing rather than trays. A useful rule of thumb is that 1 ft^2 of tower area could handle about 100 bbl/day of stock tank liquid. In some designs, the cold feed is introduced several trays below the top tray, using the upper trays as a scrubber in order to prevent liquid carry over during burping.

Field operation of a stabilizer is described as follows. Relatively cool liquid (oil) exiting the GOSP is fed to the top plate of the column where it contacts the vapor rising from below. The rising vapors strip the lighter ends from the liquid (i.e., acting as a stripping agent). At the same time, the cold liquid—acting as an internal reflux—will condense and dissolve heavier ends from the rising vapor, similar to a rectification process. The net separation is very efficient as compared to stage separation (3–7% more).

To have a stabilized product of certain specifications, in theory a stabilizer can be operated at multiple combinations of tower pressure and bottom temperature. In general, as the tower pressure is increased, more light ends will condense in the bottom. In normal operation, it is best to operate the tower at the lowest possible pressure without losing too much of the light ends at the initial feed flash. This will minimize "burping" and cause the column to operate near equilibrium. In addition, lower operating pressures require less reboiler duty with less fuel consumption.

Operating data for a 40,000-bbl/day nonrefluxed stabilizer are given in Figure 5.

7.4.3 Main Features and Applications of Stabilizers

Stabilizers used for oil production field operations should have the following features:

- They must be self-contained and require minimum utilities that are available in the field, such as natural gas for fuel.

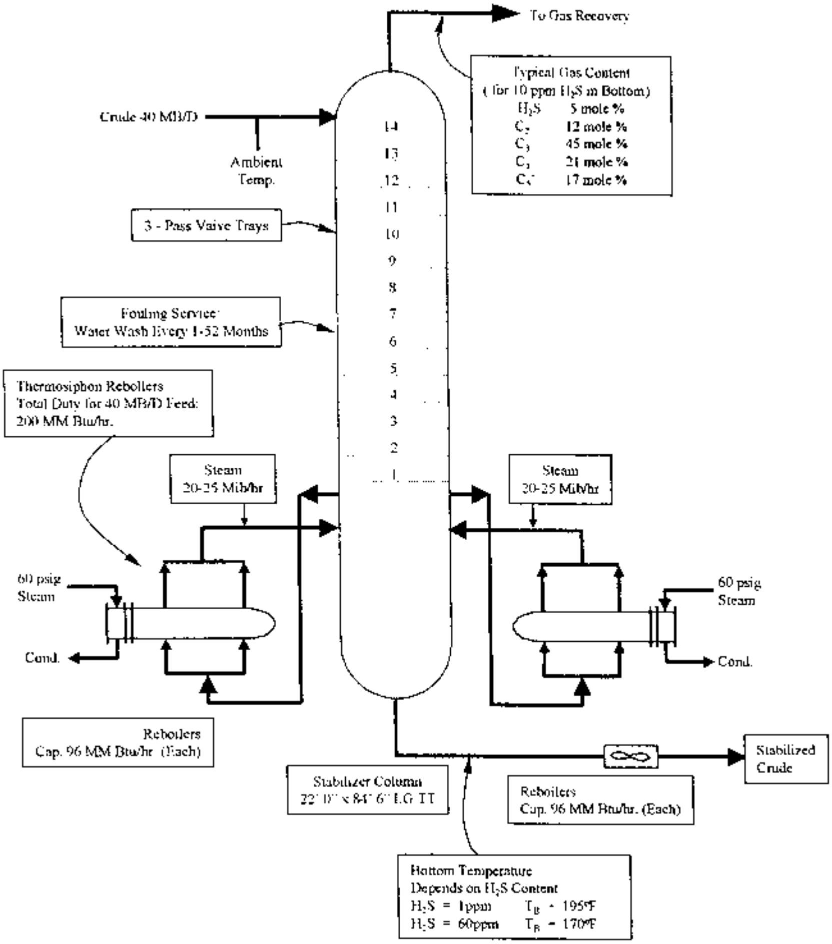

Figure 5 Typical trayed stabilizer with operating data for 40 MB/D of oil.

- Stabilizers must be capable of unattended operation and to stand fail-safe operation.
- Stabilizers must be equipped with simple but reliable control system.
- They should be designed in a way to make them accessible for easy dismantling and reassembly in the field.
- Maintenance of stabilizers should be made simple and straightforward.

Stabilizer's applications, on the other hand, are justified over simple stage separation under the following operating conditions:

- The first-stage separation temperature is between 0°F and 40°F.
- The first-stage separation pressure is greater than 1200 psig.
- The liquid gravity of the stock tank oil is greater than 45°API.
- Oil to be stabilized contains significant quantities of pentanes plus, even though the oil gravity is less than 45°API.
- Specifications are set by the market for product compositions—obtained from an oil—that require minimum light ends.

7.5 CRUDE OIL SWEETENING

Apart from stabilization problems of "sweet" crude oil, "sour" crude oils containing hydrogen sulfide, mercaptans, and other sulfur compounds present unusual processing problems in oil field production facilities. The presence of hydrogen sulfide and other sulfur compounds in the well stream impose many constraints. Most important are the following:

- Personnel safety and corrosion considerations require that H_2S concentration be lowered to a safe level.
- Brass and copper materials are particularly reactive with sulfur compounds; their use should be prohibited.
- Sulfide stress cracking problems occur in steel structures.
- Mercaptans compounds have an objectionable odor.

Along with stabilization, crude oil sweetening brings in what is called a "dual operation," which permits easier and safe downstream handling and improves and upgrades the crude marketability.

Three general schemes are used to sweeten crude oil at the production facilities:

Process	Stripping Agent
1. Stage vaporization with stripping gas	Mass (gas)
2. Trayed stabilization with stripping gas	Mass (gas)
3. Reboiled tray stabilization	Energy (heat)

1. **Stage vaporization with stripping gas**. This process—as its name implies—utilizes stage separation along with a stripping agent.

Hydrogen sulfide is normally the major sour component having a vapor pressure greater than propane but less than ethane. Normal stage separation will, therefore, liberate ethane and propane from the stock tank liquid along with hydrogen sulfide. Stripping efficiency of the system can be improved by mixing a lean (sweet) stripping gas along with the separator liquid between each separation stage.

Figure 6 represents typical stage vaporization with stripping gas for crude oil sweetening/stabilization. The effectiveness of this process depends on the pressure available at the first-stage separator (as a driving force), well stream composition, and the final specifications set for the sweet oil.

2. **Trayed stabilization with stripping gas**. In this process, a tray stabilizer (nonreflux) with sweet gas as a stripping agent is used as shown in Figure 7. Oil leaving a primary separator is fed to the top tray of the column countercurrent to the stripping sweet gas. The tower bottom is flashed in a low-pressure stripper. Sweetened crude is sent to stock tanks, whereas vapors collected from the top of the gas separator and the tank are normally incinerated. These vapors cannot be vented to the atmosphere because of safety considerations. Hydrogen sulfide is hazardous and slightly heavier than air; it can collect in sumps or terrain depressions.

   This process is more efficient than the previous one. However, tray efficiencies cause a serious limitation on the column height. For an efficiency of only 8%, 1 theoretical plate

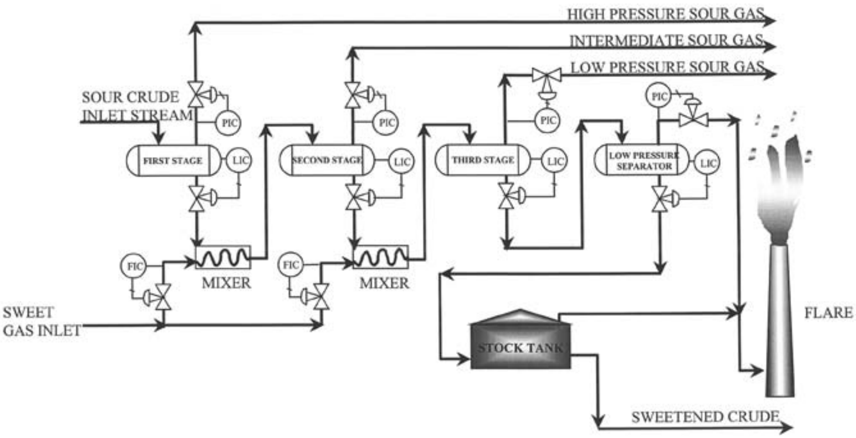

Figure 6 Crude sweetening by stage vaporization with stripping gas.

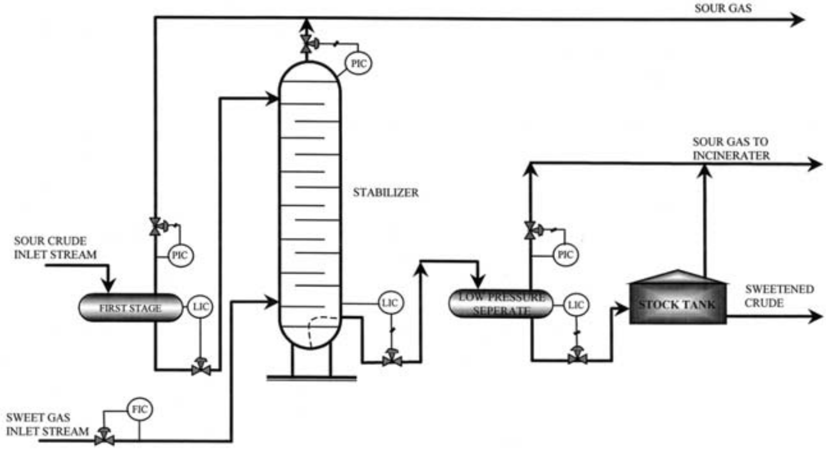

Figure 7 Crude sweetening by trayed stabilization with stripping gas.

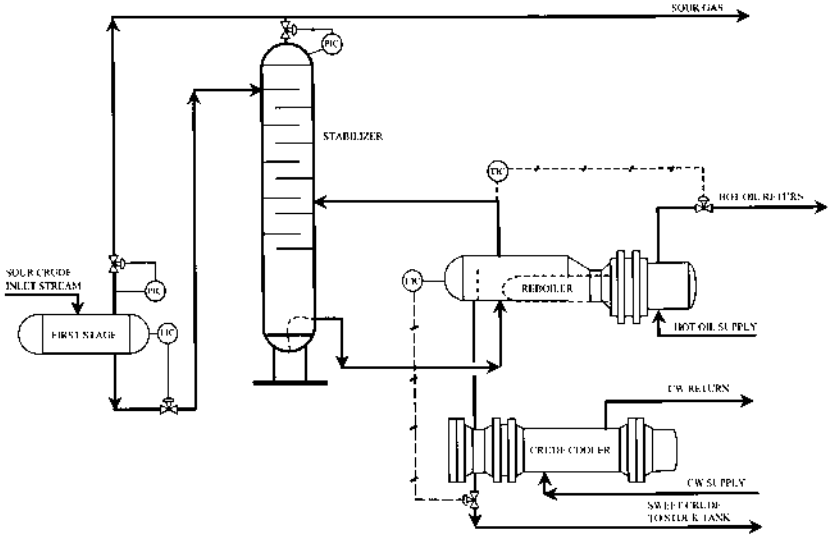

Figure 8 Crude sweetening by reboiled trayed stabilization.

would require 12 actual trays. Because trays are spaced about 2 ft apart, columns are limited to 24–28 ft high, or a maximum of two theoretical trays.

3. **Reboiled trayed stabilization**. The reboiled trayed stabilizer is the most effective means to sweeten sour crude oils. A typical

reboiled trayed stabilizer is shown in Figure 8. Its operation is similar to a stabilizer with stripping gas, except that a reboiler generates the stripping vapors flowing up the column rather than using a stripping gas. These vapors are more effective because they possess energy momentum due to elevated temperature.

Because hydrogen sulfide has a vapor pressure higher than propane, it is relatively easy to drive hydrogen sulfide from the oil. Conversely, the trayed stabilizer provides enough vapor/liquid contact that little pentanes plus are lost to the overhead.

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REVIEW QUESTIONS

1. Explain the function of crude oil stabilization operation.
2. Why do lighter crude oils need stabilization but heavier crude oil may not need stabilization?
3. What is the difference between sweet and sour crude oils?
4. What is the function of the crude oil sweetening process?
5. Why it is necessary to remove H_2S from crude oil?
6. Describe how stabilization by flashing is achieved. Under what conditions could this method of stabilization be used.
7. Describe the principle of stripping operation for crude oil stabilization.
8. What are the conditions under which crude oil stabilization should be employed?
9. Describe the principles of operation of the following crude oil sweetening processes:
    (i) Stage vaporization with stripping gas
    (ii) Trayed stabilization with stripping gas
    (iii) Reboiled trayed stabilization

8

Storage Tanks and Other Field Facilities

This chapter is devoted to the discussion of storage tanks of crude oil and other hydrocarbons, vapor recovery units (VRUs), and piping in the oil field, including gathering schemes.

8.1 STORAGE TANKS

8.1.1 Introduction

The design of storage tanks for crude oil and petroleum products requires, in general, careful consideration of the following important factors:

- The vapor pressure of the materials to be stored
- The storage temperature and pressure
- Toxicity of the petroleum material

In order to meet the environmental constraints on air pollution, to prevent fire hazards, and to avoid losses of valuable petroleum products at the same time, it is recommended to adopt the following:

- The use of floating-roof tanks for petroleum materials with a vapor pressure of 1.12–11.5 psia (at the storage temperature) or
- Using fixed-roof tanks along with the VRU system (to be described later).

These alternatives are schematically illustrated in Figure 1.

Storage tanks for crude oil are needed in order to receive and collect oil produced by wells, before pumping to the pipelines as well as to allow for measuring oil properties, sampling, and gauging.

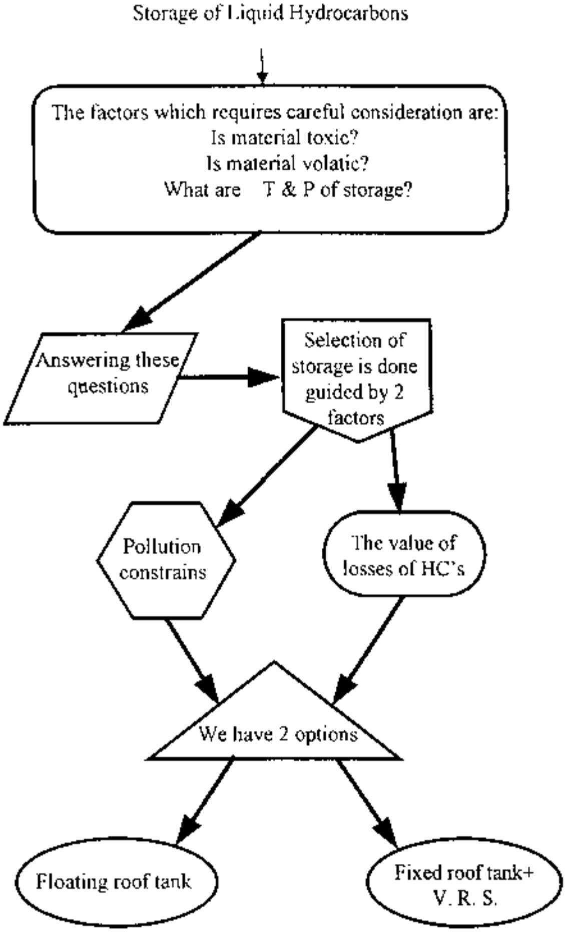

Figure 1 Vapor recovery units are used to minimize the loss of hydrocarbon vapor formed during storage.

8.1.2 Types of Storage Tank

The main features of some of the common types of storage tank used by the petroleum industry in general are presented in Table 1. The atmospheric tank, or standard storage tank, is one that is designed to be used within plus or minus a few pounds per square inch of atmospheric pressure. It may be open to the atmosphere (vented) or enclosed.

As will be explained next, an effective method of preventing vent loss in a storage tank is to use one of the many types of variable-volume tank

Table 1 Summary of Refinery Storage Tanks

Characteristics	Standard Storage Tanks	Conservation-Type Storage Tanks		
		I (Floating Roofs)	II (Variable-Vapor-Space)	III (Pressure Storage)
Evaporation losses	High	Significantly reduced	Significantly reduced	Prevented or eliminated
Operating conditions	Recommended for liquids whose vapor pressure is atmospheric or below at storage conditions (vented).	Allow no vapor space above the liquid; level (no venting)	Allow the air-vapor mixture to change volume at constant or variable pressure (no venting)	Allow the pressure in the vapor space to build up. Tanks are capable of withstanding the maximum pressure without venting.
Sub-classification	1. Rectangular 2. Cylinderical: a) Horizontal b) Vertical		1. Lifter roof, which is a gas holder mounted on a standard storage tank. 2. Vapor-dome	1. Low-pressure storage normally designed for 2.5–5 psig (up to 15 psig) 2. High pressure storage: 30–200 psig
Typical types	Cone-roof-vertical (cylinderical tanks)	Floating-roof, wiggins-Hidek type	Lifter roof tanks, wiggins dry seal type	Spheroids and hemispheroids for low pressure storage, spheres for high pressure storage
Applications	Heavy refinery products	Sour crude oils, light crude oils, light products.	Light refinery product and distillates	Spheriods are used to store aviation, motor, jet fuels. Spheres are used to store natural gasoline and LPG.

(type II in Table 1). These are built under API Standard 650. They may have *floating roofs* of the double-deck or single-deck type. These are lifter-roof types in which the roof either has a skirt moving up and down in an annular seal or is connected to the tank shell by a flexible membrane.

8.1.3 Sizing the Storage Tanks

The following is the guidelines to estimate the storage capacity for an oil-producing facility; select the number of storage tanks of crude oil.

1. To determine the capacity of storage tanks, the following rules apply:
    (a) For **one** well, the capacity per storage tank should be adequate enough for at least 2–3 days' production.
    (b) For a **group** of wells (entire field), the total tank capacity should be adequate enough for at least 3–4 days of oil production by that field.
2. To determine the number of storage tanks as a function of the number of producing wells. This relationship is represented graphically in Figure 2.

To illustrate the application of these guidelines, the following example is cited. However, it should be emphasized that their use is restricted to small- to medium-size oil field installations.

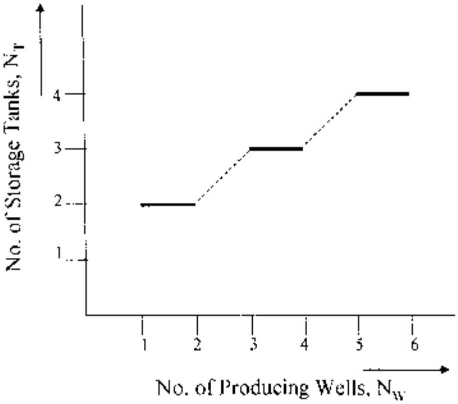

Figure 2 Guidelines for selecting the number of storage tanks as a function of the number of producing wells (for small- to medium-size installations).

Example 1

For a lease consisting of five producing wells, each with an average production capacity of 10,000 bbl/day, find the approximate total storage capacity and the number of storage tanks.

Solution

Using rule a,

$$\begin{aligned}\text{Capacity or the size of one tank} &= \text{2–3 days of oil production/well}\\ &= (2.5)(10{,}000)\\ &= 25{,}000\ \text{bbl/well}\end{aligned}$$

Now, for five wells,

$$\begin{aligned}\text{Total storage capacity} &= (5)(25{,}000)\\ &= 125{,}000\ \text{bbl}\end{aligned}$$

Using correlation,

$$\begin{aligned}\text{Storage capacity for the entire lease} &= \text{3–4 days of oil}\\ &\qquad \text{production by lease}\\ &= (3.5)(50{,}000)\\ &= 175{,}000\ \text{bbl}\end{aligned}$$

Conclusions:

The number of storage tanks according to rule a is 125,000/25,000 = 5 tanks

The number of storage tanks according to rule b, assuming storage capacity per tank the same as computed in rule a is 175,000/25,000 = 7 tanks

Finally, the number of storage tanks as computed graphically, using Figure 2 is 4. Now, comparing the results obtained, the appropriate number of storage tanks required by the lease is five to six.

8.2 VAPOR RECOVERY UNITS

8.2.1 Background

The loss of hydrocarbon vapors formed above crude oil or its products—when stored—could be minimized using what is called vapor recovery units (VRUs). If allowed to escape to the atmosphere, these vapors will

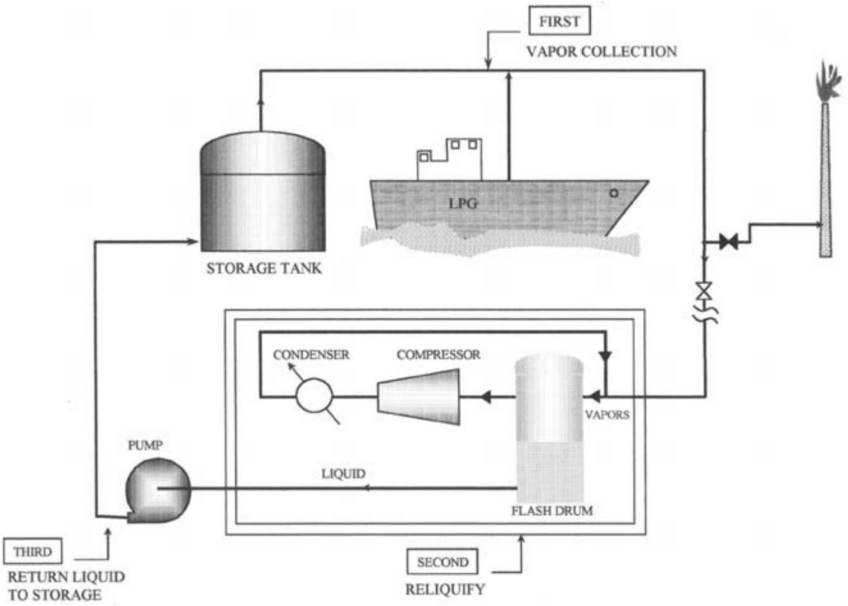

Figure 3 Main functions of vapor recovery system.

not only cause a loss of income due to loss of hydrocarbon volume and change in the API of the oil but will also lead to pollution and fire hazards.

The three main functions for the vapor recovery system are (as illustrated in Fig. 3) as follows:

1. To collect vapor from storage/loading facilities
2. To reliquefy vapors
3. To return liquid hydrocarbons to storage

Basically, when we talk about a VRU, what we are looking for is to hook our storage tanks to a "breather" system such as the following:

- During the day, when the temperature rises and vaporization of the hydrocarbons occur, excess vapors can be released and collected by the VRU.
- At night, when the vapors cool and condensation takes place leading to partial vacuum, vapors from the VRU will be admitted into the tanks.
- While pumping in and pumping out liquids to and from the storage tanks, vapors could be vented, [i.e., collected and drawn in, respectively, by such a breather system (VRU)].

8.2.2 Types of Storage Loss

In general, hydrocarbon losses in storage tanks are identified as follows:

Working losses

(a) Filling
(b) Emptying

Other losses

(a) Breathing
(b) Standing
(c) Boiling

Filling losses occur when vapors are expelled from a tank as it is filled, no matter how the vapors are produced. This loss occurs when the pressure inside the tank exceeds the relief-valve pressure. For API tanks, the relief pressure is low and, therefore, filling losses can be relatively high.

Emptying losses are experienced by the vapors that are expelled from a tank after the liquid is removed from it. Because vaporization lags behind the expansion of the vapor space during withdrawal, the partial pressure of a hydrocarbon vapor drops. Enough air enters during the withdrawal to maintain the total pressure at the barometric value. However, when vaporization into the new air reaches equilibrium, the increase in the vapor volume will cause some vapor expansion.

Breathing losses occur when vapors are expelled from a tank under one of the following conditions:

1. The thermal expansion of the existing vapors
2. An expansion caused by barometric pressure changes
3. An increase in the amount of vapors from added vaporization in the absence of a liquid level change

Breathing losses take place in most types of tanks and occurs when the tank's limits of pressure or volume changes are exceeded.

The fixed-roof API type tanks used to store stock tank oil are designed for only for a few inches of water pressure or vacuum and suffer relatively large breathing losses.

Standing losses are losses of vapor which result from causes other than breathing or a change in liquid level in tanks. Sources of standing losses are vapor escape from hatches or other openings and from glands, valves, and fittings.

Boiling losses occur when liquid boils in a tank and vapors are expelled. In other words, the vapor pressure of the liquid exceeds the surrounding pressure.

8.2.3 Vapor Recovery Methods

Ideally, it would be best to design a tank or a storage system to operate at pressures high enough to suppress evaporation; hence minimizing evaporation losses. However, this is not generally economical; also, refiners require crude oil to meet maximum vapor pressure specifications.

Various methods can be recommended to recover vapors generated in storage tanks and from other sources such as liquified petroleum gas (LPG) tankers. These usually involve one or a combination of the following schemes implemented through what is referred to as the VRU:

- **Absorption**: Usually carried out under pressure using a liquid solvent of higher molecular weight than that of the vapors being recovered. Vapors are then separated from the rich solvent, which is recycled in the process as "lean solvent."
- **Condensation**: Vapors can be totally or partially condensed by compression and cooling, as shown in Figure 3.
- **Simple cooling**: Cooling the vapors without compression may condense the vapors, but it is not normally economical unless refrigeration is applied.
- **Adsorption**: Hydrocarbon vapors mixed with noncondensable gases, such as air, can be adsorbed by molecular sieves, activated charcoal, or silica gel. Heat or depressurization will remove the adsorbed vapors from the solid bed. The vapors could then be condensed for recovery.

The basic part of equipment operating the VRU is the vapor regulator setup (see Figure 4). The basic functions of the regulator are the following:

1. Release vapor from the storage tank battery when the normal operating pressure within the system increases beyond a preset value

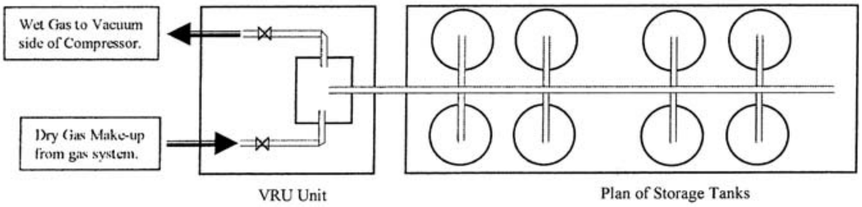

Figure 4 Vapor regulator system connected to storage tanks.

2. Add vapor to the battery system if the normal operating pressure decreases and reaches a preset value

One should mention that, in addition to this vapor regulator, other automatic relief valves are found in the VRU. The system works automatically and in harmony. The breather valve operates if excessive pressure or vacuum exists, whereas the manhole relief functions if abnormal pressure or vacuum is experienced in the system. Values of pressure settings of the different instruments in the system are illustrated in Figure 5.

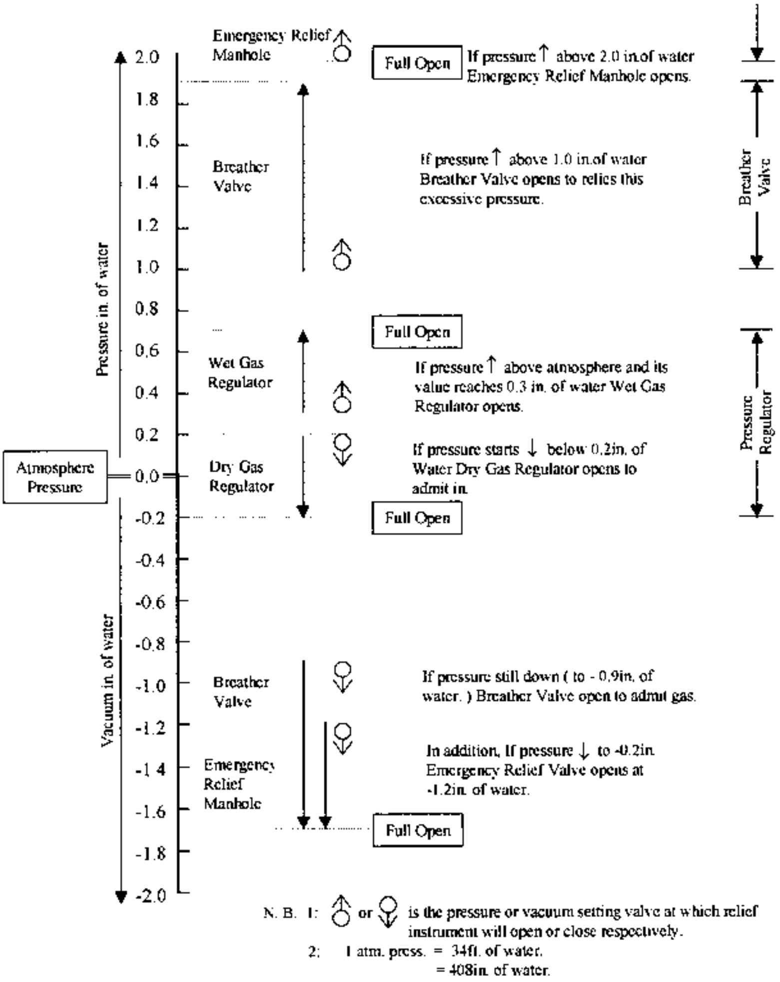

Figure 5 Chart for pressure setting for the VRU. (After Chilingar and Beeson.)

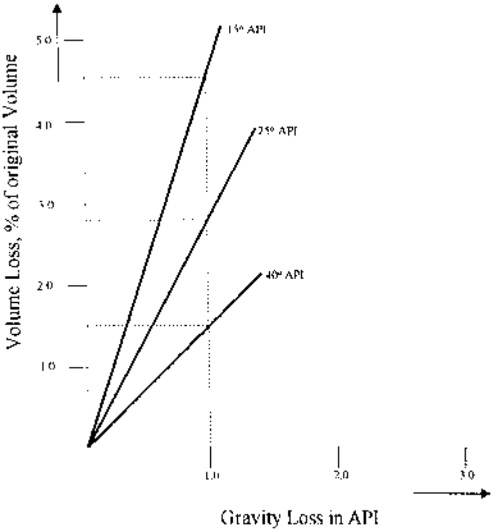

Figure 6 Change of volume with gravity decrease.

Finally, it should be pointed out that the loss of vapors from oil during storage results in the following:

A decrease in the API gravity of the oil, which degrades its quality
A reduction in the volume of oil to be sold

The loss in volume of oil per degree of API gravity reduction varies depending on the original gravity of the oil. On average, a 2% volume loss is experienced per one degree reduction in the API gravity of the oil, as exemplified in Figure 6.

8.3 PIPING AND THE OIL FIELDS

8.3.1 Introduction

Today, there is great diversity in size of pipes used to carry crude oil and refined products, ranging from 2 in. to as much as 36 in. and in some cases, even 48-in. piping is used. In general, there are four types of pipeline:

1. Oil field gathering pipelines; their function in an oil field is of great impact on production operations.
2. Pipelines which run from the oil field to loading ports and are complementary to ocean transport.

3. Long-distance pipelines, which naturally shorten the alternative sea route.
4. Pipelines which transport oil from ports of discharge to inland refineries located in industrial areas, remote from a seaport.

8.3.2 Pipeline Gathering Schemes

Figure 7 illustrates the transport of oil by pipelines. It is a collection of pipelines that takes oil from wellhead all the way to the loading ports or refineries. Crude oil is collected from each individual wellhead by small-diameter pipelines, which then converge on a collecting center. At the collecting center, oil is directed to the gas–oil separating plant (GOSP) to

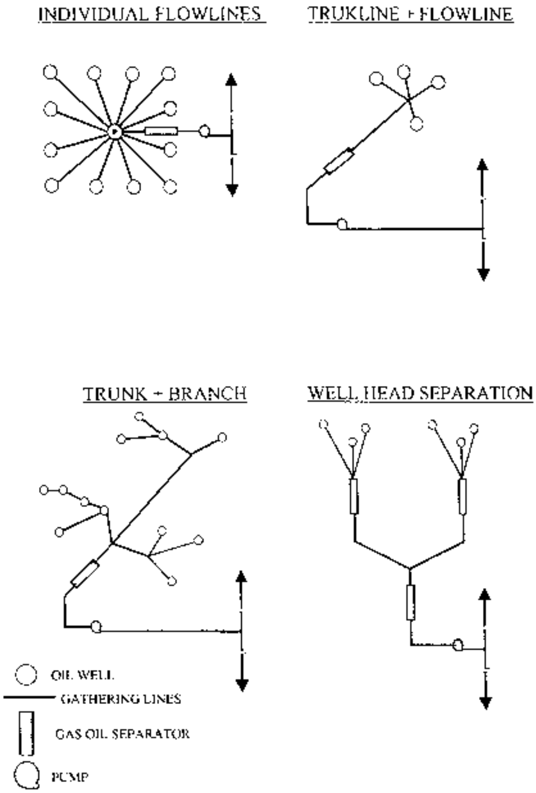

Figure 8 Oil field gathering system: typical four schemes. (After Yocum.)

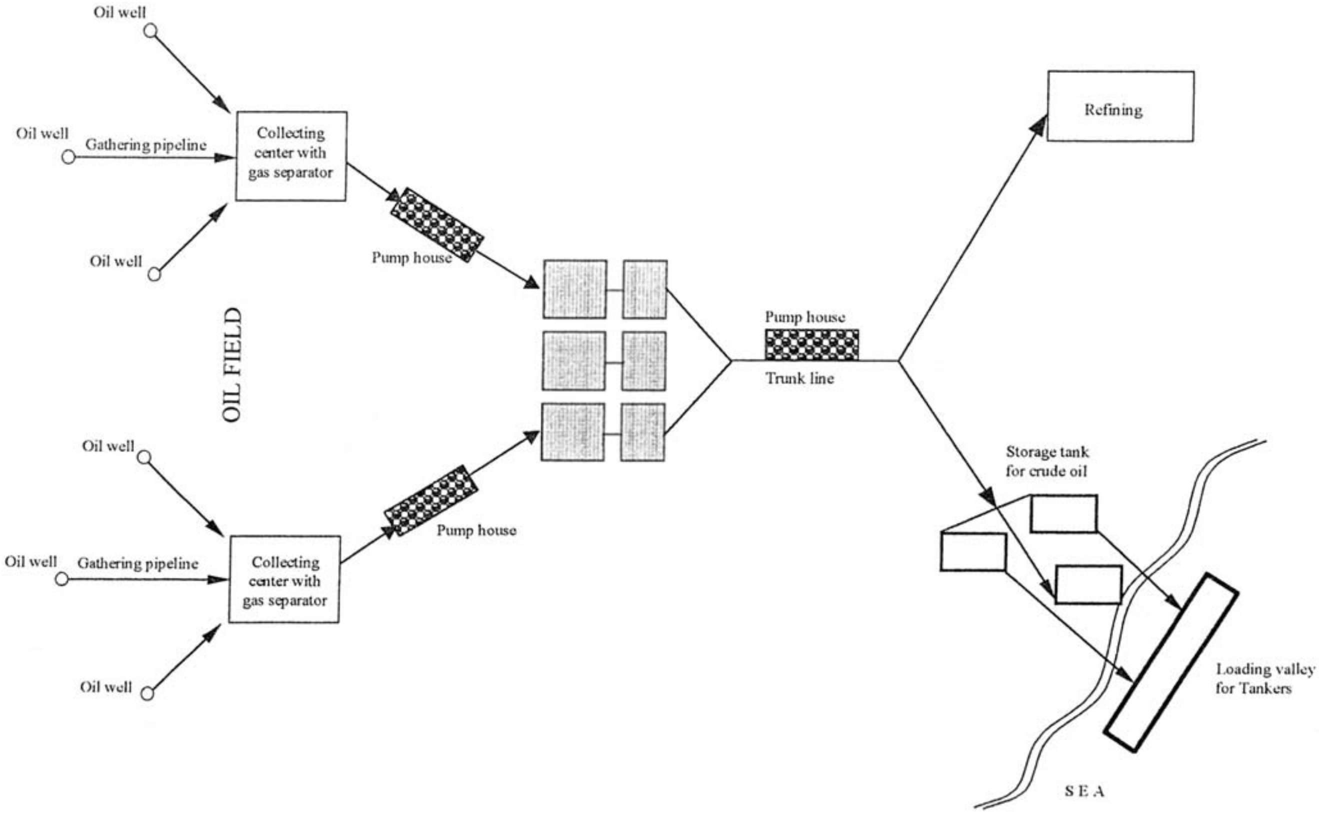

Figure 7 Piping system and oil field. (After Abdel-Aal and Schmelzlee.)

separate gas from oil. Usually, a number of collecting centers are found in an oil field.

From the collecting center, pipes of large diameter lead the crude oil to a tank farm. From there, oil is conveyed either to refineries or to storage tanks at the terminals for overseas delivery by tankers or by long-

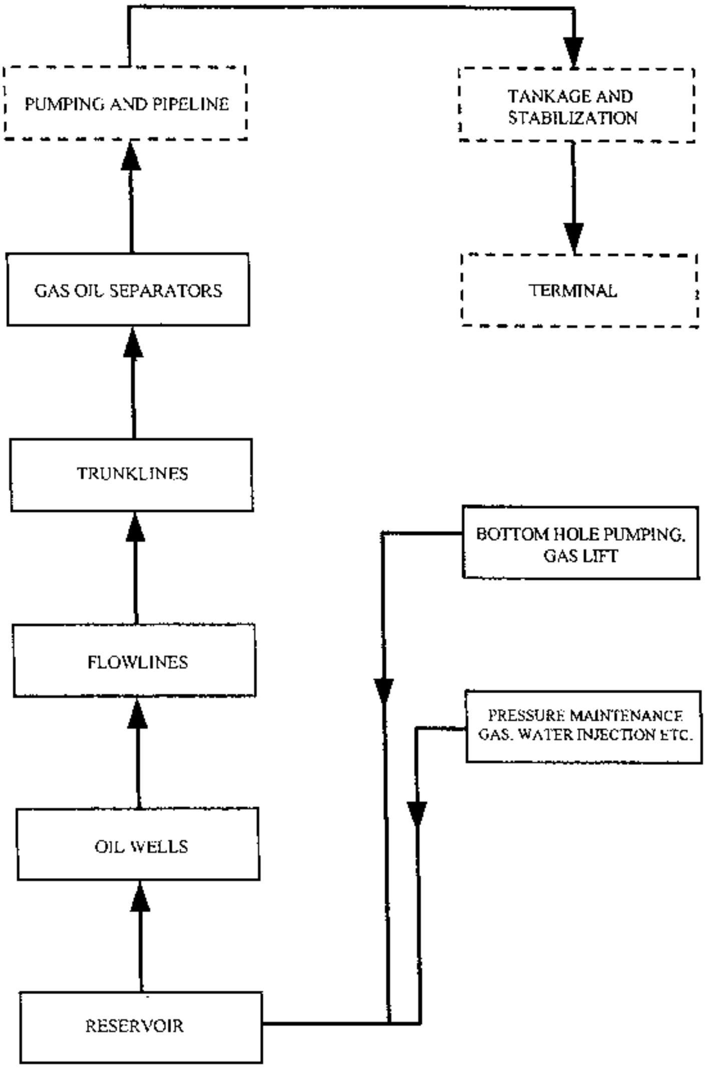

Figure 9 Different functions of piping. (After Yocum.)

distant pipelines. Figure 8 presents a block diagram for the different functions of piping in an oil field.

As far as the design of a gathering system in an oil field is concerned, four different schemes are presented:

1. Individual (gathering) flow lines through which oil wells are connected to a central GOSP.
2. Trunklines and short flow lines. Well effluents, in this scheme, are directed into a large trunkline via short flow lines. Crude oil is directed to the GOSP by a trunkline.
3. Trunkline and branckes, where major trunklines are 30–50 miles long and 24–30 in. in diameter, gather crude oil from branch trunklines, which, in turn, collect oil from short flow lines connected to the wells. The branch trunklines are smaller in diameter than the main trunklines (16–20 in.) and much shorter.
4. Wellhead separation, in which case the oil exiting different wells is delivered to the GOSP through very short gathering lines. The separated crude oil is then transferred from GOSP using trunklines.

The function of the four schemes described are diagrammatically sketched in Figure 9.

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REVIEW QUESTIONS

1. List the different types of storage tank.
2. What is the difference between floating-roof tanks and fixed-roof tanks? Which one requires the installation of a vapor recovery unit? Why?
3. A small field has 20 producing wells, each well produces an average of 8000 bbl/day. Determine the required field storage capacity and the number of storage tanks needed.
4. Describe the five types of storage-tank loss.
5. What effects does vapor loss have on the stored crude oil?
6. Under what conditions should a vapor recovery unit be included in field facilities? Why?
7. What are the main functions of vapor recovery units?
8. Describe briefly the four methods used for vapor recovery.
9. What are the main functions of the vapor regulator in a vapor recovery unit?

9

Produced Water Treatment

9.1 INTRODUCTION

Production of crude oil and natural gas is usually associated with the production of water. During the early life of the petroleum fields, water-free production of oil and gas is normally experienced. However, water will eventually be produced later. The produced water may be water that exists within the petroleum reservoir as connate water or bottom water. Alternatively, water may be produced as a result of water-flooding operations, where water is injected into the reservoir to enhance the recovery.

Water production presents serious operating, economic, and environmental problems. Production of water with the crude oil or natural gas reduces the productivity of the well due to the increased pressure losses throughout the production system. This may either result in reduced production or necessitate the installation of costly artificial lifting systems to maintain the desired production levels. Production of water also results in serious corrosion problems, which add to the cost of the operation. As discussed in the previous chapters, production of water with the crude oil or natural gas requires the use of three-phase separators, emulsion treatment, and desalting systems, which further add to the cost of the operation.

In most situations, the produced water has no value and should be disposed of. In other situations, the produced water may be used for water flooding or reservoir pressure maintenance. The produced water, collected from the separation, emulsion treatment, and desalting systems, contains hydrocarbon concentrations that are too high for environmentally safe disposal. The presence of the hydrocarbon droplets in the water makes it difficult to inject the water into disposal wells or into water-injection wells for enhanced recovery operations. This is because the hydrocarbon droplets cause severs plugging of the formation. In all cases, the produced water must be treated to lower its hydrocarbon content to acceptable limits. For the heavy oil field, produced water may be used to generate the steam needed for oil recovery. In this case, additional chemical treatment will be needed to reduce the concentration of the salt and other minerals to make the water quality adequate for steam generation.

The purpose of this chapter is to present the concepts and procedures used for selecting and sizing the equipment used for removal of oil from the produced water.

9.2 PRODUCED AND TREATED WATER QUALITY

The quality of treated water (i.e., the maximum allowable oil concentration and maximum allowable oil droplet size) is determined to meet water-injection or disposal requirements. From an environmental point of view, it should be desirable to remove all of the oil from the produced water or at least allow the technically minimum possible. This, however, can impose substantial additional operating costs. Therefore, operators would usually provide the necessary water treatment to achieve the maximum allowable oil content. To properly design an efficient and economical treatment system that achieves this objective, knowledge of the produced water quality (oil concentration and droplet size distribution) is necessary. This is best determined from laboratory analysis of actual field samples. Such samples, however, are not normally available, especially when designing a treatment system for new field development.

Theoretically speaking, it is possible to determine the droplet size distribution throughout the various components of the production system and the separation and oil treatment equipment. However, most of the parameters needed to solve the governing equations, especially those involving dispersion and coalescence, are normally unknown. As discussed in previous chapters, the design of separation and oil treatment equipment determines the maximum oil droplet size remaining in the water. Several attempts have been made to determine the oil concentration in water for properly designed separation and treatment equipment. The results showed that the dispersed oil content ranges from 1000 to 2000 mg oil per liter of water. Unfortunately, as the water leaves the separation and treatment equipment, it flows through various restrictions (such as valves and bends) in the piping system before it reaches the water treatment facility. In its journey, the oil droplets are subjected to a series of dispersion and coalescence that makes it difficult to exactly determine the oil droplet size distribution in the water to be treated. Experience showed that a conservative assumption for design purposes would be to represent the droplet size distribution by a straight-line relationship between the droplet size and cumulative oil concentration, with the maximum oil droplet diameter being between 250 and 500 μm.

9.3 PRODUCED WATER TREATMENT SYSTEM

In general, produced water always has to be treated before it is disposed of or injected into the reservoir. The purpose of the treatment is to remove enough oil from the water such that the remaining amount of oil in the water and the oil droplet size are appropriate for the disposal or injection of the water. For example, for water disposal into underground formations and water injection into the producing reservoir, the pore size of the formation determines the allowable oil droplet size in the treated water. The maximum droplet size of the remaining oil in the water should be less than the minimum pore size of the formation to avoid plugging of the formation by the oil droplets. For water disposal into the sea, as is normally practiced in offshore operations, the amount and droplet size of the oil in the water is governed by environmental constraints.

Depending on the amount and droplet size of the oil in the produced water, the required quality of the treated water, and the operating conditions, water treatment may be achieved through a single or two stages of treatment. The single, or first, stage of treatment is normally known as the *primary treatment* stage; the second stage of treatment is known as the *secondary treatment* stage.

The equipment used for water treatment serves the function of allowing the oil droplets to float to the surface of the water, where they are skimmed and removed. For primary treatment, this may be achieved by using skim tanks for atmospheric treatment or skim vessels for treatment under pressure. Plate coalescers such as the *parallel plate interceptor* and *corrugated plate interceptor* are used to promote coalescence of the oil droplets to increase their size and thus speeds their floatation to the surface. Another device, known as the *SP pack*, is also used to promote coalescence of the oil droplets. For secondary treatment, plate interceptors, SP packs, and flotation units are normally used.

For offshore operations, water disposal must be through a disposal pile, skim pile, or SP pile. The deck drains normally contains free oil and must be treated before disposal. This can be done either in similar primary treatment equipment or directly through the various disposal piles.

9.4 WATER TREATMENT EQUIPMENT

The various produced water treatment equipment mentioned are described in this section. The main function of the treating equipment is to separate the free oil droplets from the water. The fluid may contain some dissolved gas, which will be liberated in the treating equipment and must be removed. Therefore, the produced water treatment equipment are, in essence, similar to the three-phase oil–water–gas separators. The main difference is that for water treatment equipment, water is the main and continuous phase and oil represents a small volume of the fluid mixture.

9.4.1 Filters

One of the very efficient ways of removing oil droplets from water is the use of filters. In this method of water treatment, produced water is made to flow through a bed of porous medium, normally sand, where the oil droplets are trapped in the filtering medium. At least two filters arranged in parallel are used. As the filter in use gets clogged, the flow is directed to the other filter and the clogged filter is backwashed using water or solvent. The backwash fluid must be treated or disposed of properly, which adds more complications and cost to the water treatment process. Several onshore successful operations have been reported in which sand filters were used to yield treated produced water with oil content as little as 25 mg/L of water.

9.4.2 Precipitators

In this method of treatment, the produced water is directed through a bed of porous material, such as excelsior, placed inside a horizontal vessel that is similar in design to the three-phase separator to promote the coalescence of oil droplets. The coalesced large oil droplets flow upward, countercurrent to the downward flow of the water where it can be skimmed out of the vessel. Although this method has been effective in treating produced water to desired quality, clogging of the coalescing medium represented a serious problem, which limited the use of such precipitators.

9.4.3 Skim Tanks and Vessels

The skim tanks and vessels are the simplest equipment used for primary treatment of produced water. Skim tanks and skim vessels are generally similar in shape, components, and function. However, the designation of skim tanks is associated with atmospheric treatment, whereas skim vessels are used when water treatment is performed under pressures above the atmospheric pressure. These equipment are normally large in volume to provide residence time that is sufficiently long (10–30 min) for the coalescence and gravity separation of the oil droplets.

Pressure vessels are more expensive than atmospheric tanks. However, the choice is controlled by the overall requirements of the water treatment system. Pressure vessels are normally preferred over atmospheric tanks for the following reasons:

1. To avoid the potential gas venting problems associated with atmospheric tanks
2. To eliminate the potential danger of overpressure that may occur in an atmospheric tank
3. To eliminate the need for pumps that may be required to deliver the treated water to other secondary treating equipment or to other locations for disposal

The technical aspects, benefits, and cost should all be considered in deciding on the pressure rating of the skimmers.

Skimmers can be either horizontal or vertical in configuration. The shape and internal components of the skimmers are generally similar to those of the three-phase separators. Figure 1 shows a schematic of a horizontal skimmer. As shown in the figure, the produced water enters the skimmer below the water–oil interface and flows horizontally along the length of the vessel. The oil droplets coalesce and rise to the oil pad perpendicular to the direction of the water flow. The oil flows over the weir into the oil collection section and out of the skimmer. The height of the oil pad is controlled by the weir as shown in Figure 1. Alternatively,

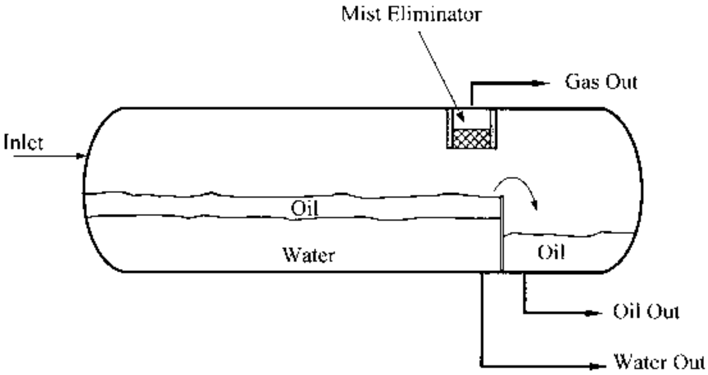

Figure 1 Schematic of a horizontal skimmer.

the height of the oil pad may be controlled by an interface level controller or by an external water leg. The treated water is withdrawn from the skimmer at the bottom of the vessel. The liberated gas leaves the vessel at the top through a mist extractor.

Figure 2 shows a schematic of a vertical skimmer that is equipped with an inlet spreader and a water outlet collector, which work to even the distribution of the incoming and outgoing flow, respectively. As with the horizontal skimmers, the produced water enters the vessel below the oil–water interface. Water flows downward while the oil droplets rise upward to the oil pad. Because of this countercurrent flow of the water and oil, vertical vessels are generally less efficient than horizonal vessels. The oil is skimmed over the weir into the oil collection section, where it is withdrawn from the vessel. The water outlet is at the bottom of the vessel through the water collector. The liberated gas leaves at the top of the vessel through a mist extractor.

Vertical vessels are preferred over horizontal vessels when treating water containing sand or other solids. A sand drain at the bottom of the vertical vessel always provide a simpler and more effective means for

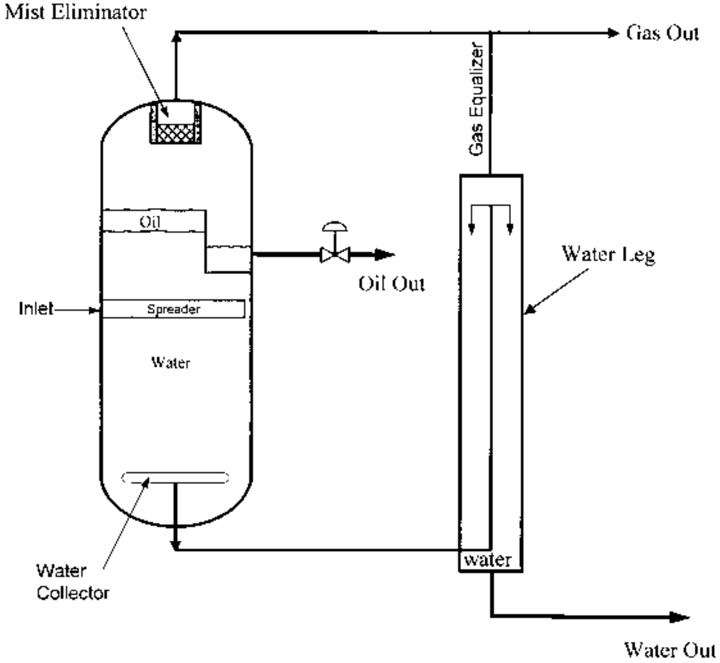

Figure 2 Vertical skimmer schematic.

cleaning the vessels as compared to the sand drains in horizontal vessels. Further, vertical vessels are better than horizontal vessels with regard to handling liquid surges. Surging in horizontal vessels tends to create internal waves, which results in a false indication of a high liquid level within the vessel and leads to false high-level shutdown.

Another type of skimmer is the API separator, which is basically a horizontal, rectangular cross-section tank. This type of skimmer is mostly used for treatment under atmospheric conditions.

9.4.4 Plate Coalescers

There are mainly two types of plate coalescer: the *parallel plate interceptors* (PPI) and the *corrugated plate interceptors* (CPI). Both types consist of a set of parallel plates that are spaced a short distance apart and are inclined by an angle of 45°.

The PPI was the first form of plate coalescers where a series of inclined parallel plates is installed inside an API separator, as shown in Figure 3. The water flow is split between the plates; therefore, the oil droplets need first to rise along the short distance between two consecutive plates where coalescence occurs (Fig. 4). Due to gravity, the large oil droplets move upward along the bottom surface of the inclined plate and then vertically upward to the oil collection section, where oil is skimmed out of the tank. Sediments in the water move downward to the bottom of the tank, where they can be removed.

The CPI is the most commonly used plate interceptor in the industry. The CPI was an improvement over the PPI, where the surface of the parallel plates was made corrugated with the axis of the corrugations being parallel to the direction of water flow. As shown in Figure 5, the water to be treated flows downward through the CPI pack. The oil rises upward, counter to the water flow and accumulates at the corrugations. The accumulated oil flows along the axis of corrugations and upward to the oil–water interface.

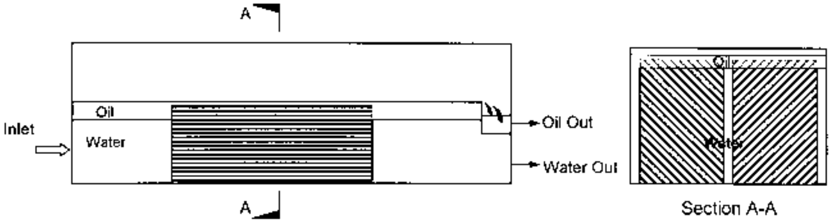

Figure 3 The parallel plate interceptor.

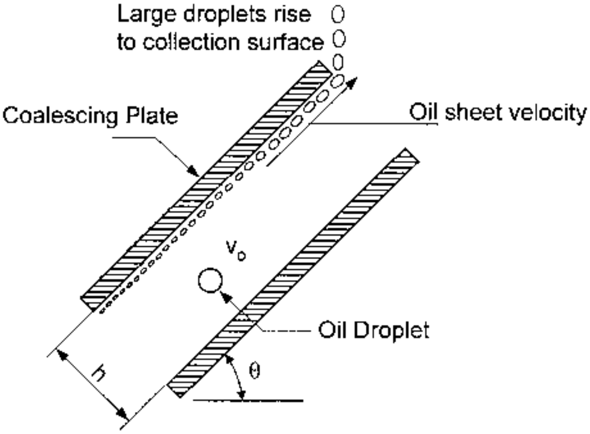

Figure 4 Oil droplets rise between two consecutive plates in the PPI.

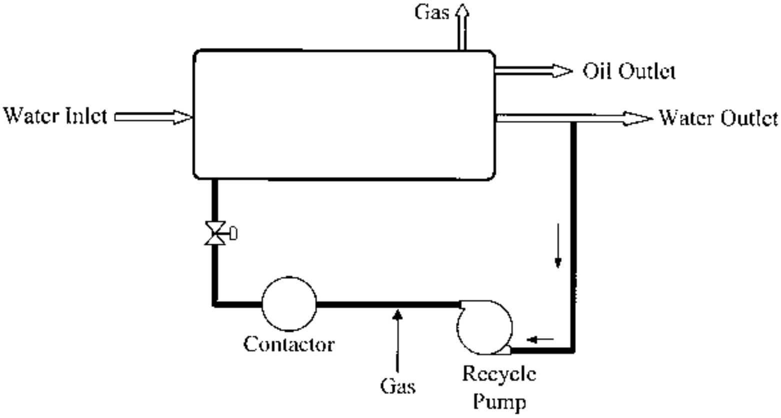

Figure 5 The corrugated plate interceptor.

Both PPI and CPI are normally used for water treatment under atmospheric conditions. The 45° inclination of the plates may present a problem when produced water contains appreciable amounts of sediments or sand, particularly oil-wet sand. The solids have the tendency to adhere to the surface of the plates at such an angle of inclination, which may cause clogging of the plates. To avoid such a problem and to enable treatment at pressures higher than atmospheric, modified CPI equipment known as *cross-flow devices* have been developed. In such equipment, the angle of inclination of the plates is made steeper than 45° (normally 60° to the horizontal) and the plate pack is placed inside a pressure vessel (vertical or horizontal) such that the water flow is perpendicular to the axis of corrugations in the plates. Vertical vessels are generally preferred for handling sediments and sand problems. Regular CPI units are less expensive and more efficient than cross-flow devices, but the later should be used for treatment under pressure and for water containing large amounts of sand or sediments.

9.4.5 Serpentine-Pipe Packs

The *serpentine-pipe pack* (SP pack) is another device that is used to promote coalescence of the oil droplets and thus facilitates their separation by gravity. The coalescence concept for the SP pack is, however, different from that of the previously described equipment. Water is forced to flow through a serpentine path that is properly sized to create turbulence that is sufficient to cause coalescence without causing shearing of oil droplets below a specified size. The SP packs are available in standard dimensions ranging from 2 to 8 in. in diameter for handling water flow ranging from 900 to 73,000 bbl/day (BPD). Such packs are designed to develop a drop size distribution curve with a maximum drop size of 1000 μm. By producing such a drop size distribution, gravity settling becomes very efficient. In fact, SP packs can result in about 50% additional oil removal as compared to gravity settling alone. The SP pack is normally placed inside any gravity settling vessel with the water inlet diameter being the same as the SP diameter. SP packs can be staged in series to allow successive coalescence and removal of oil as the water flows from one stage to the next.

9.4.6 Flotation Units

Flotation units utilize a completely different concept is removing oil droplets from water. In this type of treatment equipment, a large number of small gas bubbles are produced within the water. As the gas bubbles rise upward, they carry the oil droplet to the surface, where they accumulate and are then skimmed out of the unit. Flotation units are classified into two types based on the method by which the gas bubbles are produced. These are the *dissolved gas units* and the *dispersed gas units*.

Dissolved Gas Flotation Units

As shown in Figure 6, a portion of the treated water (between 20% and 50% of the effluent) is taken and saturated with natural gas in a contactor at a pressure between 20 and 40 psi. The amount of gas used in standard

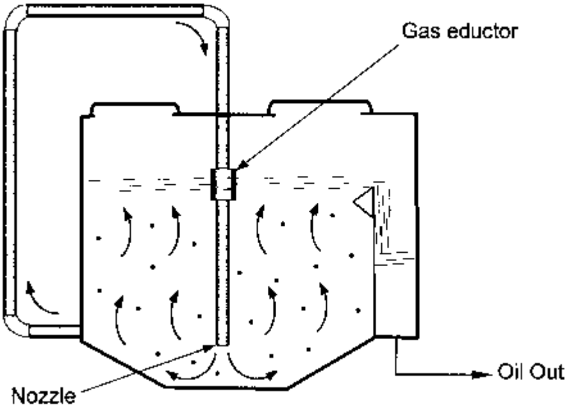

Figure 6 Dispersed gas flotation unit with eduction.

cubic feet (SCF) ranges from 0.2 to 0.5 SCF/bbl of water to be treated. The gas-saturated water is recycled back into the unit, which operates at a pressure lower than that of the gas–water contactor. Due to the reduction in pressure, the dissolved gas breaks out of solution as small bubbles. The gas bubbles carry the oil droplets with them as they move to the surface. The size and depth of the unit are determined to provide retention times between 10 and 40 min. The equipment manufacturer normally determines the detailed design parameters of the unit based on the specific operating conditions.

Dispersed Gas Flotation Units

In this type of flotation unit, the gas bubbles are created, introduced, and dispersed into the bulk of the water to be treated. This is basically done by two methods. In one method, the gas bubbles are created and dispersed in the water by inducing a vortex using a mechanical rotor driven by an electric motor. Figure 7 shows a schematic cross section of a unit utilizing this method and manufactured by Petrolite Corporation.

The vortex induced by the rotor creates vacuum within the vortex tube. Due to this vacuum, gas is withdrawn into the vortex and is dispersed in the water. The gas bubbles carry the oil droplets as froth to the surface, where the oil is skimmed and collected in the recovery channel for removal out of the unit.

The other method of creating and dispersing the gas bubbles utilizes an inductor device as shown in Figure 8. As shown in the figure, a portion

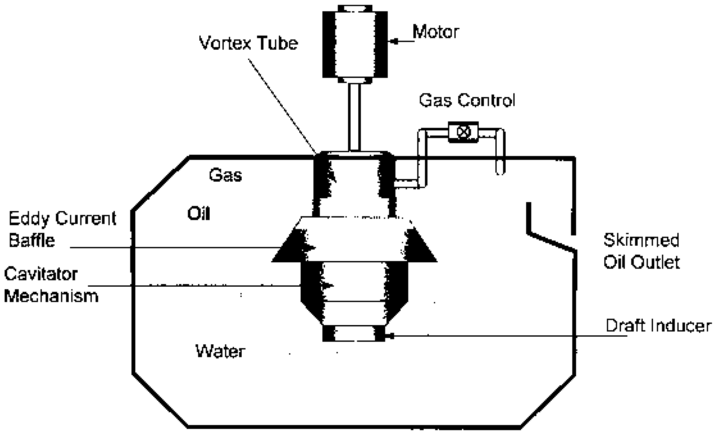

Figure 7 Unit utilizing the method of creating gas bubbles and dispersing them in the water by inducing a vortex.

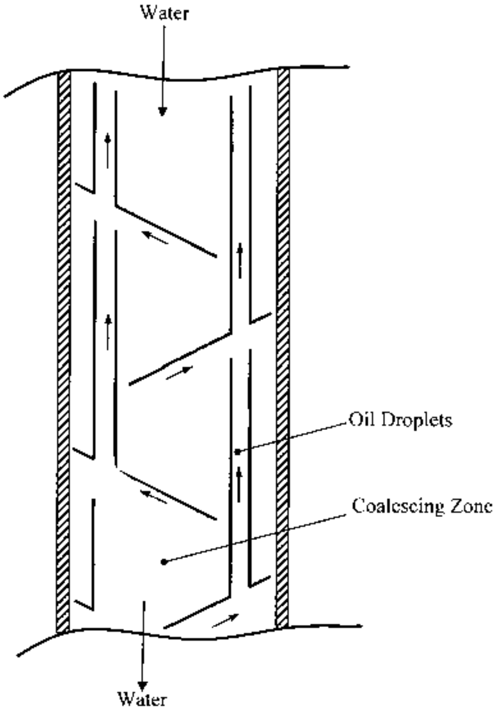

Figure 8 Unit utilizing the method of creating and dispersing gas bubbles using an inductor device.

of the treated water is recycled back to the unit using a pump. The recycled water flows through a venturi and, due to the reduction in pressure, sucks gas from the vapor space at the top of the unit. The gas is released through a nozzle near the bottom in the form of small bubbles that carry the oil droplets to the surface as they rise. Finally, the oil is skimmed and collected in a chamber for removal out of the unit.

Normally, a dispersed gas flotation unit consists of three or four of the cells described. The water to be treated moves from one cell to the next for further removal of oil. Typically, the oil removal efficiency of one cell is about 50%. Therefore, a three-cell unit will have an overall efficiency of 87%, whereas a four-cell unit will have an overall efficiency of 94%. Flotation unit manufacturers have patented design and produce standard units that are typically designed to handle produced water flow rate of about 5000 BPD. For higher flow rates, additional units are added in parallel. Flotation units are capable of removing oil droplets smaller than 30 μm.

9.5 OFFSHORE WATER DISPOSAL EQUIPMENT

Produced water in offshore operations should not be dumped directly into the sea after treatment. In addition to the treated produced water, rainwater, and equipment-washdown water represent another source of oil-contaminated water that needs to be disposed of properly. For this purpose, offshore production platforms should be equipped with some form of a disposal device that disposes of the water deep enough below the surface of the sea and away from the wave action to prevent sheens from occurring. The most common of these disposal devices are the *disposal piles*, *skim piles*, and *SP piles*; these are described in the following subsections.

Disposal Piles

Disposal piles are the simplest form of offshore water disposal devices. The disposal pile is simply a large diameter open-ended pipe that is attached to the platform and extends to a specific minimum depth below the surface of the sea. The diameter of the pile is determined based on the total flow of water to be disposed of and the water and oil gravities. In shallow water, the disposal pile should extend down to near the seafloor. In deep water, however, the depth of the pile below the normal water level is determined such that a high level in the pile will be sensed and the shutdown signal measured before the oil in the pile comes within 10 ft of the bottom.

Disposal piles are used to collect treated produced water, deck drains, treated sand, and liquids from drip pans and dispose of them deep below the surface. Disposal piles are also useful as traps for oil in the event of equipment failure or upset operating conditions. The deck drainage, normally rainwater and washdown water, is saturated with oxygen and may contain sand and other solids. Therefore, it should not be treated in the same equipment as produced water to avoid corrosion and plugging problems. Disposal piles are particularly useful for disposal of the platform drainage.

Skim Piles

The skim pile is basically a disposal pile equipped with a series of inclined baffle plates and oil collection risers, as shown in Figure 9. The presence of these baffles plates serves two functions. It reduces the distance a given oil droplet has to rise to be separated from the water and creates zones of no flow below each plate. The oil droplets rise to the zone of no flow between two successive plates, where coalescence and gravity separation occurs. The coalesced large oil droplets travel up the bottom side of the plate and into the oil collection riser to the surface of the pile where oil could be skimmed out.

Skim piles have two specific advantages over standard disposal piles. Skim piles are more efficient in separating oil from water. Skim piles also provide for some degree of cleaning sand that may be present in the water from oil.

SP Piles

In this type of device, the disposal pile is equipped with a number of equally separated SP packs and oil risers. As water flows through a SP pack, coalescence of oil droplets occurs due to the induced turbulence. As the water travels out of the SP pack to the next SP pack, the larger oil droplets rise to form an oil pad below the upper SP pack. This continues as the water goes from one SP pack to the next. The oil accumulated below the bottom SP pack rises to the oil pad above through the risers until it reaches the surface to be pumped out.

The SP packs are normally designed to develop oil droplets to a maximum size of 750 μm. The number of SP packs needed is determined

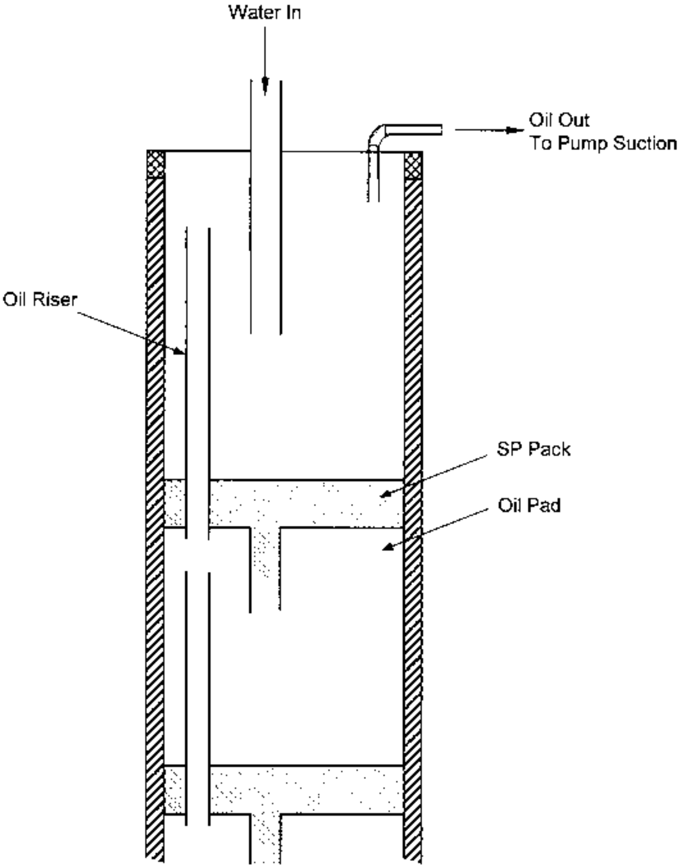

Figure 9 Schematic of a skin pile.

from the desired overall efficiency of oil removal and the calculated efficiency of a single pack will be explained later.

9.6 DESIGN OF WATER TREATMENT AND DISPOSAL EQUIPMENT

To properly design the produced water system, certain data and information must be available. In some cases, needed data could be obtained from actual measurements and analyses. In other cases, however, the engineer may have to assume reasonable values for missing or unattainable data. The following data and information are required for the design of the treatment system:

1. Flow rate of the produced water, Q_{pw} (BPD)
2. Specific gravity, γ_{pw}, and viscosity, μ_{pw}, of the produced water
3. Concentration of oil in the produced water (in mg/L or ppm)
4. Oil droplets size distribution in the produced water
5. Specific gravity of oil at flowing conditions, γ_o
6. Required effluent (treated water) quality (in mg/L or ppm)
7. Concentration of soluble oil in the effluent
8. Rainfall rate, Q_{rw}, and flow rate of washdown water, Q_{wd}

It should be noted that the equipment discussed in this chapter could not separate the soluble oil. The treating equipment are designed to separate dispersed oil only. Therefore, the soluble oil concentration should be subtracted from the allowable oil concentration of the effluent to obtain the allowable dispersed oil concentration used for design purposes. This is used to determine the size of oil droplets that must be removed from the water, which greatly affects the size and selection of the treating equipment.

In the following subsections the equations used to determine the size of the various treating equipment are first presented, then a guideline of equipment selection procedure is highlighted. Throughout the derivation of the equations, the following notations and units are used:

Q_{pw} = produced water flow rate (BPD)
Q_{rw} = deck's rainwater flow rate (BPD)
Q_{wd} = deck's washdown water flow rate (BPD)
μ_{pw} = produced water viscosity (cP)
μ = produced water viscosity (lb-s/ft^2)
d_m = smallest diameter of oil droplets to be removed (μm)
t_{rw} = retention time for water (min)
t_w = retention time for water (s)
t_o = settling time for oil droplets (s)
V_w = average water velocity through the skimmer (ft/s)
V_o = settling velocity of oil droplets (ft/s)
D = diameter of cylindrical skimmer (in.)
L = effective length of skimmer (ft)
L_s = total (seam-to-seam) length of skimmer (ft)
H = height of water in the skimmer (ft)
W = width of rectangular cross-section skimmer (ft)
A = flow area for water (ft^2)

h = perpendicular distance between plates of CPI and PPI (in.)
θ = angle of inclination of plate interceptors with the horizontal (deg)

9.6.1 Sizing Horizontal Cylindrical Skimmers

The dimensions of the skimmer must be sufficient to allow for separation of the smallest oil droplet that need to be separated and for the desired retention time. Let D be the skimmer diameter (in in.) and L_s be the skimmer length (in ft). We shall assume that the skimmer is half full with water, which moves horizontally from the inlet to the outlet of the skimmer. Due to the presence of various internal components, the effective length for separation, L, will be less than L_s. To be on the conservative side, we shall also assume that the oil droplets will settle vertically upward from the bottom of the skimmer to the surface of the water; that is, the oil droplets will travel a distance equal to the radius of the skimmer to reach the surface of the water. The time it takes the oil droplet to rise to the surface, t_o, must equal the time it takes the water to move from the inlet to the outlet of the skimmer, t_w.

From Stokes' law, the terminal rise velocity of the oil droplet, u_o, is given by

$$u = 1.787 \times 10^{-6} \frac{(\gamma_w - \gamma_o) d_m}{\mu_{\text{pw}}} \ \frac{\text{ft}}{\text{s}} \tag{1}$$

where d_m is the diameter of the oil droplet (in μm). Therefore,

$$t_o = \frac{(D/2 \times 12)}{u_o} \ \text{s}$$

$$t_o = 2.34 \times 10^4 \frac{D \mu_{\text{pw}}}{(\gamma_w - \gamma_o) d_m} \ \text{s} \tag{2}$$

The average velocity of the water, u_w, is obtained by dividing the flow rate by the flow area; that is,

$$u_w = \frac{[5.61 Q_{\text{pw}}/(24 \times 3600)]}{[0.5(\pi/4)(D/12)^2]^{-1}} \ \frac{\text{ft}}{\text{s}}$$

The water retention time is obtained by dividing the effective length, L, by the average velocity; therefore,

$$t_w = 41.9786 \frac{D^2 L}{Q_{\text{pw}}} \ \text{s} \tag{3}$$

Equating Eqs. (2) and (3), we get

$$DL = 557.43 \frac{Q_{\mathrm{pw}} \mu_{\mathrm{pw}}}{(\gamma_w - \gamma_o) d_m} \text{ in. ft} \tag{4}$$

For turbulence and short-circuiting, an efficiency factor of 1.8 is recommended; therefore, Eq. (4) becomes

$$DL = 1000 \frac{Q_{\mathrm{pw}} \mu_{\mathrm{pw}}}{(\gamma_w - \gamma_o) d_m} \text{ in. ft} \tag{5}$$

Equation (5) provides a relationship between the diameter D and effective length L of the skimmer which allows all oil droplets of diameter d_m or larger to rise to the surface of the water. Therefore, any combination of D and L that satisfies Eq. (5) is theoretically acceptable.

Another constraint on the dimensions of the skimmer is that the size should be large enough to provide the necessary retention time, t_{rw}. The retention time is obtained by dividing the volume of the skimmer occupied by water, V_{pw}, by the water flow rate, that is,

$$t_{\mathrm{rw}} = \frac{V_{\mathrm{pw}}}{Q_{\mathrm{pw}}} \tag{6}$$

where

$$V_{\mathrm{pw}} = 0.5 \left(\frac{\pi}{4}\right) \left(\frac{D}{12}\right)^2 L \text{ ft}^3$$

$$Q_{\mathrm{pw}} \left(\frac{ft^3}{\min}\right) = Q_{\mathrm{pw}}\,(\mathrm{BPD}) \left(\frac{5.61}{24 \times 60}\right)$$

Substituting in Eq. (6) and rearranging, we obtain

$$D^2 L = 1.43 t_{\mathrm{rw}} Q_{\mathrm{pw}} \text{ in.}^2 \text{ ft} \tag{7}$$

Therefore, the combination of diameter and effective length that satisfies both settling and retention time constraints should be selected. This is obtained by plotting D versus L for both Eqs. (5) and (7) and determining the values of D and L that satisfy both equations. The actual length of the skimmer, L_s, is then determined from

$$L_s = 1.33 \mathrm{L} \tag{8}$$

9.6.2 Sizing Horizontal Rectangular Cross Section (API) Skimmer

The same procedure described in Section 9.6.1 is used to determine the width, W, effective length, L, and height of the water, H, of the skim tank. In this case, we have

$$t_o = \frac{H}{u_o} \tag{9}$$

$$t_w = \frac{L}{u_{\mathrm{pw}}} \tag{10}$$

Equating Eqs. (9) and (10) with u_o given by Eq. (1) and u_{pw} given by dividing the water flow rate $[5.61Q_{\mathrm{pw}}/(24 \times 3600)]$ by the flow area (HW), we obtain

$$WL = 36.5\frac{Q_{\mathrm{pw}}\mu_{\mathrm{pw}}}{(\gamma_w - \gamma_o)d_m}\ \mathrm{ft}^2 \tag{11}$$

For this type of skimmer, the recommended efficiency factor for turbulence and short-circuiting is 1.9; therefore, Eq. (11) becomes

$$WL = 70\frac{Q_{\mathrm{pw}}\mu_{\mathrm{pw}}}{(\gamma_w - \gamma_o)d_m}\ \mathrm{ft}^2 \tag{12}$$

The retention time as obtained from Eq. (10) should also be equal to the volume of the skimmer occupied by water (HWL) divided by the water flow rate; therefore,

$$t_{\mathrm{rw}} = 15401\frac{HWL}{Q_{\mathrm{pw}}}\ \mathrm{s} \tag{13}$$

The height of the water, H, is usually limited to one-half the width, W; therefore, substituting $0.5\,W$ for H in Eq. (13) and dividing by 60, the retention time in minutes is given by

$$t_{\mathrm{rw}} = 125\frac{W^2L}{Q_{\mathrm{pw}}}\ \mathrm{min} \tag{14}$$

Rearranging, we obtain

$$W^2L = 0.008t_{\mathrm{rw}}Q_{\mathrm{pw}}\ \mathrm{ft}^3 \tag{15}$$

Any combination of W and L that satisfies both Eqs. (12) and (15) satisfies both the settling and retention time constraints.

9.6.3 Sizing Vertical Cylindrical Skimmer

In this type of skimmer, the oil droplets move vertically upward against the vertially downward flowing water. For settling to occur, the velocity of the oil droplet, u_o, must be at least equal to the average velocity of the water, u_{pw}, determined by dividing the water flow rate by the circular cross-sectional area of the skimmer; that is,

$$u_w = \frac{5.61 Q_{\text{pw}}/(24 \times 3600)}{(\pi/4)(D/12)^2} \ \frac{\text{ft}}{\text{sec}}$$

Equating u_w to u_o expressed by Eq. (1) and solving for the diameter of the skimmer, D, we obtain

$$D^2 = 6691 \frac{Q_{\text{pw}} \mu_w}{(\gamma_w - \gamma_o) d_m} \ \text{in.}^2 \tag{16}$$

Equation (16) is good for diameters up to 48 in. For larger diameters, where turbulence and short-circuiting occur, Eq. (16) should be multiplied by a factor F that is greater than 1. The value of F depends largely on the design of the inlet and outlet flow spreaders, oil collector, and baffles.

The retention time is obtained by dividing the height of the water in the skimmer, H, by the average flow velocity of the water:

$$t_{\text{rw}} = H \left(\frac{5.61 Q_{\text{pw}}/(24 \times 3600)}{(\pi/4)(D/12)^2} \right)^{-1} \ \text{s} \tag{17}$$

Convert t_{rw} into minutes in Eq. (17) and solve for the height H:

$$H = 0.7 \frac{t_{\text{rw}} Q_{\text{pw}}}{D^2} \ \text{ft} \tag{18}$$

Equation (16) determines the minimum diameter of the skimmer that satisfies the settling constraint. Any diameter that is equal to or larger than that determined from Eq. (16) is acceptable. Equation (18) determines the height of the water for the selected diameter to satisfy the retention time constraint. The total height (seam-to-seam length) of the skimmer, L_s, is obtained by adding 3 ft to H; that is,

$$L_s = H + 3 \ \text{ft} \tag{19}$$

Example 1

A field separation and treatment plant produces 8000 BPD of water containing 2000 mg of oil per liter of water with a maximum oil droplet size of 500 μm. The water viscosity and specific gravity are 1.1 cP and 1.07, respectively, and the oil specific gravity is 0.87. It is required to treat the produced water to reduce it oil content to 800 mg/L.

Taking the retention time as 10 min, determine the dimensions of (1) horizontal cylindrical skimmer, (2) horizontal, rectangular cross-section skimmer, and (3) vertical cylindrical skimmer to provide the required treatment.

Solution

The first step is to determine the size of the oil droplet that must be removed to achieve the required treated water quality. Because the relationship between the oil droplet size and concentration is approximately linear, the droplet diameter, d_m, that must be removed is obtained from

$$d_m = 500\left(\frac{800}{2000}\right) = 200\,\mu\text{m}$$

1. **Horizontal Cylindrical Skimmer.** The dimensions of the skimmer must satisfy both settling and retention time constraints, Eqs. (5) and (7), respectively. Using Eq. (5),

$$DL = \frac{(1000)(8000)(1.1)}{(1.07 - 0.87)(200)^2}$$

$$DL = 1100 \tag{E1}$$

Using Eq. (7),

$$D^2L = (1.43)(8000)(10) = 114{,}400 \tag{E2}$$

Assuming various values for D and determining the corresponding values of L from Eqs. (E1) and (E2) the following table is constructed:

	Settling Eq. (E1)		Retention time [Eq.(E2)]	
D (in.)	L(ft)	L_s(ft)	L(ft)	L_s(ft)
60	18.33	24.44	31.78	42.26
72	15.28	20.36	22.07	29.35
84	13.10	17.46	16.21	21.56
96	11.46	15.27	12.41	16.51
108	10.19	13.58	9.81	13.05
120	9.17	12.22	7.94	10.57

Examination of the table shows that the retention time constraint governs the design up to a diameter of 96 in.; then, the settling constraint governs. **Suitable selections would be 84 in. × 22 ft, or 96 in. × 17 ft.**

2. **Horizontal Recatangular Cross-Section Skimmer.** Using Eq. (12) (settling constraint),

$$WL = \frac{(70)(8000)(1.1)}{0.2(200)^2} = 77\ \text{ft}^2 \tag{E3}$$

Using the retention time constraint, Eq. (14)

$$W^2L = (0.008)(8000)(10) = 640\ \text{ft}^3 \tag{E4}$$

		L(ft)	L(ft)
W(ft)	$H_w = 0.5\,W$	Settling [Eq. (E3)]	Retention [Eq. (E4)]
5	2.5	15.40	26.6
6	3.0	12.83	17.28
7	3.5	11.00	13.06

Again, the retention time constraint governs the design of the skimmer. **The suitable selections are skimmers that are 6 ft wide by 18 ft long**, **and 7 ft wide by 14 ft long.** The height of the water in the skimmer is limited to one-half of the width.

3. **Vertical Cylindrical Skimmer.** The settling constraint [Eq. (16)] provides the minimum diameter for the skimmer:

$$D^2 = \frac{(6691)(8000)(1.1)}{0.2(200)^2} = 7360$$

$$D_{\text{min}} = 85.79\ \text{in.}$$

Because the diameter is greater than 48 in., the settling equation should be multiplied by a factor that is greater than 1 to account for turbulence and short-circuiting. Using a factor of 1.5 gives

$$D_{\text{min}} = 105\ \text{in.}$$

Taking a diameter of 108 in. and substituting in Eq. (18) (the retention-time constraint), the height of the water is determined:

$$H = \frac{(0.7)(8000)(10)}{(108)^2} = 4.8 \text{ ft}$$

The height of the skimmer (seam-to-seam length), L_s, is given by

$$L_s = H + 3 = 4.8 + 3 = 7.8 \text{ ft}$$

Therefore, **the skimmer is 108 in. in diameter and 8 ft high.**

Any of the three skimmers will provide the required water treatment. The final selection depends on cost, availability, and space limitation.

9.6.4 Sizing of Skimmers with PPI

In this case, the oil droplets need only to rise to the underside of plates in a time t_o that is equal to the time it takes the water to travel through the effective length of the skimmer, t_w. We shall assume that the effective length in the settling process is 70% of the actual length of the PPI pack and that the plates occupy 10% of the flow area (HW).

$$t_o = \frac{(h/12\cos\theta)}{u_o} \tag{20}$$

where h is defined as in the notation on page 234. Substituting for u_o from Eq. (1)

$$t_o = 4.681\frac{h\mu_{\text{pw}}}{(\gamma_w - \gamma_o)d_m\cos\theta} \text{ s} \tag{21}$$

The retention time of water is given by

$$t_w = \frac{L}{u_w} = 0.7\frac{L_s}{u_w} \tag{22}$$

where

$$u_w = \frac{5.61Q_{\text{pw}}/(24 \times 3600)}{0.9HW}\ \frac{\text{ft}}{\text{s}} \tag{23}$$

Therefore,

$$t_w = 9.703 \times 10^3\frac{LHW}{Q_{\text{pw}}} \text{ s} \tag{24}$$

Equating Eqs. (21) and (24),

$$HWL = 4.824\frac{Q_{\text{pw}}h\mu_{\text{pw}}}{(\gamma_w - \gamma_o)d_m\cos\theta} \tag{25}$$

For efficient settling of the oil droplets, it has been reported that the Reynolds number (Re) with the characteristic dimension being four times the hydraulic radius should not exceed 1600. This can be used to develop an equation for the minimum flow area (HW) as follows. The hydraulic radius, R, is obtained by dividing the area between plates by the wetted perimeter, that is,

$$R = \frac{(h/12)(W/\cos\theta)}{2((W/\cos\theta) + (h/12))}$$

$$R = \frac{(h/12)}{2(1 + (h\cos\theta/12W))}$$

Considering that $(h\cos\theta/12) \ll W$, R may be approximated by

$$R = \frac{h}{24}$$

Therefore,

$$\text{Re} = \frac{\rho_w D u_w}{\mu g} = 1600$$

where g is the acceleration of gravity, ft/sec^2. Substituting for u_w from Eq. (23), $D = 4R$, $\rho_w = 62.4\,\gamma_w$ lb/ft^3, and $\mu = 2.088 \times 10^{-5}\,\mu_{\text{pw}}$ and solving for HW, we obtain

$$HW = 6.982 \times 10^{-4}\frac{Q_{\text{pw}}h\gamma_w}{\mu_{\text{pw}}}\ \text{ft}^2 \tag{26}$$

Equation (26) is first used to determine the minimum value of HW, then Eq. (25) is used to determine L for any value of HW above the minimum determined.

9.6.5 Sizing of Skimmers with CPI

The corrugated plate interceptor packs are normally available in standard sizes with $H = 3.25$ ft, $W = 3.25$ ft, $L = 5.75$ ft, $h = 0.69$ in., and $\theta = 45°$. The size of the skimmer used with the CPI is determined by the number of CPI packs installed. It should be noted that the flow rate through the standard CPI pack should not exceed 20,000 BPD in order to stay within the Reynolds number limitation.

Using Eq. (25) and the dimensions of standard CPI packs, the number of CPI packs needed, N, is determined using

$$N = 0.11 \frac{Q_{\mathrm{pw}} \mu_{\mathrm{pw}}}{(\gamma_w - \gamma_o) d_m} \tag{27}$$

9.6.6 Sizing of Skimmers with Cross-Flow Devices

As described earlier, cross-flow devices require the installation of internal spreaders and collectors for uniform distribution of the water among the plates. Because spreaders do not guarantee a 100% uniform distribution of the flow, a spreader efficiency must be introduced in sizing skimmers with cross-flow devices. It has been suggested to use a spreader efficiency of 75% for this purpose. Equation (25) could be used for sizing after dividing the right-hand side by 0.75; therefore,

$$HWL = 6.432 \frac{Q_{\mathrm{pw}} h \mu_{\mathrm{pw}}}{(\gamma_w - \gamma_o) d_m \cos\theta} \tag{28}$$

9.6.7 Determining the Reqired Number of SP Packs

As previously described, the SP packs are used to coalesce the oil droplets and greatly increase their size to speed the separation process. Depending on the treatment requirements, more than one stage may be needed. In such a case, the stages will be arranged in series. The number of SP pack stages, n, can be determined as follows. Let C_i be the oil concentration in the influent water and C_o be the maximum allowed oil concentration in the effluent (treated) water. Therefore, the overall separation or treatment efficiency, E, is determined from

$$E = 1 - \frac{C_o}{C_i} \tag{29}$$

Alternatively, because the droplet size distribution is approximately linear, E could be determined from

$$E = 1 - \frac{d_o}{d_i} \tag{30}$$

where d_o is the largest oil droplet diameter in the effluent and d_i is the largest oil droplet diameter in the influent water.

Now, let d_m be the oil droplet diameter that can be separated by a skimmer of a certain size. The function of the SP pack is to coalesce such oil droplets to increase the diameter to $d_{\max}$ (typically, $d_{\max}$ for standard

SP packs is approximately 1000 μm). The efficiency of the SP pack could then be determined from

$$E_{\mathrm{SP}} = 1 - \frac{d_m}{d_{\max}} \tag{31}$$

Therefore, for n SP packs staged in series, the overall efficiency could be expressed in terms of the efficiency of the SP pack by

$$E = 1 - (1 - E_{\mathrm{SP}})^n \tag{32}$$

9.6.8 Sizing of Flotation Units

Flotation units manufacturers have various standard units with specific dimensions and flow rate limitations. The role of the facilities engineer is, therefore, limited to determining whether flotation units are needed for the treatment process and the number of units needed. The need for flotation units depends on the required quality of the treated (effluent) water. Normally, if the maximum allowable oil droplet size in the effluent water is about 30 μm or less, flotation units should be considered. The efficiency of the flotation unit, which is typically between 87% and 94% depending on the number of cells, is then used to determine the quality (oil concentration, or maximum oil droplet size) of the influent water that could be treated by the flotation unit from Eq. (29) or (30). The quality of the influent water is then compared to the quality of the produced water to determine whether a primary treatment stage is needed.

Example 2

For the same conditions of Example 1, design the required treatment facilities such that the treated water does not contain more than 80 mg of oil per liter of water.

Solution

The maximum oil droplet size that can exist in the treated water is determined from

$$d_m = \frac{(80)(500)}{2000} = 20\,\mu\mathrm{m}$$

This size is too small to be separated by skimmers of any reasonable size.

Therefore, either flotation units or SP packs should be used. Let us examine both options.

1. **Using flotation units.** Flotation units are normally available as standard units for a flow rate of 5000 BPD. Because we have 8000 BPD of water to be treated, two units must be used in parallel. Assume that a four-cell unit is being used and that the efficiency of the unit is 90%. The quality of the water that could be treated by the flotation unit is, therefore, determined from

   $$E = 1 - \left(\frac{d_{mo}}{d_{mi}}\right)$$

   $$0.9 = 1 - \left(\frac{20}{d_{mi}}\right)$$

   $$d_{mi} = 200\,\mu\text{m}$$

   Because this is smaller than the droplet size in the produced water, a primary treatment of the water is necessary before it can be treated in the flotation unit. The primary treatment is required to remove all oil droplets of 200 μm and larger. This is the same as the requirement for Example 1. Therefore, any of the three skimmers designed in the previous example could be used. The treatment system could, therefore, consist of a horizontal cylindrical skimmer 96 in. × 17 ft, or a vertical skimmer of 108 in. diameter by 8 ft high, and two flotation units in parallel.
2. **Using SP packs.** The overall required efficiency is first calculated from

   $$E = \frac{2000 - 80}{2000} = 0.96$$

   Assume that we will use the vertical skimmer designed in Example 1. The droplet size that can be treated in such a skimmer is 200 μm. Therefore, the efficiency of the SP pack (assuming it grows 1000-μm droplets) is

   $$E_{\text{SP}} = 1 - \left(\frac{200}{1000}\right) = 0.8$$

   From Eq. (32),

   $$0.96 = 1 - (1 - 0.8)^n$$
   $$n = 2$$

Therefore, the treatment could be achieved using two 108-in.-diameter vertical skimmers with SP packs.

The selection of any of the above two options will depend on cost and other considerations such as the availability of gas for flotation units.

9.6.9 Sizing of Offshore Disposal Devices

Disposal Piles

As mentioned earlier, disposal piles are not designed for produced water treatment; they are simply used to dispose of the treated produced water. However, disposal piles are also used to dispose of the deck drains, normally rainwater or washdown water. Such water is usually contaminated by oil and, therefore, the disposal pile will be expected to provide the means to settle oil droplets of certain size. In most cases, the disposal pile will be designed to separate oil droplets 150 μm and larger. This, however, depends on local regulations.

Equation (16) developed for vertical skimmers is used to determine the diameter of the disposal pile; the equation is rewritten as follows:

$$D^2 = 6691 \frac{Q_w \mu_w}{(\gamma_w - \gamma_o)} \text{ in.}^2 \tag{33}$$

where Q_w is given by

$$Q_w = Q_{\text{pw}} + Q_{\text{rw}} + Q_{\text{wd}} \text{ BPD} \tag{34}$$

Normally, the larger of the rainwater and washdown water flow rates is used in Eq. (34), as it is unlikely to have both at the same time. The water viscosity, μ_w, is taken as 1 cP.

The length of the pile submerged below the normal sea level is determined such that the high-level and shutdown alarms are sensed before the oil reaches 10 ft from the bottom. This could be determined from a hydrostatic pressure balance at the bottom of the pile. The normal tide and expected storms should be considered.

Let L_{w} be the normal water level, L_{st} be the design storm range, L_t be the average tide range, and assigning 4 ft for the alarm and shutdown levels, the length of the pile submerged, L, will be given by

$$L = 10 + \frac{(4 + L_{\text{st}} + L_t + L_w)\gamma_o}{\gamma_w - \gamma_o} \text{ ft} \tag{35}$$

In deep waters, however, the minimum length should be 50 ft. In shallow water, the length of the pile should extend as deep as possible.

Skim Piles

Due to the complexity of the flow regime in skim piles, no analytical equation exists for sizing such piles. However, an adequate and simple equation that is based on field experience, developed in this subsection, may be used for determining the size of skim pile.

Let L be the length of the submerged section of the pile as determined from Eq. (35). The length of the baffle section, L_b, is then determined from

$$L_b = L - 15 \text{ ft} \tag{36}$$

If d is the diameter of the pile (in in.) and assuming that the baffles and oil collection risers occupy 25% of the volume, the volume occupied by water, v, is then given by

$$V = 0.75\left(\frac{\pi D^2}{4 \times 1144}\right) L_b \text{ ft}^3 \tag{37}$$

The retention time in the baffle section, t_r, is obtained by dividing the volume, V, by the water flow rate, Q_w, which is determined from Eq. (34); therefore,

$$t_r = \frac{V}{5.61 Q_w/(24 \times 60)} \text{ min} \tag{38}$$

Experience has shown that a retention time of 20 min is sufficient in meeting disposal requirements. Therefore, combining Eqs. (37) and (38) assigning a value of 20 to t_r, and solving for $D^2 L_b$, we get

$$D^2 L_b = 19.1 Q_w \text{ in.}^2\text{ ft} \tag{39}$$

SP Piles

The procedure for determining the number of SP packs needed is the similar to that used for skimmers. A pile diameter is first selected and the efficiency per SP stage is determined. The number of stages is then determined from Eq. (32) for an overall efficiency of 90%.

Example 3

An offshore production platform having a deck area of 3000 ft^2 produces 4500 BPD of water. Primary and secondary treatment stages are used and result in a final water quality of 45 mg oil per litre of water, which meets the offshore disposal requirements. The average rainfall in the area is 1.5 in./h and the deck is equipped with a 50-gpm washdown hose.

Design the necessary disposal facility to handle the treated produced water and deck drain. Assume a water viscosity of 1.1 cP and a difference in specific gravity of 0.2.

Solution

The first step is to calculate the total amount of water to be disposed of:

$$\text{Rainwater} = 1.5\,\frac{\text{in.}}{\text{h}} \times \frac{1}{12}\,\frac{\text{ft}}{\text{in.}} \times 3000\ \text{ft}^2 \times 24\,\frac{\text{h}}{\text{day}} \times \frac{1}{5.61}\,\frac{\text{bbl}}{\text{ft}^3} = 1604\ \text{BPD}$$

$$\text{Washdown water} = 50\ \text{gpm} \times (60 \times 24)\,\frac{\text{min}}{\text{day}} \times \frac{1}{48}\,\frac{\text{bbl}}{\text{gal}} = 1500\ \text{BPD}$$

Because the rainwater rate is larger than the washdown rate, it will be considered in the design. Therefore,

$$Q_w = 4500 + 1604 = 6104\ \text{BPD}$$

We first check the simplest disposal device (i.e., the disposal pile). Using Eq. (33) with $d_m = 150\ \mu\text{m}$,

$$D^2 = \frac{(6691)(6104)(1.1)}{0.2(150)^2} = 9983.57$$

$$D = 99.9\ \text{in.}$$

This is too large. Therefore, we check the skim pile. Using Eq. (39),

$$D^2 L_b = (19.1)(6104) = 116{,}586\ \text{in.}^2\ \text{ft}$$

Assume various values for D and calculate the corresponding values of L_b:

D(in.)	36	42	48
L_b(ft)	89.9	66.1	50.6

Therefore, we can use a 48-in.-diameter pile with a 50-ft baffle section. The total submerged length is, therefore Eq. (36),

$$L = Lb + 15$$

or

$$L = 50 + 15 = 65\ \text{ft}$$

which is greater than the minimum required length of 50 ft.

The skim pile is an adequate choice. The SP pile could also be used where the treated produced water should be introduced below the SP packs because it does not need any treatment. The skim pile is, however, simpler and less expensive.

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REVIEW QUESTIONS AND EXERCISE PROBLEMS

1. Discuss the reasons for treating produced water that will be
    (a) Disposed off into the sea
    (b) Disposed off into an underground formation
    (c) Used as injection water for improved oil recovery
    (d) Used for generating steam for thermal recovery operations
2. Discuss the advantages and disadvantages of treating produced water under pressure versus atmospheric treatment.
3. Describe the functions of the following equipment in promoting efficient produced water treatment:
    (a) Parallel plates interceptors
    (b) Corrugated plate interceptors
    (c) Serpentine-pipe coalescers
4. Describe the basis of operation of flotation units and discuss the reasons for using such units.
5. Describe the function and basis of operation of the various types of offshore disposal pile.
6. Discuss the effects of the various parameters in the settling equation [Eq. (1)] on the efficiency of separating oil droplets from produced water. Discuss possible means of controlling these parameters to improve the separation (treatment) process.
7. Design the water treatment and disposal facilities for an offshore production and treatment platform for the conditions listed. Perform the design calculations for all types of treatment and disposal equipment discussed in this chapter and then determine the best combination of facilities giving reasons for your decisions.

Produced water rate	12,000 BPD
Oil content	2000 mg oil/L water
Maximum oil droplet size	500 μm
Required treated water quality	70 mg/L
Dissolved oil	25 mg/L
Water specific gravity	1.074
Oil specific gravity	0.89
Water viscosity	1.03 cP
Treating temperature	70°F
Platform area	4500 ft^2
Average annual rainfall rate	1.7 in./h
Available washdown hose	50 gpm

10

Overview of Gas Field Processing

10.1 PLANNING THE SYSTEM

Part IV of the book is devoted to field treatment and processing operations of natural gas and other associated products. These include dehydration, acidic gas removal (H_2S and CO_2), and the separation and fractionation of liquid hydrocarbons [known as natural gas liquid (NGL)]. Sweetening of natural gas almost always precedes dehydration and other gas plant processes carried out for the separation of NGL. Dehydration, on the other hand, is usually required before the gas can be sold for pipeline marketing and it is a necessary step in the recovery of NGL from natural gas.

For convenience, a system involving field treatment of a gas project could be divided into two main stages, as shown in Figure 1:

1. Stage I, known as gas treatment or gas conditioning
2. Stage II, known as gas processing

The gas treatment operations carried out in stage I involve the removal of gas contaminants (acidic gases), followed by the separation of water vapor (dehydration), as explained in Chapters 11 and 12, respectively. Gas processing, stage II, on the other hand, comprises two operations: NGL recovery and separation from the bulk of gas and its subsequent fractionation into desired products. The purpose of a fractionator's facility is simply to produce individual finished streams needed for market sales. Fractionation facilities play a significant role in gas plants, as given in Chapter 13.

Gas field processing in general is carried out for two main objectives:

1. The necessity to remove impurities from the gas
2. The desirability to increase liquid recovery above that obtained by conventional gas processing

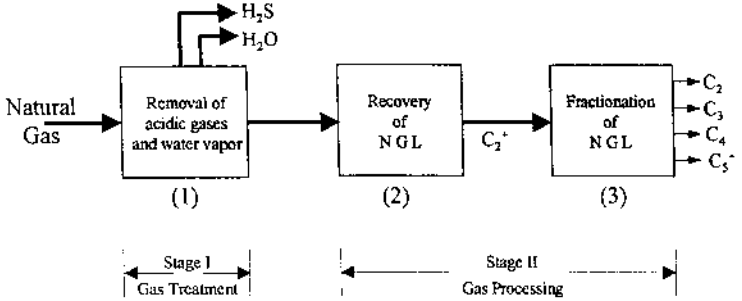

Figure 1 Operations involved in the treatment and processing of natural gas.

Natural gas field processing and the removal of various components from it tend to involve the most complex and expensive processes. Natural gas leaving the field can have several components which will require removal before the gas can be sold to a pipeline gas transmission company. All of the H_2S and most of the water vapor, CO_2, and N_2 must be removed from the gas. Gas compression is often required during these various processing steps. To illustrate this point, a sour gas leaving a gas–oil separation plant (GOSP) might require first the use of an amine unit (MEA) to remove the acidic gases, a glycol unit (TEG) to dehydrate it, and a gas compressor to compress it before it can be sold.

It is also generally desirable to recover NGL present in the gas in appreciable quantities. This normally includes the hydrocarbons known as C_3^+. In some cases, ethane C_2 could be separated and sold as a petrochemical feed stock. NGL recovery is the first operation in stage II, as explained earlier. To recover and separate NGL from a bulk of a gas stream would require a change in phase; that is, a new phase has to be developed for separation to take place by using one of the following:

1. An energy-separating agent; examples are refrigeration for partial or total liquefaction and fractionation
2. A mass-separating agent; examples are adsorption and absorption (using selective hydrocarbons, 100–180 molecular weight).

The second operation in stage II is concerned with the fractionation of NGL product into specific cuts such as LPG (C_3/C_4) and natural gasoline.

It should be pointed out that the fact that all of the field processes do not occur at or in the vicinity of the production operation does not change the plan of the system of gas processing and separation or its basic needs. This requires superimposing the fractionation facilities to the gas field processing scheme.

Most important in the design of a system for gas field processing, the following parameters should be evaluated and considered in the study:

1. Estimated gas reserve (both associated and free)
2. The gas flow rate and composition of the feed gas
3. Market demand, both local and export, for the products
4. Geographic location and methods of shipping of finished products
5. Environmental factors
6. Risks involved in implementing the project and its economics

Of these factors, the gas/oil reserve might be the paramount factor. Several schemes can be recommended for field processing and separation of natural gas, but the specific solution is found to be a function of the composition of the gas stream, the location of this source, and the markets available for the products obtained.

10.2 BACKGROUND

10.2.1 What Is Natural Gas?

Natural gas is the gas obtained from natural underground reservoirs either as free gas or gas associated with crude oil. It generally contains large amounts of methane (CH_4) along with decreasing amounts of other hydrocarbons. Impurities such as H_2S, N_2, and CO_2 are often found with the gas. It also generally comes saturated with water vapor. Other information on natural gas and its usage under different commercial products are found in Table 1 in the Introduction.

10.2.2 Why Field Processing?

The principal market for natural gas is achieved via transmission lines, which distribute it to different consuming centers, such as industrial, commercial, and domestic.

Field processing operations are thus enforced to treat the natural gas in order to meet the requirements and specifications set by the gas transmission companies. The main objective is to simply obtain the natural gas as a main product free from impurities. In addition, it should be recognized that field processing units are economically justified by the increased liquid product (NGL) recovery above that obtained by conventional separation.

10.2.3 Types of Gas Reservoir

It is of interest in this chapter to comment briefly on the types of gas reservoir that supply us with natural gas. It is misleading to infer that few well-known types can rigidly describe reservoirs producing natural gas.

At one end, some fields produce saturated associated gas (gas associated with crude oil); on the other end, a dry gas (free gas) is produced from some fields. In between these two ends, one can find numerous types of reservoir in which the hydrocarbons vary in composition and, hence, the gas produced. Some of the factors contributing to these changes are as follows:

1. The contents of heavier components
2. The percentage of acidic gases
3. The presence of inert gases

The type of processing operations recommended for any specific operation will, consequently, depend on the kind of gas under consideration along with the distribution of the hydrocarbons found in the gas. For example, natural gas separated via a GOSP to be transmitted for distribution imposes specific duties on the field processing operations that are not justified if the gas were to be diverted for gas injection application in oil recovery.

For discussion purposes, Figure 2 illustrates some diversified processing operations involved in the treatment of natural gas produced by different reservoirs.

10.2.4 Gas Specifications and Standard Conditions

Market sales of natural gas require some specifications set by the consumers regarding the maximum contents allowable for the following: acidic gases and sulfur, oxygen and carbon dioxide, water vapor, and liquefiable hydrocarbons. The thermal heating content of the gas sets another value to be met as a minimum. These specifications are given in the Introduction of the book.

As for the standard conditions used in the gas industry for reporting the volumetric properties of the gas, two common standard conditions of temperature and pressure are used:

1. Universal scientific, reported at 32°F and 760 mm Hg
2. Natural gas industry, reported at 60°F and 14.7 psia

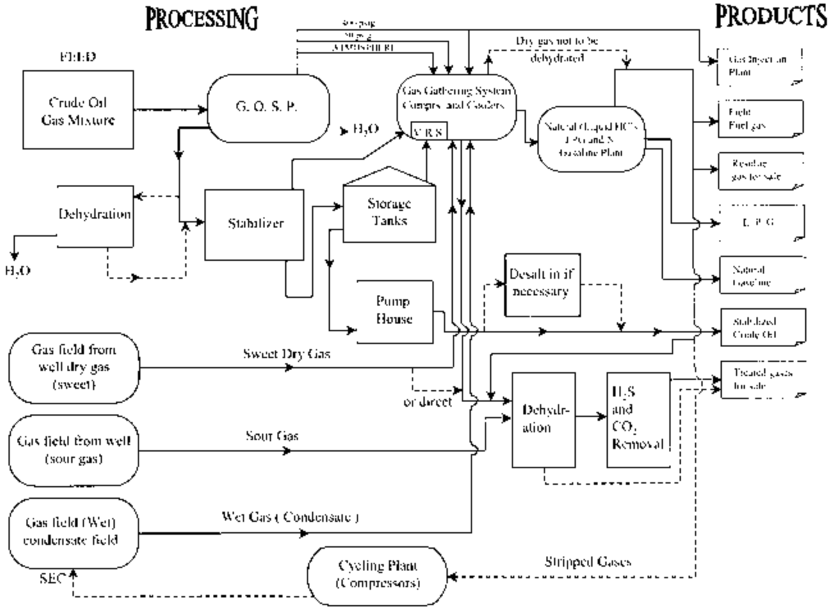

Figure 2 Schematic presentation of processing operations.

Under the universal standard conditions, for an ideal gas, the following is applied:

1 g mole occupies 22.4 liters
1 kg mole occupies 22.4 cubic meters
1 lb mole occupies 359 cubic feet

10.3 EFFECT OF IMPURITIES FOUND IN NATURAL GAS

Field processing operations of natural gas, which is classified as a part of gas engineering, generally include the following:

1. Removal of water vapor, dehydration
2. Removal of acidic gases (H_2S and CO_2)
3. Separation of heavy hydrocarbons

Before these processes are detailed in the following chapters, the effect each of these impurities has on the gas industry, as end user, is briefly outlined:

Water vapor	H_2S and CO_2	Liquid hydrocarbons
It is a common impurity. It is not objectionable as such. (a) Liquid water accelerates corrosion in the presence of H_2S gas. (b) Solid hydrates, made up of water and hydrocarbons, plug valves, fittings in pipelines, and so forth.	Both gases are harmful, especially H_2S, which is toxic if burned; it gives SO_2 and SO_3 which are nuisance to consumers. • Both gases are corrosive in the presence of water. • CO_2 contributes a lower heating value to the gas.	Their presence is undesirable in the gas used as a fuel. • The liquid form is objectionable for burners designed for gas fuels. • For pipelines, it is a serious problem to handle two-phase flow: liquid and gas.

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11

Sour Gas Treating

11.1 INTRODUCTION

Natural gas usually contains some impurities such as hydrogen sulfide (H_2S), carbon dioxide (CO_2), water vapor (H_2O), and heavy hydrocarbons such as mercaptans. These compounds are known as "acid gases." Natural gas with H_2S or other sulfur compounds (such as COS, CS_2 and mercaptans) is called "sour gas," whereas gas with only CO_2 is called "sweet gas." It is usually desirable to remove both H_2S and CO_2 to prevent corrosion problems and to increase heating value of the gas.

Sweetening of natural gas is one of the most important steps in gas processing for the following reasons:

1. **Health hazards.** At 0.13 ppm, H_2S can be sensed by smell. At 4.6 ppm, the smell is quite noticeable. As the concentration increases beyond 200 ppm, the sense of smell fatigues, and the gas can no longer be detected by odor. At 500 ppm, breathing problems are observed and death can be expected in minutes. At 1000 ppm, death occurs immediately.
2. **Sales contracts.** Three of the most important natural gas pipeline specification are related to sulfur content, as shown in Table 1. Such contracts depend on negotiations, but they are quite strict about H_2S content.
3. **Corrosion problems.** If the partial pressure of CO_2 exceeds 15 psia, inhibitors usually can only be used to prevent corrosion. The partial pressure of CO_2 depends on the mole fraction of CO_2 in the gas and the natural gas pressure. Corrosion rates will also depend on temperature. Special metallurgy should be used if CO_2 partial pressure exceeds 15 psia. The presence of H_2S will cause metal embrittlement due to the stresses formed around metal sulfides formed.

Table 1 Natural Gas Pipeline Specifications

Characteristic	Specification	Test method
Water content	4–7 lb/MMSCF maximum	ASTM (1986), D 1142
Hydrogen sulfide content	0.25 grain/100 SCF maximum	GPA (1968), Std. 2265 GPA (1986), Std. 2377
Gross heating value	950 Btu/SCF minimum	GPA (1986), Std. 2172
Hydrocarbon dew point	15°F at 800 psig maximum	ASTM (1986), D 1142
Mercaptan content	0.2 grain/100 SCF maximum	GPA (1968), Std. 2265
Total sulfur content	1–5 grain/100 SCF maximum	ASTM (1980), D 1072
Carbon dioxide content	1–3 mol% maximum	GPA (1990), Std. 2261
Oxygen content	0–0.4 mol% maximum	GPA (1990), Std. 2261
Sand, dust, gums, and free liquid	Commercially free	
Delivery temperature (°F)	120°F maximum	
Delivery pressure (psia)	700 psig minimum	

MMSCF ≡ 1,000,000 standard cubic feet.

11.2 GAS-SWEETENING PROCESSES

There are more than 30 processes for natural gas sweetening. The most important of these processes can be classified as follows:

1. **Batch solid bed absorption.** For complete removal of H_2S at low concentrations, the following materials can be used: iron sponge, molecular sieve, and zinc oxide. If the reactants are discarded, then this method is suitable for removing a small amount of sulfur when gas flow rate is low and/or H_2S concentration is also low.
2. **Reactive solvents.** MEA (monoethanol amine), DEA (diethanol amine), DGA (diglycol amine), DIPA (di-isopropanol amine), hot potassium carbonate, and mixed solvents. These solutions are used to remove large amounts of H_2S and CO_2 and the solvents are regenerated.
3. **Physical solvents.** Selexol, Recitisol, Purisol, and Fluor solvent. They are mostly used to remove CO_2 and are regenerated.
4. **Direct oxidation to sulfur.** Stretford, Sulferox LOCAT, and Claus. These processes eliminate H_2S emissions.
5. **Membranes.** This is used for very high CO_2 concentrations. AVIR, Air Products, Cynara (Dow), DuPont, Grace, International Permeation, and Monsanto are some of these processes.

11.3 SELECTION OF SWEETENING PROCESS

There are many factors to be considered in the selection of a given sweetening process. These include the following:

1. Type of impurities to be removed (H_2S, mercaptans)
2. Inlet and outlet acid gas concentrations
3. Gas flow rate, temperature, and pressure
4. Feasibility of sulfur recovery
5. Acid gas selectivity required
6. Presence of heavy aromatic in the gas
7. Well location
8. Environmental consideration
9. Relative economics

Figure 1 can be utilized for the initial selection of the proper process. It depends on the sulfur content in the feed and the desired product. Several commercial processes are available, as shown in the schematic flowsheet of Figure 2. This diagram can help in selecting the suitable process. If the sulfur recovery is not an option, then indirect processes can be considered. These processes can be classified into liquid-phase processes and dry-bed processes. The latter is selected if the sulfur content in the feed is low, as supported by Figure 1. In this case, the total sulfur to be removed should not be large. The dry-bed processes can be classified as iron oxide sponge or zinc oxide process, in this case, because the oxides are not expensive, the corresponding sulfide might not be regenerated on site but rather by a contracting company, and this might lower the operational cost and does not require further processing for sulfur recovery. Alternatively, the molecular sieve bed can be used when the bed is regenerated on site.

If the operating conditions require the removal of a large amount of H_2S and/or CO_2, liquid-phase processes are used. If it is required to selectively remove H_2S, then physical solvents will be suitable, they also remove COS and CS_2. If the feed contain high amount of heavy hydrocarbons (C_3^+), then the use of physical solvents may result in a significant loss of these heavy hydrocarbons because they are released from the solvent with the acid gases and cannot be recovered economically. On the other hand, if the feed contains a high amount of H_2S and CO_2 and it is required to remove both of them, then chemical solvents such as amines or carbonates could be used. The amine processes offer good reactivity at low cost and good flexibility in design and operation. However, carbonates can be used if it is required to remove COS and CS_2 from the feed. The carbonate process also can be run at a lower utility cost.

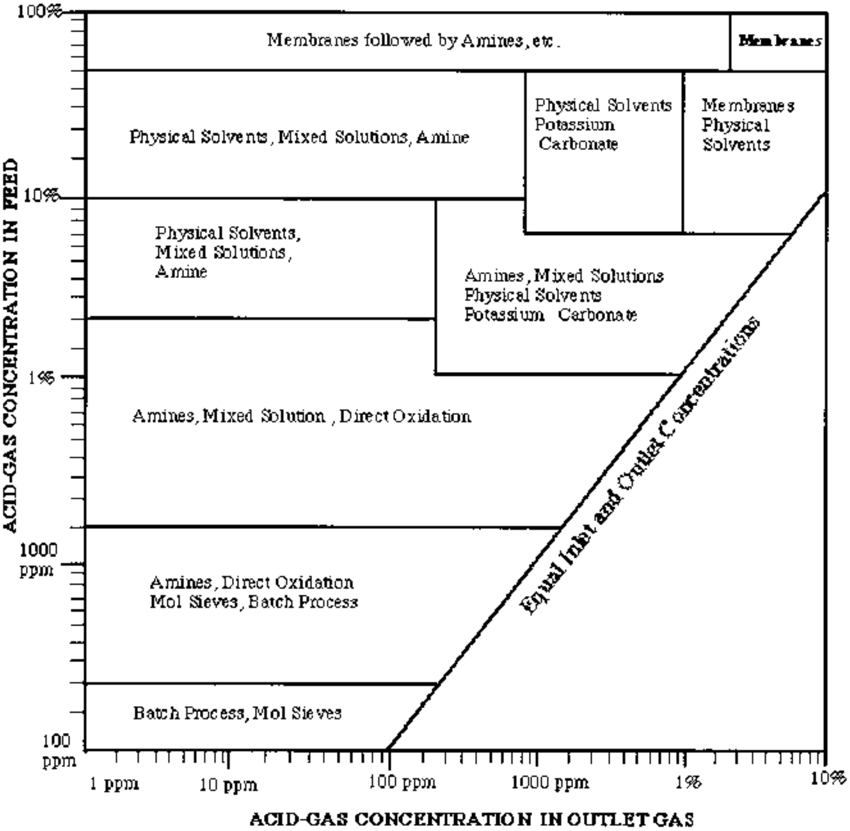

Figure 1 Selection of gas-sweetening processes.

If the sulfur recovery is an option, then a direct conversion process can be used. They are only selective for H_2S. Stretford, Lo Cat, and Sulferox are some of the process used for removal of H_2S from natural gas. In the case of the regeneration of solid fixed beds or amine regeneration units, concentrated streams of H_2S are obtained. In these cases, the Claus process will be the most suitable for sulfur recovery.

Figure 1 can be used as a guide for the selection of the proper process depending on the acid–gas concentration in the feed and the required degree of acid removal and its concentration in the outlet stream, after Tennyson and Schaff [1]. However, final selection should consider environmental, economical, and any other considerations.

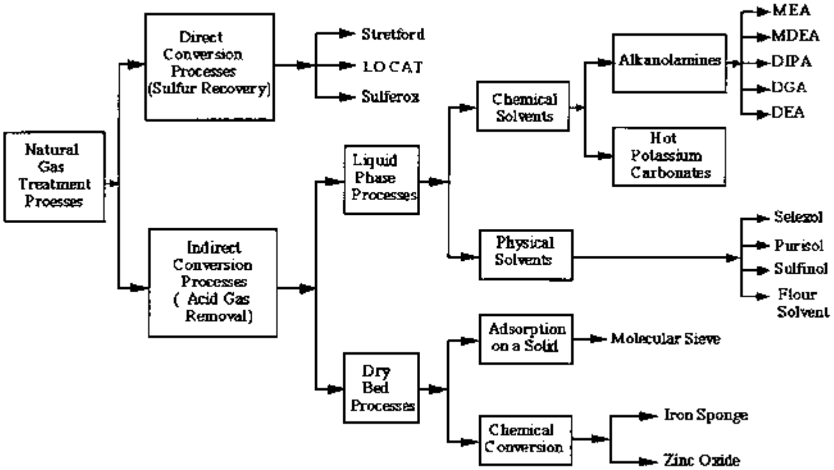

Figure 2 Alternatives for natural gas sweetening.

11.4 BATCH PROCESSES

In this case, H_2S is basically removed and the presence of CO_2 does not affect the processes. Usually, batch processes are used for low-sulfur-content feeds.

11.4.1 Iron Sponge

Iron sponge fixed-bed chemical absorption is the most widely used batch process. This process is applied to sour gases with low H_2S concentrations (300 ppm) operating at low to moderate pressures (50–500 psig). Carbon dioxide is not removed by this treatment. The inlet gas is fed at the top of the fixed-bed reactor filled with hydrated iron oxide and wood chips. The basic reaction is the formation of ferric sulfide when H_2S reacts with ferric oxide:

$$2Fe_2O_3 + 6H_2S \rightarrow 2Fe_2S_3 + 6H_2O \tag{1}$$

The reaction requires an alkalinity pH level 8–10 with controlled injection of water. The bed is regenerated by controlled oxidation as

$$2Fe_2S_3 + 3O_2 \rightarrow 2Fe_2O_3 + 6S \tag{2}$$

Some of the sulfur produced might cake in the bed and oxygen should be introduced slowly to oxide this sulfur, Arnold and Stewart [2]:

$$S_2 + 2O_2 \rightarrow 2SO_2 \tag{3}$$

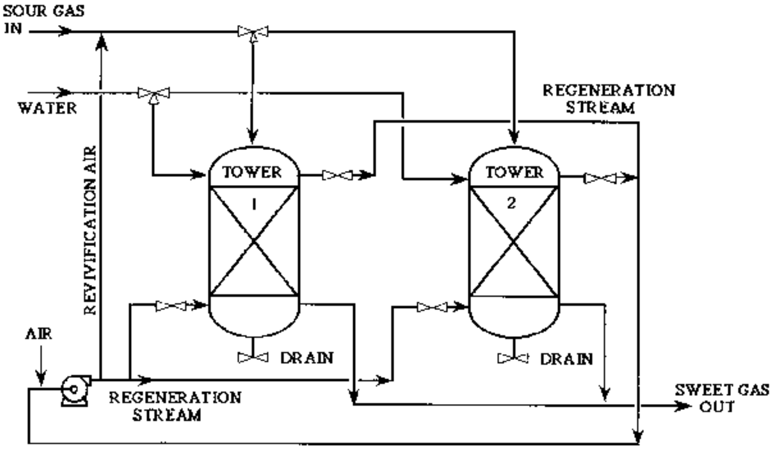

Figure 3 Typical iron oxide process flowsheet.

Repeated cycling of the process will deactivate the iron oxide and the bed should be changed after 10 cycles.

The process can be run continuously, in this case, small amounts of air or oxygen are continuously added to the inlet sour gas so that the produced sulfur is oxidized as it forms. The advantage of this process is the large savings in labor cost for loading and unloading of the batch process. In this case, higher sulfur recovery per pound of iron oxide is obtained also.

A typical flow diagram of high-pressure continuous operation of the iron oxide process is shown in Figure 3. In this case, one of the towers is on stream removing H_2S from the sour gas while the second tower is in the regeneration cycle by air blowing. The last regeneration step should be carried out with caution because reaction (2) is highly exothermic and the rate of reaction must be controlled. Care must be taken in replacing the exhausted iron sponge beds from the two towers shown in Figure 2. On opening the beds, entering air causes a sharp rise in temperature which can result in spontaneous combustion of the bed. The entire bed should be wetted before recharging. There are only two types of ferric oxide which can be used for gas sweetening: α-Fe_2O_3–H_2O and γ-Fe_2O_3–H_2O; they react easily with H_2S and can be regenerated easily by oxidation to the corresponding ferric oxide forms.

In some cases, it is more economical to operate the process with a single bed and the exhausted bed of iron sulfide is trucked away to a disposal site. The tower is then recharged with a new bed of iron oxide and the tower is put back to service. The spent bed is left to react slowly with oxygen in the air, as shown in reactions (2) and (3), which can be done by a contracting company.

Design of Iron Sponge Bed

It is noted from Figure 1 that such a bed (batch process) can handle an acid gas feed of up to 200 ppm and bring it down to 1 ppm in the exit sweet gas stream.

The design of the iron sponge bed requires the following data:

Q_g(MMSCFD) = Gas flow rate
P(psig) = Operating pressure
T(°F) = Operating temperature
X_{AG}(ppm) = H_2S inlet concentration
Z = Compressibility factor from charts
SG = Specific gravity of the gas (assume 0.7 if not given)

Calculation Procedure

1. Calculate column diameter. First the minimum diameter is calculated as

$$d_m^2(\text{in.}^2) = 360\frac{Q_g TZ}{P}$$

Then, the maximum diameter to prevent channeling is

$$d_{\max}^2(\text{in.}^2) = 1800\frac{Q_g TZ}{P}$$

The selected diameter should be between these two values.

2. Calculate daily consumption of iron sponge (R, ft^3/day):

$$R = 0.0133 Q_g (X_{\mathrm{AG}})$$

3. Select bed height between 10 and 20 ft. The bed volume (V) becomes

$$V(\text{ft})^3 = 0.7854\, D^2 L$$

where D and L are in feet. The bed is replaced after V/R days.

Example 1

Natural gas flowing at 2 MMSCFD has a specific gravity of 0.7 and acid gas concentration of 25 ppm (MMSCFD = 10^6 standard cubic feet per day). The operating pressure and temperature are 1000 psig and 120°F, respectively. Design an iron sponge unit.

Solution

The feed acid concentration is 25 ppm, which is below 200 ppm, so a batch process such as iron sponge unit can be used.

$$d^2_{\text{min}} = \frac{360(2)(580)(0.86)}{1014.7} = 353.43 \text{ and } d_{\text{min}} = 18.8\,\text{inch}$$

$$d^2_{\text{max}} = \frac{1800(2)(580)(0.86)}{1014.7} = 1769.7 \text{ and } d_{\text{max}} = 42\,\text{inch}$$

A diameter between 18.8 and 42 in. can be used. Select a diameter of 24 in. (2 ft). Daily iron consumption R is

$$R = 0.0133(2)(25) = 0.665\,\text{ft}^3/\text{day}$$

Select a bed height of 10 ft. The bed volume becomes

$$V = 0.7854(2)^2(10) = 31.416\,\text{ft}^3$$

$$\text{Bed life} = \frac{31.416}{0.665} = 47\,\text{days}$$

Thus, the bed should be replaced in 6 weeks.

11.4.2 Zinc Oxide

Zinc oxide can be used instead of iron oxide for the removal of H_2S, COS, CS_2, and mercaptans. However, this material is a better sorbent and the exit H_2S concentration can be as low as 1 ppm at a temperature of about 300°C. The zinc oxide reacts with H_2S to form water and zinc sulfide:

$$ZnO + H_2S \rightarrow ZnS + H_2O$$

A major drawback of zinc oxide is that it is not possible to regenerate it to zinc oxide on site, because active surface diminishes appreciably by sintering. Much of the mechanical strength of the solid bed is lost due to fines formation, resulting in a high-pressure-drop operation.

The process has been decreasing in use due to the above problems and the difficulty of disposing of zinc sulfide; Zn is considered a heavy metal.

11.4.3 Molecular Sieves

Molecular sieves (MSs) are crystalline sodium alumino silicates and have very large surface areas and a very narrow range of pore sizes. They possess highly localized polar charges on their surface that act as adsorption sites for

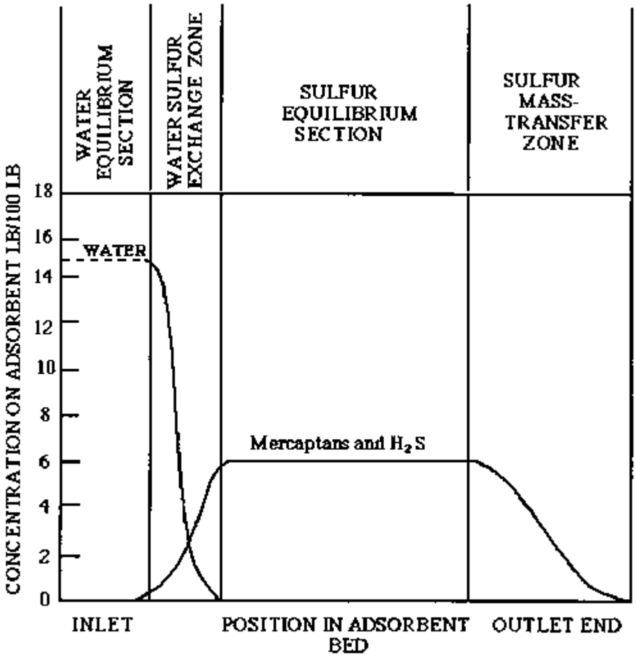

Figure 4 Adsorption zone in a molecular sieve bed.

polar materials at even very low concentrations. This is why the treated natural gas could have very low H_2S concentrations (4 ppm).

In order for a molecule to be adsorbed, it first must be passed through a pore opening and then it is adsorbed on an active site inside the pore. There are four major zones in a sieve bed as shown in Figure 4. In the presence of water, which is a highly polar compound, H_2O is first adsorbed in the bed, displacing any sulfur compounds. Water equilibrium with the sieves is established in this zone. The second zone (zone 2) is the water–sulfur compound exchange zone, where water is still displacing the sulfur compounds but some sulfur sites will be left adsorbed. An adsorption front for sulfur compounds is formed and a concentration profile is established. The concentration profile of sulfur compounds declines along the bed. Zone 3 is the sulfur equilibrium section, showing the highest concentration of sulfur compounds in the bed depending on the MS capacity. Zone 4 is the mass transfer section for sulfur compounds. As this profile reaches the end of the bed, this will mean some sulfur compounds (H_2S) will appear in the gas stream

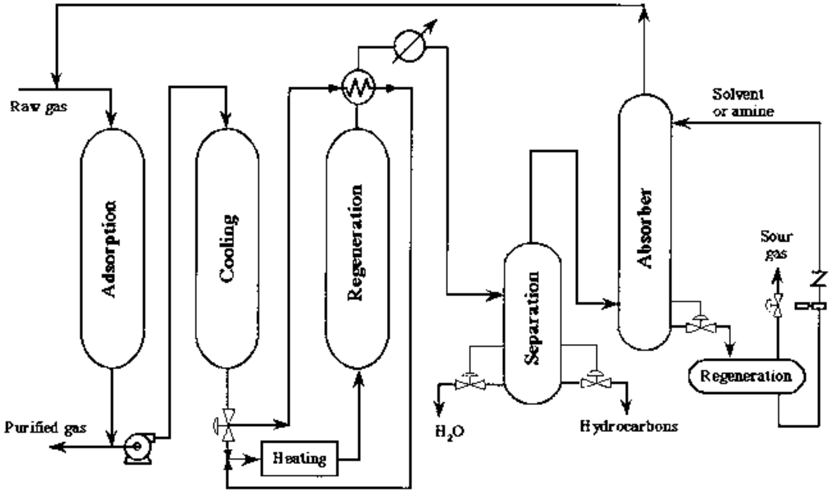

Figure 5 Sweetening of natural gas by molecular sieves.

(unabsorbed), this will mean that the bed is exhausted and requires regeneration.

If it is desired to remove H_2S, a MS of 5 A* is selected, however, if it is also desired to remove mercaptans, 13 X* is selected. In either case, selection made to minimize the catalytic reaction

$$H_2S + CO_2 \Leftrightarrow COS + H_2O$$

Olefins, aromatics, and glycols are strongly adsorbed, which may poison the sieves.

Commercial applications require at least two beds so that one is always on line while the other is being regenerated. The schematic diagram of the process is shown in Figure 5. The sulfur compounds are adsorbed on a cool, regenerated bed in the sweetening. The saturated bed is regenerated by passing a portion of the sweetened gas, preheated to about 400–600°F or more, for about 1.5 h to heat the bed. As the temperature of the bed increases, it releases the adsorbed H_2S into the generation gas stream. The sour effluent gas is flared off, with about 1–2% of the treated gas lost.

An amine unit can be added to this process to recover this loss; in this case, H_2S will be flared off from the regenerator of the amine unit. In case this flaring is prohibited environmentally, the H_2S can be sent to a gathering center for the sulfur recovery unit, if it exists on site.

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*5A and 13 X are the basic types of molecular sieves commercially available. Their basic properties and applications are fully described by Campbell [3].

11.5 LIQUID-PHASE PROCESSES

This is one of the most commonly used processes for acid gas treatment. Chemical solvents are used in the form of aqueous solution to react with H_2S and CO_2 reversibly and form products which can be regenerated by a change of temperature or pressure or both. Physical solvents can be utilized to selectively remove sulfur compounds. They are regenerated at ambient temperature by reducing the pressure. A combination of physical and chemical solvents can be used. A comparison of chemical solvents (amines, carbonates) and physical solvents is shown in Table 2.

Table 2 Comparison of Chemical and Physical Processes

	Chemical adsorption		Physical absorption
Feature	Amine processes	Carbonate processes	
Absorbents	MEA, DEA, DGA, MDEA	K_2CO_3, K_2CO_3 + MEA K_2CO_3+DEA K_2CO_3+arsenic trioxide	Selexol, Purisol, Rectisol
Operating pressure (psi)	Insensitive to pressure	> 200	250–1000
Operating temperature (°F)	100–400	200–250	Ambient temperature
Recovery of absorbents	Reboiled stripping	Stripping	Flashing, reboiled, or stream stripping
Utility cost	High	Medium	Low/medium
Selectivity H_2S, CO_2	Selective for some amines (MDEA)	May be selective	Selective for H_2S
Effect of O_2 in the feed	Formation of degradation products	None	Sulfur precipitation at low temperature
COS and CS_2 removal	MEA: not removed DEA: slightly removed DGA: removed	Converted to CO_2 and H_2S and removed	Removed
Operating problems	Solution degradation; foaming; corrosion	Column instability; erosion; corrosion	Absorption of heavy hydrocarbons

11.5.1 Amine Processes

The most widely used for sweetening of natural gas are aqueous solutions of alkanolamines. They are generally used for bulk removal of CO_2 and H_2S. The properties of several amines are shown in Table 2. The low operating cost and flexibility of tailoring solvent composition to suit gas compositions make this process one of most commonly selected. A liquid physical solvent can be added to the amine to improve selectivity.

A typical amine process is shown in Figure 6. The acid gas is fed into a scrubber to remove entrained water and liquid hydrocarbons. The gas then enters the bottom of absorption tower which is either a tray (for high flow rates) or packed (for lower flow rate). The sweet gas exits at the top of tower.

The regenerated amine (lean amine) enters at the top of this tower and the two streams are contacted countercurrently. In this tower, CO_2 and H_2S are absorbed with the chemical reaction into the amine phase. The exit amine solution, loaded with CO_2 and H_2S, is called rich amine. This stream is flashed, filtered, and then fed to the top of a stripper to recover the amine, and acid gases (CO_2 and H_2S) are stripped and exit at the top of the tower. The refluxed water helps in steam stripping the rich amine solution. The regenerated amine (lean amine) is recycled back to the top of the absorption tower.

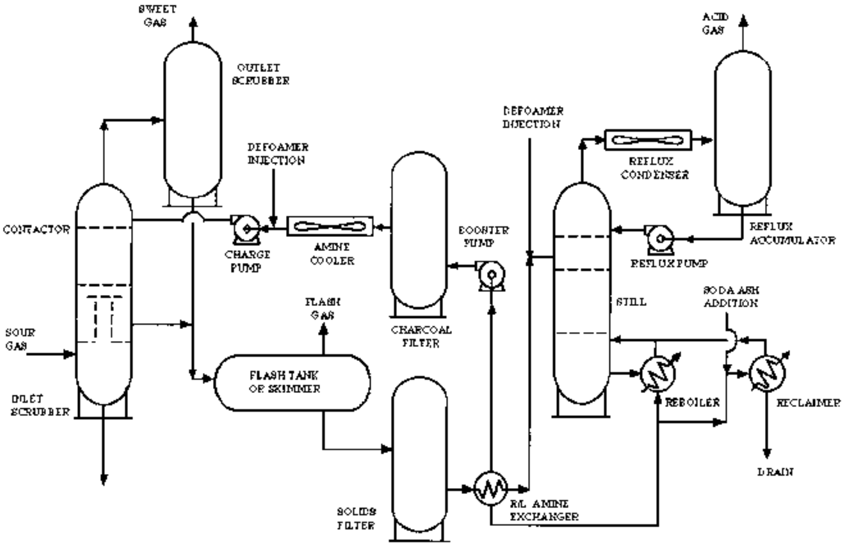


Figure 6 Flowsheet for the amine process.

Table 3 Comparison of Amine Solvents

Solvent	MEA	DEA	DIPA	DGA	MDEA
Chemical formula	RNH_2	R_2NH	R'_2NH	$RO(CH_2)_2NH_2$	R_2CH_3N
Molecular weight	61	105	133	105	119
Amine type	Primary	Secondary	Secondary	Primary	Tertiary
Vapor pressure, 100°F (mm Hg)	1.05	0.058	0.01	0.16	0.0061
Freezing point (°F)	15	20	16	– 40	– 25
Relative capacity (%)	100	58	46	58	51
Solution weight (%)	20	30	35	60	50
K[a]	2.05	1.45	0.95	1.28	1.25
Loading (mol AG/mol amine)	0.35	0.5	0.7	0.3	0.4
H_2S/CO_2 selectivity[b]	1	1	2	1	3
Solvent conc. (wt%)	15–20	20–35	30–40	45–65	40–55
AG (mol/mol)	0.3–0.4	0.5–0.6	0.3–0.4	0.3–0.4	0.3–0.45
Circulation (gal/mol AG)	100–165	60–125	—	50–75	65–110
Steam rate (lb/gal)	1.0–1.2	0.9–1.1	—	1.1–1.3	0.9–1.1
Reboiler temp. (°F)	240	245	255	260	250
Heat of reaction (Btu/lb AG, H_2S)	620	550	—	675	500
CO_2	660	630	0	850	600

[a] Circulation rate (gpm) $= KQ_g$ (MMSCFD) X_{AG} (mol% acid gas) for $P > 400$ psig and $T < 120$°F.

[b] 10 is highly selective; 1 is not selective.

The operating conditions of the process depends on the type of the amine used. Some of these conditions are given in Table 3. Primary amines are the strongest to react with acid gases; but the stable bonds formed make it difficult to recover by stripping. Secondary amines have a reasonable capacity for acid gas absorption and are easily recovered. Tertiary amines have a lower capacity, but they are more selective for H_2S absorption.

Among the amines discussed here, DEA is the most common. This is may be due to the fact it is less expensive to install and operate. Specific details for each amine follow.

Monoethanolamine Solvent

Monoethanolamine (MEA) is a primary amine and the strongest amine among others. It can produce pipeline specification gas. It reacts with H_2S

and CO_2 as follows:

$$2(RNH_2) + H_2S \Leftrightarrow (RNH_3)_2S \tag{4}$$

$$(RNH_3)_2S + H_2S \Leftrightarrow 2(RNH_3)HS \tag{5}$$

$$2(RNH_2) + CO_2 \Leftrightarrow RNHCOONH_3R \tag{6}$$

These reactions are reversible and are forward in the absorber (at low temperature) and backward in the stripper (at high temperature).

Monoethanolamine will react with H_2S and CO_2 nonselectively. Unfortunately, it will react irreversibly with carbonyl sulfide and carbon disulfide, which can result in a loss of solution and a buildup of solid products in the circulating solution. These products could also be a source of corrosion.

The MEA process is usually using a solution of 15–20% MEA (wt%) in water. Loading is about 0.3–0.4 mol of acid removed per mole of MEA. The circulation rate is between 2 and 3 mol of MEA per mole of H_2S, as expected from reactions (4) and (5). However, commercial plants use a ratio of 3 to avoid excessive corrosion. Solution strength and loading above this limits result in excessive corrosion. MEA forms foam easily due to the presence of contaminants in the liquid phases; this foam results in carryover from the absorber. These contaminants could be condensed hydrocarbons, degradation products, iron sulfide, as well as corrosion products and excess inhibitors. Solids can be removed by using a filter; hydrocarbons could be flashed; degradation products are removed using a reclaimer. The number of trays used in absorbers in commercial units is between 20 and 25 trays. However, the theoretical number of trays calculated from published equilibrium data is about three to four. If we assume an efficiency of 35% for each tray, then the actual number of trays is 12. It has been reported that the first 10 trays pick up all of the H_2S and at least another 10 trays are of not much value. Thus, it is suggested to use 15 trays. It is recommended that MEA be used if the feed does not contain COS or CS_2, which form stable products and deplete the amine. If the feed has these compounds, a reclaimer must be used, where a strong base like NaOH is used to regenerate and liberate the amine. This base has to be neutralized later.

Diethanolamine Process

Diethanolamine (DEA) is replacing MEA and becoming the most widely used gas sweetening solvent. It is a secondary amine with lower reactivity and corrosivity than MEA. Moreover, it reacts with COS and CS_2 and the product can be regenerated. It has a lower vapor pressure (lower losses) and lower heat of reaction (easier to generate) than MEA. The basic reactions with CO_2 and H_2S are the same as MEA:

$$2R_2NH + H_2S \Leftrightarrow (R_2NH_2)_2S \tag{7}$$

$$(R_2NH_2)_2S + H_2S \Leftrightarrow 2R_2NH_2SH \tag{8}$$

$$2R_2NH + CO_2 \Leftrightarrow R_2NCOONH_2R_2 \tag{9}$$

Based on these reactions, 1.7 lbs of DEA can be circulated to react with the same amount of acid gas as 1.0 lb of MEA. A higher strength of 35% (wt) can be used because of its lower corrosivity. Loading up to 0.65 mol of acid gas per mole of DEA can cause fewer operational problems than MEA because the elimination of the degradation products and the absence of a reclaimer. Corrosion is less because DEA is weaker than MEA. Foaming is reduced probably due to absence of degradation and corrosion products.

Di-isopropanolamine Process

Di-isopropanolamine (DIPA) is a secondary amine and is used most frequently in the ADIP process licensed by Shell. DIPA reacts with COS and CS_2 and the products are easily regenerated. At low pressure, DIPA is more selective to H_2S, and at higher pressures, DIPA removes both CO_2 and H_2S. It is noncorrosive and requires less heat for rich amine regeneration.

Methyldiethanolamine

Methyldiethanolamine (MDEA) is commonly used in the 20–50% (wt) range. Lower weight percent solutions are typically used in very low-pressure and selectivity applications. Acid gas loading as high as 0.7–0.8 mol acid gas/mol amine are practical in carbon steel equipment. Higher loading may be possible with few problems. Corrosion is much reduced in this case even under these high loadings. In the presence of oxygen, MDEA forms corrosive acids which, if not removed from the system, can result in buildup of iron sulfide. Other advantages include lower vapor pressure, lower heat of reaction, higher resistance to degradation, and high selectivity for H_2S.

The overwhelming advantage of MDEA is its selectivity for H_2S in the presence of CO_2. At high CO_2/H_2S ratios, a major portion of CO_2 can be slipped through the absorber and into the sales gas while removing most of H_2S. The enhanced selectivity of MDEA for H_2S results from its inability to form carbamate with CO_2. Selectivity absorption of H_2S can be enhanced by controlling the residence time per tray to 1.5–3.0 s and increasing the temperature in the absorber. Both of these conditions favor H_2S absorption with CO_2 rejection.

Mixed Amines

Mixtures of amines are generally mixtures of MDEA and DEA or MEA and are used to enhance CO_2 removal by MDEA. Such mixtures are referred to as MDEA-based amines. The secondary amine generally comprises less than 20% (mole) of the total amine. At lower MEA and DEA concentrations, the overall amine strength can be as high as 55% (wt), without invoking corrosion problems.

Amine mixtures are particularly useful for low-pressure applications because MDEA becomes less capable of CO_2 pickup sufficient enough to meet pipeline specifications. At higher pressures, amine mixtures appear to have little or no advantage over MDEA. Mixed amines are also useful for cases where the CO_2 content of the feed gas is increasing over time due to field aging.

Design of Amine Units

The design of amine units are similar for different solvents; however, a reclaimer should be added in the case of more reactive amines such as MEA to recover these amines. The main equipment in the amine processes is the absorber, and emphasis here will be on the design of the absorber. The main operating conditions such as circulation rate, solvent concentration, acid gas loading, steam rate for regeneration are given in Table 3.

Absorber Design

For 35 wt% DEA and using 0.5 mol acid gas/mol DEA, the amine circulation rate q_{Am} is

$$q_{\mathrm{Am}}(\mathrm{gpm}) = 126\, Q_g X_{\mathrm{AG}} \tag{10}$$

where Q_g is the gas flow rate (MMSCFD) and X_{AG} the mole percent of acid gas in the feed. Assuming an amine drop size of 100 μm, the diameter d (in.) of the absorber can be calculated as (Arnod and Stewart)

$$d^2 = 504 \frac{TZQ_g}{P} \left[\left(\frac{\rho_g}{\rho_l - \rho_g} \right) C_d \right]^{1/2} \tag{11}$$

where T is the temperature (°R), Z is the compressibility factor, P is the pressure (psia), and C_d is the drag coefficient. The density of gas and liquid are ρ_g and ρ_l (lb/ft^3), respectively. Usually 20 valve-type trays are used with a spacing of 24 in.

Example 2

A natural gas is flowing at 100 MMSCFD at 1000 psia and 100°F. It contains 2 mol% CO_2 and 1.9 mol% H_2S. It is being treated by the amine process using DEA. The outlet gas contains no acid components. Calculate the flow rate of DEA required and the tower diameter and height.

Solution

$$\text{Total acid gas} = 2 + 1.9 = 3.9 \text{ mol\%}$$

$$\text{Amine flow rate } (q_{\text{Am}}) = 126X_{\text{AG}}$$

$$q_{\text{Am}} = 126(3.9) = 491.4 \text{ gpm}$$

The tower diameter is

$$d^2 = 504\left(\frac{560(0.85)(100)}{1000}\right)\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)C_d\right]^{1/2}$$

Assuming the drag coefficient (C_d) is 0.7,

$$\rho_g = \frac{PM}{2RT} = \frac{(1000)(16)}{(0.85)(10.7)(560)} = 3.14\,\frac{\text{lb}}{\text{ft}^3}$$

where P is the pressure and M is the molecular weight. Assuming an average molecular weight of natural gas the same as methane (16), then

$$\rho_L = \rho_{\text{water}} \text{ at } 100^\circ\text{F}$$
$$= 65.1\,\frac{\text{lb}}{\text{ft}^3}$$

$$\frac{\rho_g}{\rho_l - \rho_g} = \frac{3.14}{65.1 - 3.14} = 0.05068 = 0.0507$$

$$d^2 = 504(47.6)[0.0507(0.7)]^{1/2}$$

then

$$d = 67.2 \text{ in.} = 5.6 \text{ ft}$$

Use 24 trays with a spacing of 24 ft with a height of (24)(2) = 48 ft.

11.5.2 Hot Potassium Carbonate Process

In this process, hot potassium carbonate (K_2CO_3) is used to remove both CO_2 and H_2S. It can also remove (reversibly) COS and CS. It works best when the CO_2 partial pressure is in the range 30–90 psi. The following reactions occur in this case:

$$K_2CO_3 + CO_2 + H_2O \Leftrightarrow 2KHCO_3 \tag{12}$$

$$K_2CO_3 + H_2S \Leftrightarrow KHS + KHCO_3 \tag{13}$$

It can be seen from reaction (12) that a high partial pressure of CO_2 is required to keep $KHCO_3$ in solution, and in Eq. (13), H_2S will not react if the CO_2 pressure is not high. For this reason, this process cannot achieve a low concentration of acid gases in the exit stream and a polishing process is needed (molecular sieve). An elevated temperature is also necessary to ensure that potassium carbonate and reaction products ($KHCO_3$ and KHS) remain in solution. Thus, this process cannot be used for gases containing H_2S only.

The hot carbonate process which is given in Figure 7 is referred to as the "hot" process because both the absorber and the regenerator operate at elevated temperatures, usually in the range (230–240°F). In Figure 7, the sour gas enters at the bottom of the absorber and flows countercurrently to the carbonate liquid stream. The sweet gas exits at the top of the absorber. The absorber is operated at 230°F and 90 psia. The rich

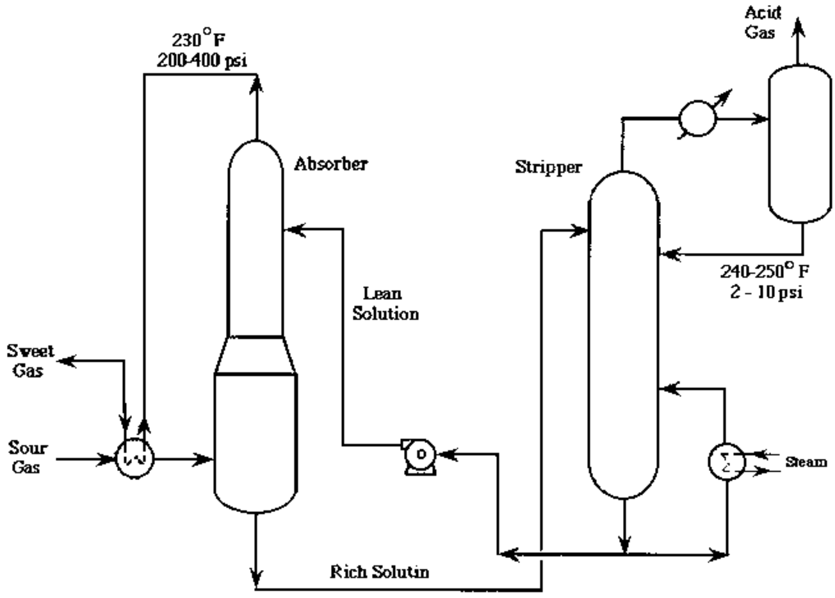


Figure 7 Hot carbonate process.

carbonate solution exits from the bottom of the absorbed and is flashed in the stripper, which operates at 245°F and atmospheric pressure, where acid gases are driven off. The lean carbonate solution is pumped back to the absorber.

The strength of the potassium carbonate solution is limited by the solubility of potassium bicarbonate ($KHCO_3$) in the rich stream. The high temperature of the system increases $KHCO_3$ solubility, but the reaction with CO_2 produces 2 mol of $KHCO_3$ per mole of K_2CO_3 reacted. For this reason, $KHCO_3$ in the rich stream limits the lean solution of K_2CO_3 concentration to 20–35% (wt).

11.5.3 Physical Solvent Processes

Organic liquid (solvents) are used in these processes to absorb H_2S (usually) preferentially over CO_2 at high pressure and low temperatures. Regeneration is carried out by releasing the pressure to the atmosphere and sometimes in vacuum with no heat. If we apply Henry's law,

$$P_i = HX_i$$

or

$$X_i = \frac{Y_i}{H} P \tag{14}$$

This implies that acid gas absorbed in liquid phase (X_i) is proportional to its gas mole fraction (Y_i) and inversely to Henry's constant (which is constant for a given temperature). Much more importantly, the solubility is proportional to the total gas pressure (P). This means that at high pressure, acid gases will dissolve in solvents, and as the pressure is released, the solvent can be regenerated.

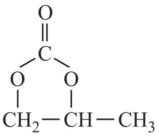
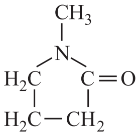
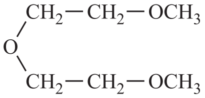
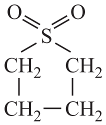
The properties of four of the important solvents used in natural gas processing are given in Table 4.

Fluor Process

This process uses propylene carbonate to remove CO_2, H_2S, C_2^+, COS, CS_2, and H_2O from natural gas. Thus, in one step, the natural gas can be sweetened and dehydrated.

Figure 8 shows a typical process flowsheet with regeneration consisting of three flash drums. The first flash drum gas containing mostly hydrocarbons is compressed and recycled. The second flash drum derives expansion turbine. The third flash contains mainly acid gases. The process is used to remove the bulk CO_2 down to 3%.

Table 4 Properties of Physical Solvents

Process	Fluor	Purisol	Selexol	Sulfinol[a]
Solvent	Propylene carbonate	*N*-Methyl pyrrilidone	Diethylene dimethyl ether	Sulfolane
Structure				
Molecular weight	102.09	99.13	134.17	120.17
Freezing (°F)	−56	−12	−83	77
Boiling (°F)	467	396	324	546
Gas solubility (cm^3 gas at 1 atm, 75°F/cm^3 solvent)				
H_2S	13.3	43.3	25.5	
CO_2	3.3	3.8	3.6	
COS	6.0	10.6	9.8	
C_3	2.1	3.5	4.6	

[a] Sulfinol is a mixture of sulfolane + DIPA + H_2O.

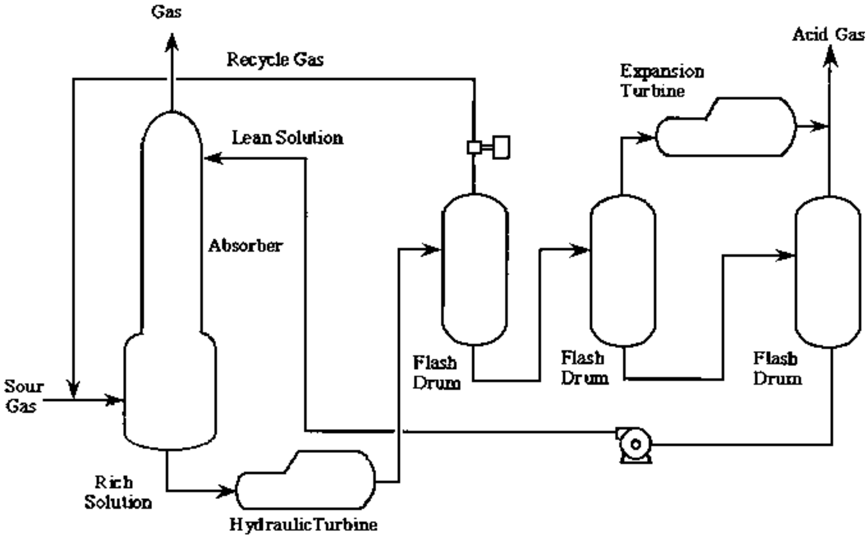


Figure 8 Fluor process.

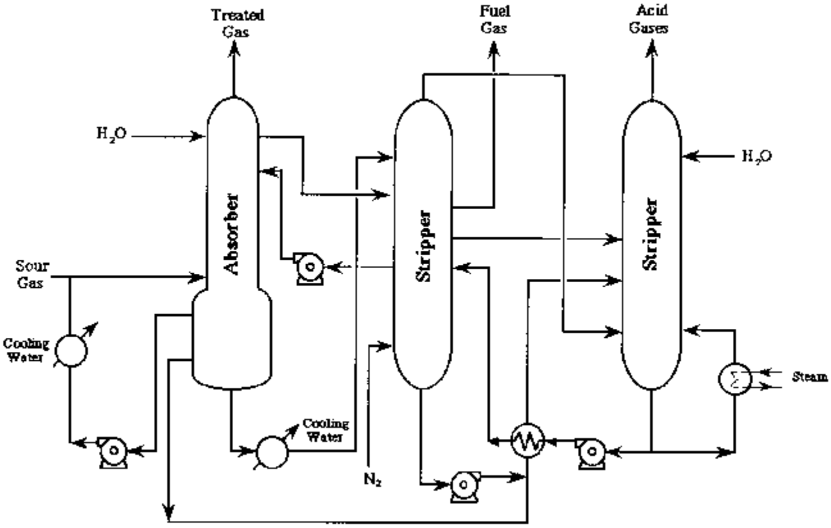


Figure 9 Purisol process.

Purisol Process

This process uses *N*-methyl-2-pyrrolidone also known as NMP as a solvent licensed by Lurgi. The solvent removes H_2S, CO_2, H_2O, RSH, and hydrocarbons and elastomers. The feature of this solvent is that it is highly selective for H_2S, as shown in Table 3. It has a boiling point of 396°F, which is rather low to be used in amine mixed solvents.

Figure 9 shows a typical flowsheet for the purisol process. Regeneration is accomplished by two strippers, where dissolved hydrocarbons are stripped off as fuel gas nitrogen in the first drum and acid gases are stripped in the second stripper; regenerated NMP is recycled back to the absorber.

Selexol Process

This process uses a mixture of dimethyl ether of propylene glycols as a solvent. It is nontoxic and its boiling point is not high enough for amine formulation. Figure 10 shows the flowsheet of the Selexol process. A cool stream of natural gas is injected in the bottom of the absorption tower operated at 1000 psia. The rich solvent is flashed in a high flash drum at 200 psia, where methane is flashed and recycled back to the absorber and joins the sweet gas stream. The solvent is then flashed at atmospheric pressure and acid gases are flashed off. The solvent is then stripped by

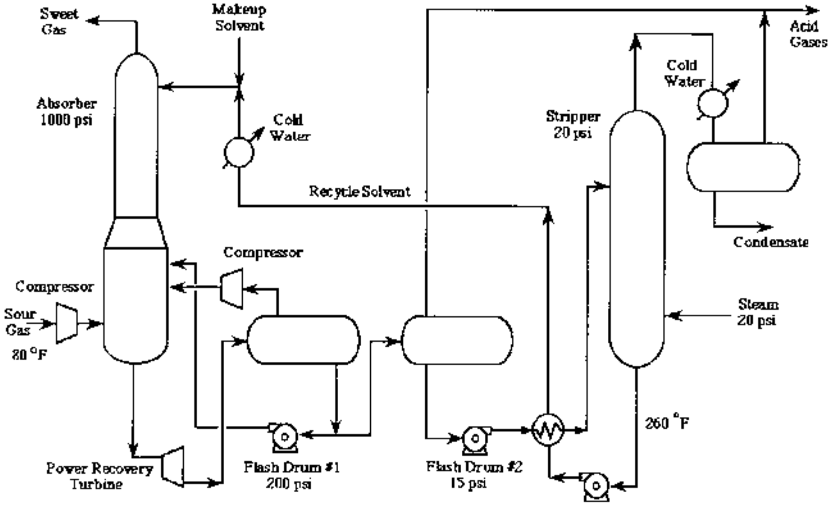


Figure 10 Selexol process.

steam to completely regenerate the solvent, which is recycled back to the absorber; any hydrocarbons will be condensed and any remaining acid gases will be flashed from the condenser drum.

This process is used when there is a high acid gas partial pressure and no heavy hydrocarbons. DIPA can be added to this solvent to remove CO_2 down to pipeline specifications.

Sulfinol Process

This process uses a solvent which is 40% sulfolane (tetrahydrothiophene 1-1 dioxide), 40% DIPA (di-isopropanolamine), and 20% water. It is an excellent example of enhancing amine selectivity by adding a physical solvent such as sulfolane. Sulfolane is an excellent solvent of sulfur compounds such as H_2S, COS, and CS_2. Aromatics, heavy hydrocarbons, and CO_2 are soluble to a lesser extent. Sulfinol is usually used for H_2S/CO_2 ratios greater than 1:1 or where CO_2 removal is not required to the same extent as H_2S. A high loading of 1.5 mol acid gas per mole of sulfinol can be achieved. The sulfinol process uses a conventional solvent absorption and regeneration cycle, as shown in Figure 11. The sour gas components are removed from the feed gas by countercurrent contact with a lean solvent stream under pressure. The absorbed impurities are then removed from the rich solvent by stripping with steam in a heated regenerator column. The hot lean solvent is then cooled for reuse in the

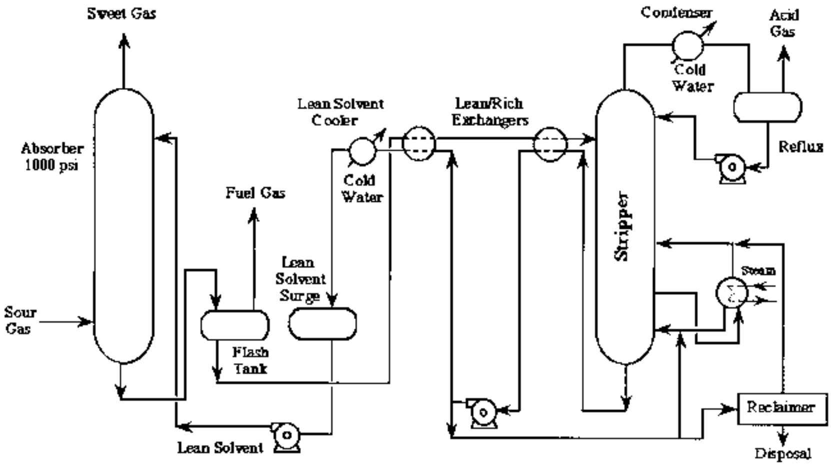


Figure 11 Sulfinol process.

absorber. Part of the cooling may be by heat exchange with the rich solvent for partial recovery of heat energy. In some applications where there is a large amount of hydrogen sulfide in the feed, the overall Sulfinol/Claus plant has a balance that permits the omission of the heat exchanger. The gas flashed from the rich solvent after partial depressuring is shown in Figure 11 as fuel gas. In some cases, it is desirable to treat this gas stream with Sulfinol solvent to control the acid gas content of the plant fuel supply.

The Sulfinol solvent reclaimer is used in a small ancillary facility for recovering solvent components from higher boiling products of alkanolamine degradation or from other high-boiling or solid impurities. A Sulfinol reclaimer is similar to a conventional MEA reclaimer, but it is much smaller than that in an MEA plant of comparable gas treating capacity. Usually, the Sulfinol reclaimer need not be started up until several months after the treating plant is started up.

Comparison of Physical Solvents

Purisol has the highest capacity for absorption of acid gases and it is the most selective; however, it is the most volatile. Selexol is more selective than Fluor solvent, but it dissolves propane. All solvents exhibit significant affinity for heavy paraffins, aromatics, and water. Water absorption make them good dessicants.

The loading capacity of physical solvent is much higher than amines. If we consider Figure 12, we see that at a partial pressure of H_2S of

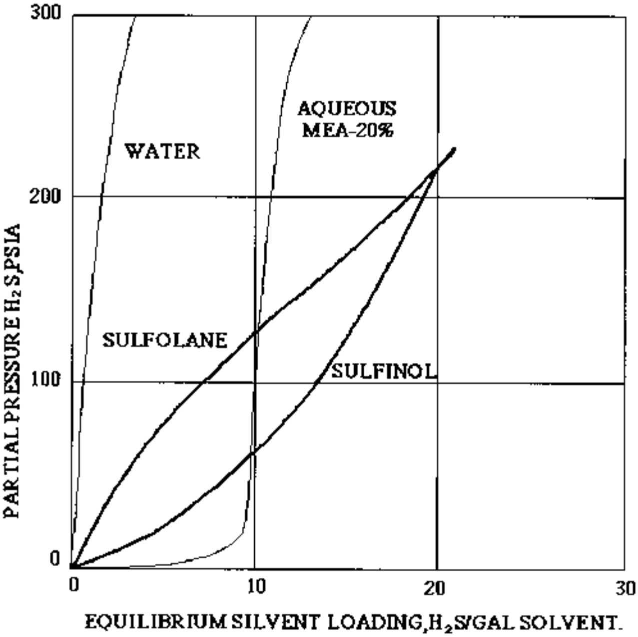


Figure 12 Equilibrium solvent loadings.

200 psia, the loading (mol H_2S/gal solvent) of MEA (20% solution) is about 11.5, and at the same time, sulfolane (physical solvent) is about 18 and sulfinol (which is a mixed solvent) is about 19.

11.6 DIRECT CONVERSION PROCESSES

There are many processes used to convert H_2S to sulfur; however, our discussion here is limited to those processes applied to natural gas. Generally, H_2S is absorbed in an alkine solution containing an oxidizing agent which converts it to sulfur. The solution is regenerated by air in a flotation cell (oxidizer). The following processes are used for this purpose.

11.6.1 Stretford Process

The absorbing solution is dilute Na_2CO_3, $NaVO_3$, and anthraquinone disulfonic acid (ADA). The reaction occurs in four steps:

$$H_2S + Na_2CO_3 = NaHS + NaHCO_3 \tag{15}$$

$$4\mathrm{NaVO_3} + 2\mathrm{NaHS} + \mathrm{H_2O} = \mathrm{Na_2V_4O_9} + 4\mathrm{NaOH} + 2\mathrm{S} \tag{16}$$

$$\begin{aligned} &\mathrm{Na_2V_4O_9} + 2\mathrm{NaOH} + \mathrm{H_2O} + 2\mathrm{ADA(quinone)} \\ &\quad = 4\mathrm{NaVO_3} + 2\mathrm{ADA(hydroquinone)} \end{aligned} \tag{17}$$

$$\mathrm{ADA(hydroquinone)} + \tfrac{1}{2}\mathrm{O_2} = \mathrm{ADA(quinone)} \tag{18}$$

The process uses anthraquinone disulfonic acid (ADA) as the organic oxygen carrier. One of the products is finely divided sulfur and the process is capable of treating natural gas of very low H_2S concentrations. In the oxidizer or regenerator, the reduced anthraquinone disulfonic acid is reoxidized by blowing air, as shown by reaction (18). The precipitated sulfur is overflown as froth.

The Stretford process is shown in Figure 13. Sour gas enters the bottom of absorber and sweet gas exits at the top. The Stretford solution enters at the top of the absorber and some time should be allowed for reaction to take place in the bottom part of the absorber, where H_2S is selectively absorbed. The reaction products are fed to the oxidizer, where air is blown to oxidize ADA (hydroquinone) back to ADA (quinone). The sulfur froth is skimmed and sent to either a filtration or centrifugation unit. If heat is used, molten sulfur is produced; otherwise a filter sulfur cake is obtained. The filtrate of these units along with the liquid from the oxidized are sent back to the absorber.

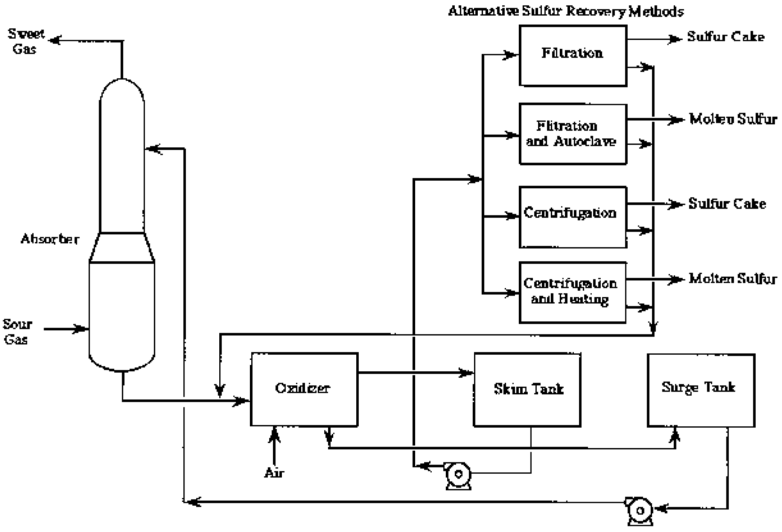


Figure 13 Stretford process.

11.6.2 LOCAT Process

This process uses as extremely dilute solution of iron chelates. A small portion of the chelating agent is depleted in some side reactions and is lost with precipitated sulfur. In this process (Fig. 14), sour gas is contacted with the chelating reagent in the absorber and H_2S reacts with the dissolved iron to form elemental sulfur. The reactions involved are the following:

$$H_2S + 2Fe^{3+} \rightarrow 2H^+ + S + 2Fe^{2+} \tag{19}$$

The reduced iron ion is regenerated in the generated by blowing air as

$$\tfrac{1}{2}O_2 + H_2O + 2Fe^{2+} \rightarrow 2(OH)^- + 2Fe^{3+} \tag{20}$$

The sulfur is removed from the regenerator to centrifugation and melting. Application of heat is not required because of the exothermic reaction.

11.6.3 Sulferox Process

Chelating iron compounds are also the heart of the sulferox process. Sulferox is a redox technology, as is the LoCat; however, in this case, a

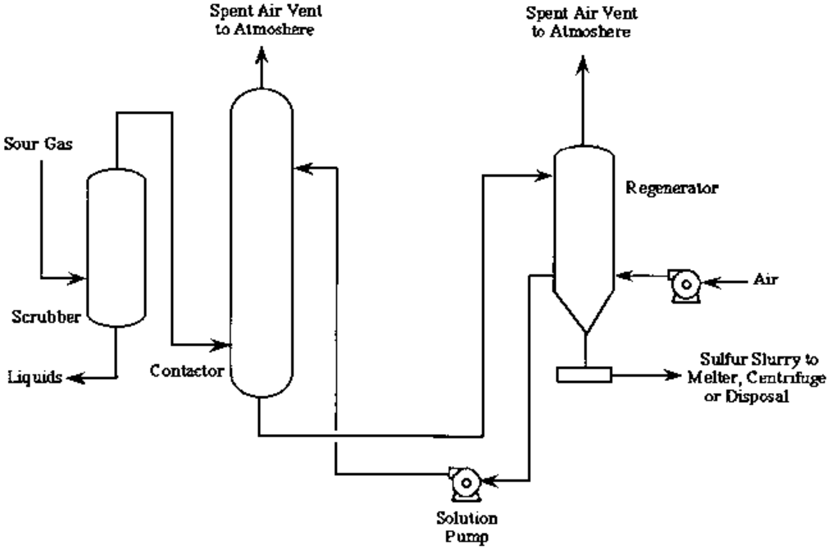


Figure 14 LOCAT process.

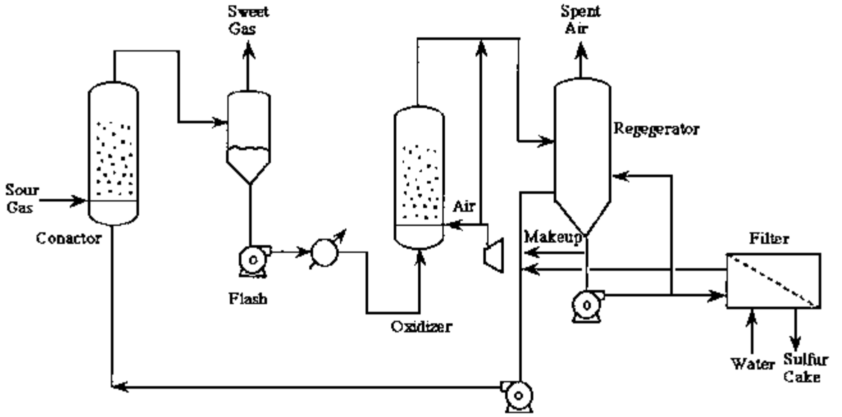


Figure 15 Sulferox process.

concentrated iron solution is used to oxidize H_2S to elemental sulfur. Patented organic liquids or chelating agents are used to increase the solubility of iron in the operating solution. As a result of high iron concentrations in the solution the rate of liquid circulation can be kept low and, consequently, the equipment is small.

As in the LOCAT process, there are two basic reactions; the first takes place in the absorber, as in reaction (19), and the second takes place in the regenerator, as in reaction (20). The key of the Sulferox technology is the ligand used in the process. The application of this ligand allows the process to use high total iron concentrations ($>1\%$ by weight).

In Figure 15, the sour gas enters the contactor, where H_2S is oxidized to give elemental sulfur. The treated gas and the Sulferox solution flow to the separator, where sweet gas exits at the top and the solution is sent to the regenerator where Fe^{2+} is oxidized by air to Fe^{3+} and the solution is regenerated and sent back to the contactor. Sulfur settles in the regenerator and is taken from the bottom to filtration, where sulfur cake is produced. At the top of the regenerator, spent air is released. A makeup Sulferox solution is added to replace the degradation of the ligands. Proper control of this degradation rate and purging of the degradation products will ensure smooth operation of the process.

11.6.4 Membrane Processes

Polymeric membranes separate gases by selective permeation of gas species in these membranes. The gas dissolves at the contact surface of the membrane and permeate across the membrane under the partial pressure gradient across the membrane wall. The rate of permeation of gas A(q_A) can be expressed as

$$q_A = \frac{PM}{t} A_m \Delta P_A \tag{21}$$

where PM is the gas permeability in the membrane, A_m and t are the surface area and thickness of the membrane, respectively, and ΔP_A is the partial pressure of gas A across the membrane.

The basic idea of the process is to flow sour gas on one side of the membrane where only acid gases diffuse across the membrane to the permeate side and the rest of the gas exits as sweet gas, as shown in Figure 16. Two module configurations are usually used: the spiral module and the hollow-fiber module. Spiral-wound membranes consist of sandwich of four sheets wrapped around a central core of a perforated collecting tube. The whole spiral-wound element is housed inside a metal shell. The feed gas enters at the left end of the shell, enters the feed channel, and flows through this channel in the axial direction of the spiral to the right end of the assembly, as shown in Figure 17. The exit sweet gas leaves the shell at this point. The acid gases permeate perpendicularly through the membrane. This permeate then flows through the permeate channel to the perforated collecting tube, where it leaves the apparatus at one end. The direction of flows in spiral-wound module is shown in Figure 18.

The hollow-fiber module consists of a bundle of very small-diameter hollow fibers. The module resembles a shell and tube heat exchanger. Thousands of fine tubes are bound together at each end into a tube sheet that is surrounded by metal shell (see Fig. 19). The membrane area per unit volume is up to 3000 ft^2/ft^3. Acid gases diffuse through the very thin membrane of the tubes and exit at the bottom of the module. Sweet gas exits at the top.

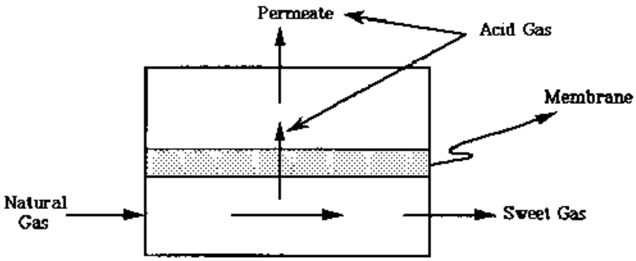


Figure 16 Basic operation of cross-flow pattern in the membrane.

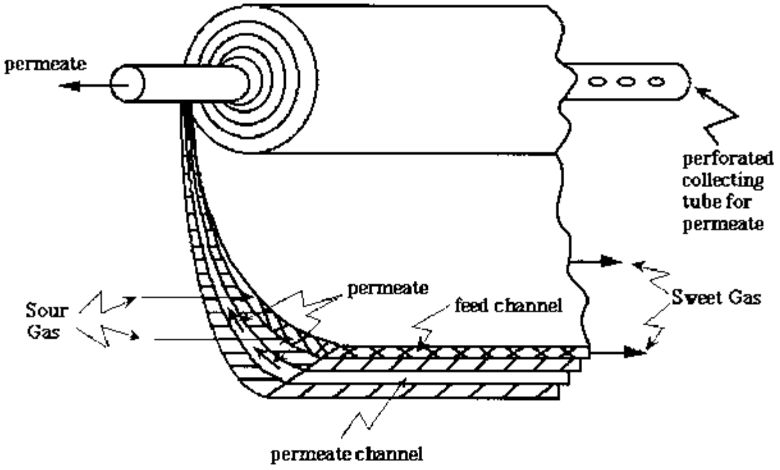


Figure 17 Spiral-wound elements and assembly.

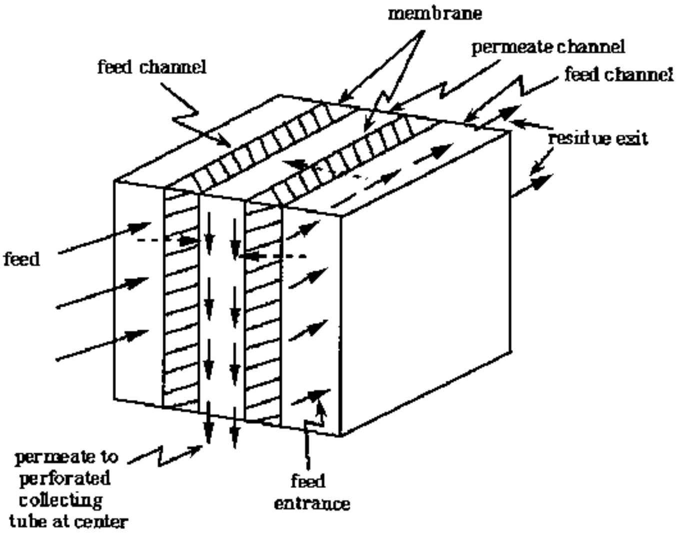


Figure 18 Gas flow paths for spiral-wound module.

In both modules, high pressure should be maintained to ensure high permeation rates. Table 5 lists the permeability (PM) of some gases in different membranes. The permeation rates of different gases in a commercial membrane is given in Table 6. It is noted from Table 6 that

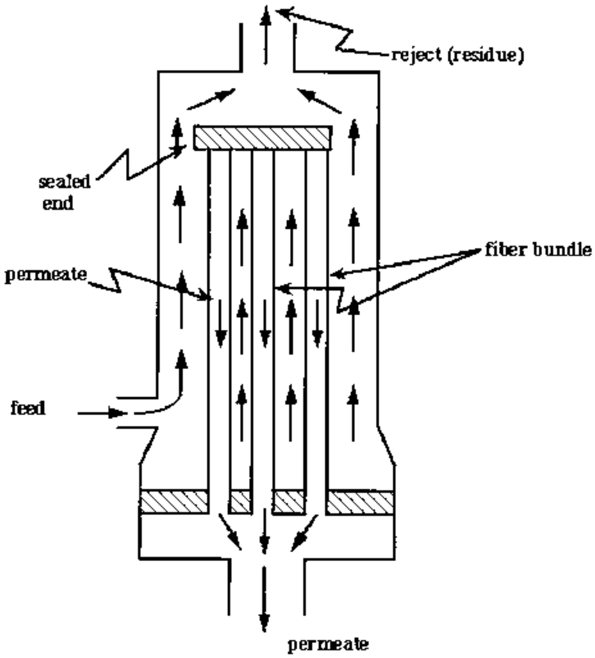


Figure 19 Hollow-fiber separator assembly.

Table 5 Permeabilities of Various Gases in Membranes

	Permeability (PM) $\frac{\text{cm}^3(\text{STP}) \cdot \text{cm}}{\text{s} \cdot \text{cm}^2/\text{cmHg}} \times 10^{10}$						
Material	Temp. (°C)	He	H_2	CH_4	CO_2	O_2	N_2
Silicone rubber	25	300	550	800	2700	500	250
Natural rubber	25	31	49	30	131	24	8.1
Polycarbonate (Lexane)	25–30	15	12	—	5.6, 10	1.4	—
Nylon 66	25	1.0	—	—	0.17	0.034	0.008
Polyester (Permasep)	—	—	1.65	0.035	0.31	—	0.031
Silicone–polycarbonate copolymer (57% silicone)	25	—	210	—	970	160	70
Teflon FEP	30	62	—	1.4	—	—	2.5
Ethyl cellulose	30	35.7	49.2	7.47	47.5	11.2	3.29
Polystyrene	30	40.8	56.0	2.72	23.3	7.47	2.55

Source: Reference [4]. STP, standard temperature and pressure.

Table 6 Gas Permeation Rates

Gas	Spiral wound membrane
H_2	100.0
He	15.0
H_2O	12.0
H_2S	10.0
CO_2	6.0
O_2	1.0
Ar	—
CO	0.3
CH_4	0.2
N_2	0.18
C_2H_6	0.1

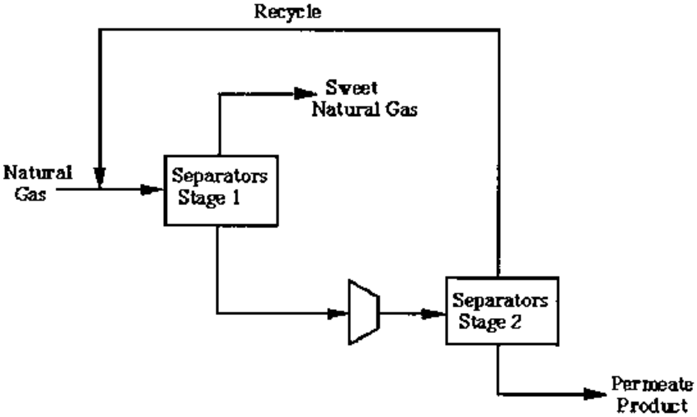


Figure 20 Two-stage membrane process.

H_2S and CO_2 will permeate much faster than other hydrocarbons. However, the permeate gas will also contain hydrogen and water vapor if they were present in the natural gas feed. It is possible to enhance the mass transfer rates by either blowing an inert gas in the permeate side or using amines on the permeate side to chemically react with the acid gas permeated and removed from the surface of the membrane. Several membrane modules may be connected in series and parallel arrangements to meet specific requirements. In Figure 20, a two-stage system is shown. If a gas stream contains a high acid content (50%), it is possible to reduce it in a single module to about 20% and to use amine absorption unit to reduce it to natural gas acid concentration, as shown in Figure 21.

11.6.5 Environmental Aspects

Hydrogen sulfide is one of the major pollutants in petroleum field processing, and acid gas treatment is one of the most important steps to eliminate H_2S. However, in the course of this treatment, some side products can cause pollution problems and must be eliminated.

In amine treatment, the following pollutants are produced and the corresponding countermeasures are given in Table 7.

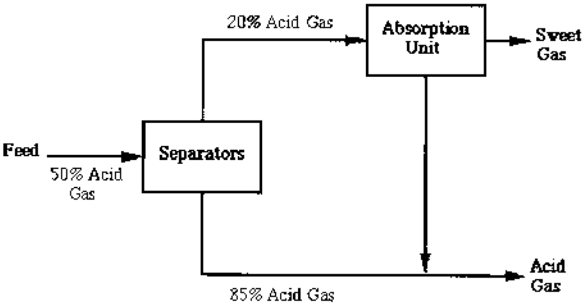


Figure 21 Bulk acid gas removal by the membrane process.

Table 7 Pollution Prevention Alternatives

Waste	Pollution prevention alternatives
Amines	• Use an amine reclaimer in the system to allow reuse of amine and minimize the volume of waste generated. • Use an amine filter to extend the life of solution and maintain efficiency.
Filters	• Change filters only when necessary; use differential pressure as indicator of needed change. • Use reusable filters. • When handling filters, care should be taken to prevent oil spillage. • Isolate all drained fluids in tight container for recycling.
Iron sponge and Iron sulfide scale	• Consider alternative methods of removing H_2S from gas stream. • Treat production streams with biocide or scale inhibitor to reduce iron sulfide formation.

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REVIEW QUESTIONS AND EXERCISE PROBLEMS

1. Name the main categories of the sweetening processes.
2. What constituents from natural gas are removed by the sweetening process?
3. What is the difference between "acid gases" and "sour gases"?
4. The sulfinol process is an example of:

   (a) Physical
   (b) Dry bed
   (c) Chemical
   (d) Chemical/physical (hybrid).

5. Some sweetening processes have a "dual function"; they bring in gas dehydration as well. Which of the following sweetening processes fall into this category?

   (a) MEA
   (b) DGA
   (c) DEA
   (d) Sulfinol
   (e) Molecular sieves.

6. A sour crude oil contains 1.2 SCF of dissolved H_2S per 1000 gallons of crude oil. If the buyer's specs for crude oil is 50 ppm maximum, calculate:

   (a) The H_2S content of the sour crude oil in ppm
   (b) The amount of H_2S gas (SCF) to be removed per 1000 bbl of oil in order to meet the specs.

7. Why is the selective removal of H_2S with respect to CO_2 in some special cases desirable?

8. Most amine solvents are regenerated by:

   (a) Lowering both T and P
   (b) Increasing both T and P
   (c) Lowering T and increasing P
   (d) Lowering P and increasing T.

12

Gas Dehydration

12.1 INTRODUCTION

Natural gas dehydration is the process of removing water vapor from the gas stream to lower the dew point of that gas. Water is the most common contaminant of hydrocarbons. It is always present in the gas–oil mixtures produced from wells. The dew point is defined as the temperature at which water vapor condenses from the gas stream. The sale contracts of natural gas specify either its dew point or the maximum amount of water vapor present. There are three basic reasons for the dehydration of natural gas streams:

1. **To prevent hydrate formation.** Hydrates are solids formed by the physical combination of water and other small molecules of hydrocarbons. They are icy hydrocarbon compounds of about 10% hydrocarbons and 90% water. Hydrates grow as crystals and can build up in orifice plates, valves, and other areas not subjected to full flow. Thus, hydrates can plug lines and retard the flow of gaseous hydrocarbon streams. The primary conditions promoting hydration formation are the following:
    1. Gas must be at or below its water (dew) point with "free" water present.
    2. Low temperature.
    3. High pressure.
2. **To avoid corrosion problems.** Corrosion often occurs when liquid water is present along with acidic gases, which tend to dissolve and disassociate in the water phase, forming acidic solutions. The acidic solutions can be extremely corrosive, especially for carbon steel, which is typically used in the construction of most hydrocarbon processing facilities.

3. **Downstream processing requirements.** In most commercial hydrocarbon processes, the presence of water may cause side reactions, foaming, or catalyst deactivation. Consequently, purchasers typically require that gas and liquid petroleum gas (LPG) feedstocks meet certain specifications for maximum water content. This ensures that water-based problems will not hamper downstream operations.

12.2 PREDICTION OF HYDRATE FORMATION

In this section, methods for determining the operating conditions leading to hydrate formation are presented. In particular, methods are presented to determine the following:

1. Hydrate formation temperature for a given pressure
2. Hydrate formation pressure for a given temperature
3. Amount of water vapor that saturates the gas at a given pressure and temperature (i.e., at the dew point)

At any specified pressure, the temperature at which the gas is saturated with water vapor is being defined as the "dew point". Cooling of the gas in a flow line due to heat loss can cause the gas temperature to drop below the hydrate formation temperature.

Two methods are discussed next for predicting the conditions leading to hydrate formation: approximate methods and analytical methods.

12.2.1 Approximate Methods

1st: To determine hydrate formation pressure or temperature. This method utilizes the chart shown in Figure 1; it involves the following steps:

1. Calculate the average molecular weight of the gas mixture M_G
2. Calculate the specific gravity of the gas, γ_G, where $\gamma_G = M_G/M_{\text{air}}$.
3. Given the gas temperature (T) and its specific gravity (γ_G), find the corresponding gas pressure for hydrate formation using Figure 1.
4. Similarly, at a given gas pressure, the hydrate formation temperature can be determined from Figure 1.

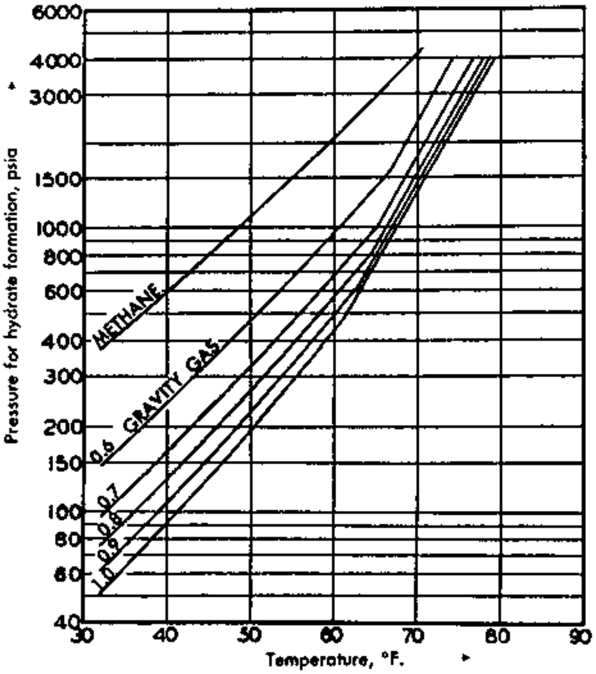

Figure 1 Pressure–temperature curves for predicting hydrate formation. (From Ref. 1.)

Example 1

If the specific gravity of a natural gas stream is calculated and found to be 0.69, find the hydrate formation pressure (in psia) at a temperature of 50°F.

Solution

From Figure 1, a pressure of about 330 psia is determined for the corresponding conditions of temperature and specific gravity stated.

2nd: To Determine the amount of water vapor at dew point condition. This method utilizes McKetta–Wehe chart [2], which relates the gas temperature T and its pressure P to the water vapor content of the natural gas at saturation conditions, as presented in Figure 2.

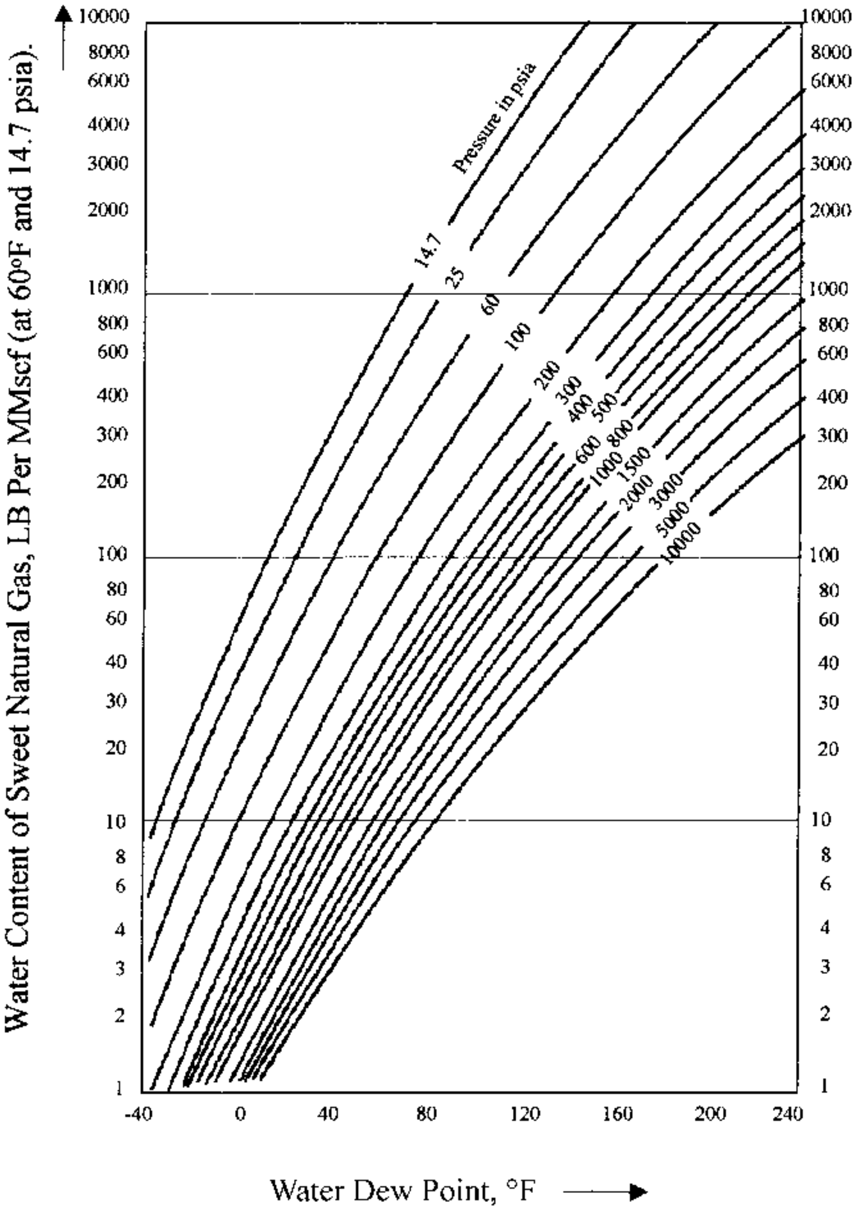

Figure 2 Water content of sweet, lean natural gas. (From Ref. 2.)

Example 2

Natural gas, saturated with water vapor at conditions of 1000 psia and 90°F is exposed to cooling in a flow line due to heat losses, where the temperature reaches 35°F and the pressure remains the same.

(a) Calculate how much liquid water will drop out of the gas.

(b) Assuming that the gas flowing through the pipeline is to reach a delivery point at 300 psia pressure, find the corresponding dew point of the gas.

Solution

(a) From Figure 2, we obtain the following:

Water content of the gas stream at the initial conditions (1000 psia and 90°F) is 46 lb/MMCF (MMCF: 10_6 cubic feet).

Water content of the gas stream at conditions of 1000 psia and 35°F is 7.6 lb/MMCF.

Water to be separated is $46 - 7.6 = 38.4$ lb/MMCF (MMCF: 10_6 cubic feet).

(b) The natural gas, once it reaches the delivery point at 300 psia, carries with it a water content of 7.6 lb/MMCF. Applying these two parameters to Figure 2, one can read the dew point temperature of 12°F.

12.2.2 Analytical Methods

The calculations presented in this section are concerned with finding the hydrate formation temperature T at a given pressure P, or the pressure P at which hydrate formation takes place for a given operating temperature T. A knowledge of the temperature and pressure of a gas stream at the wellhead is important for determining whether hydrate formation can be expected when the gas is expanded into the flow lines. In general, the temperature at the wellhead can change as the reservoir conditions or production rate change over the production life of the well. Wells, therefore, that initially flowed at conditions where no hydrate formation occurred in downstream equipment may require hydrate inhibition, or vice versa. The computational approach is analogous to the one used in the dew point calculation for a multicomponent mixture of hydrocarbon gases. The basic equation is given by

$$\sum_{i=1}^{n} \frac{y_i}{K_i} = 1 \tag{1}$$

where y_i is the mole fraction of component i in the gas phase, on a water–free basis, K_i is the vapor–solid equilibrium constant for component i, defined by $K_i = y_i/x_i$, and x_i is the mole fraction of component i in the solid phase, on a water-free basis.

Given the gas pressure P and the mole fraction of its components y_i, the hydrate formation temperature T is obtained through a trial-and-error procedure according to the following steps:

1. Assume a value for T.
2. Final the values of K for the components of the gas at P and T using the K charts.
3. Calculate the values of x_i, where $x_i = y_i/K_i$.
4. Find the sum of y_i/K_i for all components.
5. Convergence to the desired value of T is obtained by trial-and-error until

$$\sum_{i=1}^{n} \frac{y_i}{K_i} = 1$$

A sample chart for the vapor–solid equilibrium constant for isobutane is given in Figure 3 [3].

12.3 METHODS USED TO INHIBIT HYDRATE FORMATION

Hydrate formation in natural gas is promoted by high-pressure, low-temperature conditions and the presence of liquid water. Therefore, hydrates can be prevented by the following:

1. Raising the system temperature and/or lowering the system presure (*temperature/pressure control*)
2. Injecting a chemical such as methanol or glycol to depress the freezing point of liquid water (*Chemical injection*)
3. Removing water vapor from the gas liquid–water drop out that is depressing the dew point (*dehydration*).

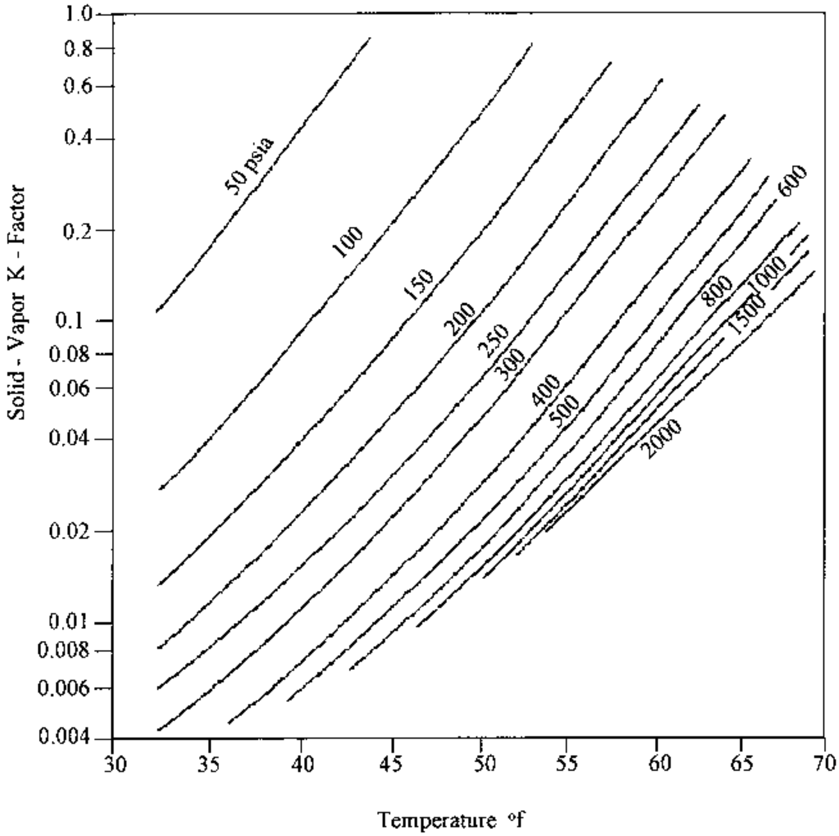

Figure 3 Vapor–solid equilibrium constants for isobutane (Ki). (From Ref. 3.)

12.3.1 Temperature/Pressure Control

Methods recommended for temperature control of natural gas streams include the following:

1. **Downhole regulators or chokes.** In this method, a pressure regulator (choke) is installed downhole (in the well). This causes the largest portion of the desired pressure drop from the bottom-hole flowing pressure to the surface flow line pressure to occur where the gas temperature is still high. The bottom-hole temperature will be sufficiently high to prevent hydrate formation as the pressure is reduced. At the surface, little or

no pressure reduction may be required, thus hydrate formation is also avoided at the surface.

2. **Indirect heaters.** Both wellhead and flow line indirect heaters are commonly used to heat natural gas to maintain the flowing temperature above the hydrate formation temperature. The primary purpose of the wellhead heater is to heat the flowing gas stream at or near the wellhead, where chocking or pressure reduction frequently occurs. Flow line heaters, on the other hand, provide additional heating if required. They are particularly used for cases where the conditions necessitate a substantial reduction in pressure between the wellhead stream and the next field processing facility. A typical example is an offshore production field with its treating facilities on land. Heat is utilized to compensate for the loss of temperature as the gas expands. Another significant factor that contributes to the reduction of the gas temperature in a gas line is the ground temperature. A temperature drop of about 80°F is experienced by a gas flow line travelling a distance of about 5000 ft.

12.3.2 Chemical Injection

Methanol and glycols are the most commonly used chemicals, although others (such as ammonia) have been applied to lower the freezing point of water, thus reducing (or preventing) hydrate formation.

The application of hydrate inhibitors should be considered for such cases:

- A system of gas pipelines, where the problem of hydrate formation is of short duration.
- A system of gas pipelines which operate at a few degrees below the hydrate formation temperature.
- Gas gathering systems found in pressure-declining fields.
- Gas lines characterized by hydrate formation in localized points.

Inhibitors function in the same manner as "antifreeze" when added to liquid water. Thus, the principle underlying the use of hydrate inhibitors is to lower the formation of the hydrate by causing a depression of the hydrate formation temperature.

Methanol Injection

Methanol is the most commonly used nonrecoverable hydrate inhibitor. It has the following properties:

1. It is non corrosive.
2. It is chemically inert; no reaction with the hydrocarbons.
3. It is soluble in all proportions with water.
4. It is volatile under pipeline conditions, and its vapor pressure is greater than that of water.
5. It is not expensive.

Methanol is soluble in liquid hydrocarbons (about 0.5% by weight). Therefore, if the gas stream has high condensate contents, a significant additional volume of methanol will be required. This makes this method of hydrate inhibition unattractive economically because methanol is nonrecoverable. In such a situation, it will be necessary to first separate the condensate from the gas. Some methanol would also vaporize and goes into the gas. The amount of methanol that goes into the gas phase depends on the operating pressure and temperature.

In many applications, it is recommended to inject methanol some distance upstream of the point to be protected by inhibition, in order to allow time for the methanol to vaporize before reaching that point.

Glycol Injection

Glycol functions in the same way as methanol; however, glycol has a lower vapor pressure and does not evaporate into the vapor phase as readily as methanol. It is also less soluble in liquid hydrocarbons than methanol. This, together with the fact that glycol could be recovered and reused for the treatment, reduces the operating costs as compared to the methanol injection.

Three types of glycols can be used: ethylene glycol (EG), diethylene glycol (DEG), and triethelyne glycol (TEG). The following specific applications are recommended:

1. For natural gas transmission lines, where hydrate protection is of importance, EG is the best choice. It provides the highest hydrate depression, although this will be at the expense of its recovery because of its high vapor pressure.
2. Again, EG is used to protect vessels or equipment handling hydrocarbon compounds, because of its low solubility in multicomponent hydrocarbons.
3. For situations where vaporization losses are appreciable, DEG or TEG should be used, because of their lower vapor pressure.

It is of importance to mention that when hydrate inhibitors in general are injected in gas flow lines or gas gathering networks, installation of a high-pressure free-water knockout at the wellhead is of value in the operation. Removing of the free water from the gas stream ahead of the injection point will cause a significant savings in the amount of the inhibitor used.

The amount of chemical inhibitor required to treat the water in order to lower the hydrate formation temperature may be calculated from the Hammerschmidt equation:

$$\Delta T = \frac{KW}{M(100 - W)} \tag{2}$$

where ΔT is the depression in hydrate formation temperature (°F), W is weight percent of inhibitor for water treatment, K is a constant that depends on the type of inhibitor, and M is the molecular weight of the inhibitor. Values of M and K for various inhibitors are given in Table 1 [2].

Table 1 Properties of Chemical Inhibitors

Inhibitor	M	K
Methanol	32.04	2335
Ethylene Glycol	62.07	2200
Propylene Glycol	76.10	3590
Diethylene Glycol	106.10	4370

Example 3

A gas well produces 10 MMSCF/day along with 2000 lbs of water and 700 barrels per day (BPD) of condensate having a density of 300 lbs/bbl. The hydrate formation temperature at the flowing pressure is 75°F. If the average flow line temperature is 65°F, determine the amount of methanol needed to inhibit hydrate formation in the flow line given that the methanol solubility in condensate is 0.5% by weight and that the ratio of the lbs methanol in vapor/MMSCF of gas to the weight percent of methanol in water is 0.95.

Solution

To prevent hydrate formation in the flow line, we need to lower the hydrate formation temperature to 65°F or less. Therefore, the depression in hydrate formation temperature, ΔT, is

$$\Delta T = 75 - 65 = 10^\circ\mathrm{F}$$

Using Eq. (2),

$$10 = \frac{2335\,W}{32.04(100 - W)}$$

Therefore, $W = 12.07\%$

$$\begin{aligned}
\text{Required methanol in water} &= (0.1207)(2000) = 241.4\ \text{lb/day}\\
\text{Pounds of methanol in vapor} &= (0.95)(12.07) = 11.47\ \text{lbs/MMSCF}\\
&= (11.47)(10)\ \text{MMSCF/day}\\
&= 114.7\ \text{lbs/day}\\
\text{Methanol dissolved in condensate} &= (0.005)(300)\ \text{lbs/bbl} \times 700\ \text{bbl/day}\\
&= 1050\ \text{lbs/day}\\
\text{Total amount of methanol} &= 241.4 + 114.7 + 1050 = 1406\ \text{lbs/day}
\end{aligned}$$

From the results, we see that 1050 lbs of methanol (75% of the total) are dissolved in the condensate and thus do not contribute to the treatment. Such treatment is, therefore, economically unacceptable. It is evident that the condensate must be separated first before the treatment.

12.3.3 Dehydration Methods

The most common dehydration methods used for natural gas processing are as follows:

1. Absorption, using the liquid desiccants (e.g., glycols and methanol)
2. Adsorption, using solid desiccants (e.g., alumina and silica gel)
3. Cooling/condensation below the dew point, by expansion and/or refrigeration

This is in addition to the hydrate inhibition procedures described earlier. Classification of dehydration methods is given in Figure 4 [3].

12.4 ABSORPTION (GLYCOL DEHYDRATION PROCESS)

12.4.1 Basic Principles

The basic principles of relevance to the absorption process are as follows:

1. In this process, a hygroscopic liquid is used to contact the wet gas to remove water vapor from it. Triethylene glycol (TEG) is the most common solvent used.

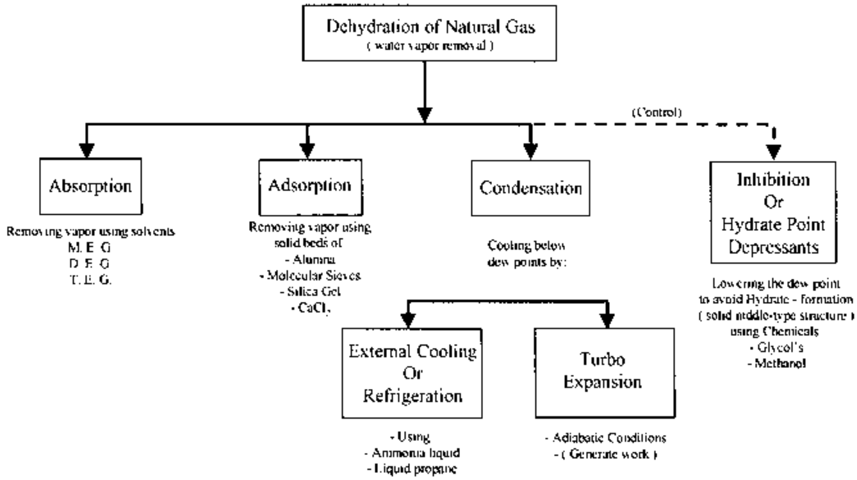

Figure 4 Classification of gas dehydration methods. (From Ref 3.)

2. Absorption, which is defined as the transfer of a component from the gas phase to the liquid phase, is more favorable at a *lower temperature* and *higher pressure*. This result is concluded by considering the following relationship (which is a combination of Raoult's law and Dalton's law):

$$\frac{P_i}{P} = \frac{Y_i}{X_i} = K_i$$

where P_i is the pressure of pure component i, P is the total pressure of the gas mixture (system), X_i is the mole fraction of component i in the liquid phase, Y_i is the mole fraction of component I in the vapor phase, and K_i is the equilibrium constant, increasing with temperature and decreasing with pressure.

Now, if temperature decreases (where P is constant), P_i decreases, which means that the water vapor concentration in the gas, Y_i, decreases, allowing more absorption of water in the liquid phase. The same conclusion is reached if the total pressure P is increased.

3. The actual absorption process of water vapor from the gas phase using glycol is dynamic and continuous. Therefore, the gas flow cannot be stopped to let a vapor and the liquid reach an equilibrium condition. Accordingly, the system under considera-

tion must be designed to allow for a close approach to equilibrium while the flow continues. Two means are provided to accomplish this task for a countercurrent flow of the feed natural gas and liquid (solvent or glycol):

(i) Tray column, or stage wise operation (equilibrium concept)
(ii) Packed column or continuous-contact operation (rate concept)

This countercurrent system allows for the "wet" gas to enter the bottom of the column and contact the *rich* glycol (high water content) at its exit point. On the other hand, as the gas works its way up the column, it encounters the *leanest* glycol (lowest water content) before the gas leaves the column.

12.4.2 Absorption System

The absorption process is shown schematically in Figure 5. The wet natural gas enters the absorption column (glycol contactor) near its bottom and flows upward through the bottom tray to the top tray and out at the top of the column. Usually six to eight trays are used. Lean (dry) glycol is fed at the top of the column and it flows down from tray to tray, absorbing water vapor from the natural gas. The rich (wet) glycol leaves from the bottom of the column to the glycol regeneration unit. The dry natural gas passes through mist mesh to the sales line.

The glycol regeneration unit is composed of a reboiler where steam is generated from the water in the glycol. The steam is circulated through the packed section to strip the water from glycol. Stripped water and any lost hydrocarbons are vented at the top of the stripping column. The hydrocarbon losses are usually benzene, toluene, xylene, and ethyl benzene (BTXE) and it is important to minimize these emissions. The rich glycol is preheated in heat exchangers, using the hot lean glycol, before it enters the still column of the glycol reboiler. This cools down the lean glycol to the desired temperature and saves the energy required for heating the rich glycol in the reboiler [4].

12.4.3 Pressure and Temperature Considerations

The absorption process improves at higher pressures because the higher-pressure gas will contain less water vapor as compared to a lower-pressure gas at the same temperature. The effect of pressure on the process, however, is not significant for pressures below about 3000 psi. However, the gas

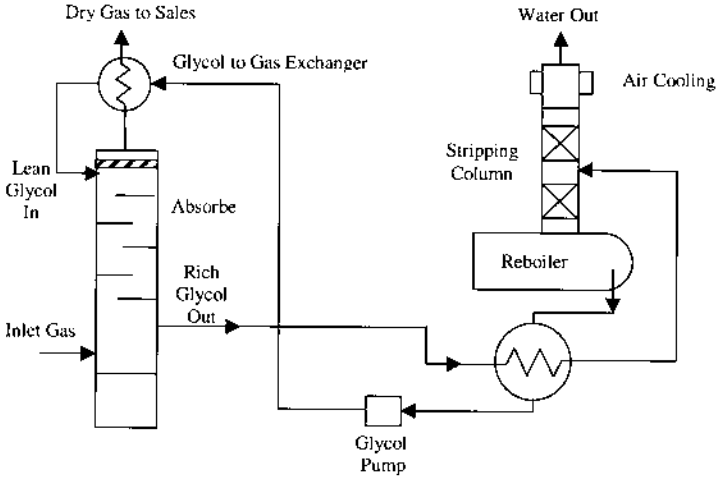

Figure 5 Flow diagram of TEG dehydration.

pressure should not be too high, as this increases the pressure rating of the column and, consequently, increases its cost. A high operating pressure would also require high glycol pumping power. On the other hand, if the gas pressure is too low, the column size would be too large. Normally, most operations are designed at pressures between 1000 and 1200 psi.

Glycol regeneration is better achieved at lower pressures. Usually, the glycol regeneration takes place at atmospheric pressure. In some cases, the process takes place under vacuum to achieve higher lean glycol concentrations; this, however, makes the system too complicated and very expensive.

The inlet gas temperature should not be too low in order to avoid condensation of water vapor and hydrocarbons. Also a low gas temperature means a low glycol temperature. At low temperatures (below 50°F), glycol becomes too viscous and more difficult to pump. Also, at low temperatures (below 60–70°F), glycol can form a stable emulsion with the hydrocarbon in the gas and may also cause foaming. On the other hand, high gas temperatures increase the gas volume, thus requiring a large-size column, and increase the water vapor content of the gas. Also, a high gas temperature results in high glycol losses. The inlet glycol temperature should not be lower than the gas temperature in order to avoid condensation of water and hydrocarbons. Normally, the gas temperature is maintained between 80°F and 110°F. The inlet glycol temperature is normally kept at about 10°F above the exiting gas temperature.

In the glycol regenerator, the glycol temperature is normally raised up to between 370°F and 390°F. This results in a lean glycol concentration of about 98.5–98.9%. A higher temperature will cause degradation of the glycol. To achieve higher glycol concentrations, stripping gas may be used.

12.4.4 Process Design

The procedure for determining the absorber column diameter and height along with solved example for a case study are presented below.

1. **Column diameter.** The column diameter can be calculated using the Souder–Brown correlation:

$$U_{\max} = C_{SB}\left(\frac{\rho_l - \rho_v}{\rho_v}\right)^{1/2} \tag{3}$$

where $U_{\max}$ is the maximum gas superficial velocity (ft/h), C_{SB} is the Souder–Brown coefficient (ft/h) (assume a value 660), ρ_l is the glycol density (lb/ft^3), and ρ_v is the gas density (lb/ft^3).

The cross-section area of the column can be calculated as

$$A = \frac{V}{U_{\max}} \tag{4}$$

where V is the gas flow rate at operating conditions.

$$V = \frac{nZRT}{P} \tag{5}$$

The compressibility factor Z can be calculated from the corresponding state method as

$$Z = 1 + \frac{B_r P_r}{T_r} \tag{6}$$

$$B_r = B^\circ + wB'$$

$$B^\circ = 0.083 - \frac{0.422}{T_r^{1.6}}$$

$$B' = 0.139 - \frac{0.172}{T_r^{4.2}}$$

$$P_r = \frac{P}{P_c} \quad \text{(reduced pressure of natural gas)}$$

$$T_r = \frac{T}{T_c} \quad \text{(reduced temperature of natural gas)}$$

Thus the diameter is

$$D = \sqrt{\frac{4A}{\pi}} \tag{7}$$

2. **Column height.** The number of theoretical plates can be calculated using the McCabe–Thiele method. The distribution of water between TEG and natural gas (equilibrium line) is almost linear, as shown in Figure 6:

$$Y = 0.1042X \tag{8}$$

where Y is the water content in the gas and X is the rich glycol concentration. If the operating line can also be assumed to be linear and the tray efficiency (η) can be assumed to be 25%, then the actual number of trays can be calculated. If the tray spacing is assumed to be 2 ft, the height H is calculated as

$$H = \frac{N(\text{spacing})}{\eta}$$

Example 4 (Case Study)

Calculate the diameter and height for an absorption column for the dehydration of "wet" natural gas under the following operating conditions:

(a) Natural gas flow rate = 98 MMSCFD
(b) Saturated with water at 1000 psig, 100°F gas

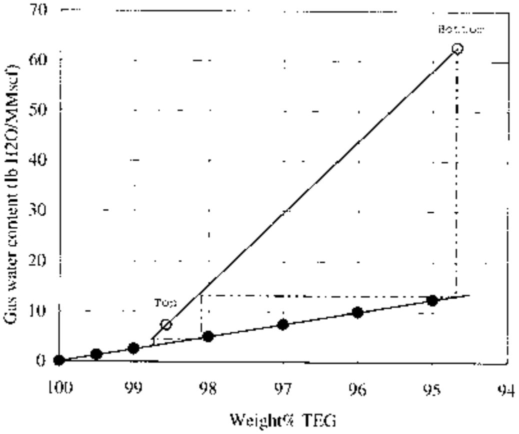

Figure 6 McCabe–Thiele diagram.

(c) Target gas water content = 7 lb/MMSCF
(d) Use TEG for dehydration (98.5% purity, the balance is water)
(e) No stripping gas used

Solution

Absorber Diameter

Calculate Z, the compressibility factor:

$T_r = 1.6317$, $P_r = 1.5217$
$B^\circ = -0.1100$, $B' = 0.1170$
$B_r = -0.1092$

$$Z = 1 + \frac{(-0.1092)(1.5217)}{1.6317} = 0.898$$

Calculate the volumetric flow rate, V:

$PV = nZRT$

$$V = \frac{(98 \times 10^6/379.5)(0.9)(10.7)(560)}{1015}$$

$$V = 953\ \frac{\text{ft}^3}{\text{min}}$$

Applying the Souder–Brown equation;

$$U_{\max} = 660\left(\frac{(70 - 3.79)}{3.79}\right)^{1/2} = 2759\ \frac{\text{ft}}{\text{h}} = 46\,\frac{\text{ft}}{\text{min}}$$

The cross-section area of the column is now found to be

$$A = \frac{953}{46} = 20.7\,\text{ft}^2$$

$$D = \sqrt{\frac{4(20.7)}{\pi}} = 5.1\,\text{ft}$$

Absorber height

To calculate the number of theoretical plates, first plot the operating line:

At tower bottom, calculate Y_1 from Figure 6:
Y_1 = water content of inlet gas = 63 lbs H_2O/MMSCF
Calculate X_1 from column material balance, where X_1 is the rich glycol weight percent
At tower top,
Y_2 = water content in exit gas = 7 lbs H_2O/MMSCF (given)

X_2 = lean glycol weight percent = 98.5 (given)

$$V(Y_1 - Y_2) = L(X_2 - X_1)$$

where L is the circulation (ft^3/h).

If we assume that 2 gal TEG/lb H_2O are needed, then the TEG circulation rate is found to be

$$\begin{aligned} L &= G(Y_1 - Y_2)(2) \\ &= \frac{98}{\text{day}}\left(\frac{1\,\text{day}}{24\,\text{h}}\right)\left(\frac{1\,L}{60\,\text{min}}\right)(63-7)(2\,\text{gal}) \\ &= 7.6\,\frac{\text{gal}}{\text{min}}\ (\text{gpm}) \end{aligned}$$

We can find X_1 from the material balance equation:

$$98\,\text{MMSCD}\ (Y_1 - Y_2) = (X_2 - X_1)\frac{\text{ft}^3}{\text{day}}$$

$$\begin{aligned} 98(63-7) &= \frac{\text{ft}^3}{9.5}\left(\frac{7.6\ \text{gal}}{\text{min}}\right)\left(\frac{60\ \text{min}}{1\ \text{h}}\right)\frac{24\,\text{h}}{1\,\text{day}}(98.5 - X_1) \\ 5488 &= 1459.2(98.5 - X_1) \\ 3.761 &= 98.5 - X_1 \\ X_1 &= 98.5 - 3.76 \\ &= 94.7 \end{aligned}$$

From Figure 6, we can find the theoretical number of stages: $N = 1.5$ stages. Assuming 25% efficiency,

$$N_{\text{act}} = \frac{1.5}{0.25} = 6\,\text{stages}$$

Assuming a spacing of 2 ft, the height H is found to be 12 ft.

The dew point of the exit sales gas can be estimated from Figure 2 and found to be equal to −10°F.

12.5 ADSORPTION: SOLID-BED DEHYDRATION

When very low dew points are required, solid-bed dehydration becomes the logical choice. It is based on fixed-bed adsorption of water vapor by a selected desiccant. A number of solid desiccants could be used such as silica gel, activated alumina, or molecular sieves.

The properties of these materials are shown in Table 2. The selection of these solids depends on economics. The most important property is the

Table 2 Properties of Solid Desiccants

Desiccant reference	Silica gel	Activated alumina	Molecular sieves
Pore diameter (Å)	10–90	15	3, 4, 5, 10
Bulk density (lb/ft^3)	45	44–48	43–47
Heat capacity (Btu/lb°F)	0.22	0.24	0.23
Minimum dew point (°F)	-60 to -90	-60 to -90	-100 to -300
Design capacity (wt%)	4–20	11–15	8–16
Regeneration stream temp. (°F)	300–500	350–500	425–550
Heat of adsorption (Btu/lb)	—	—	1800

Source: Ref. 5.

capacity of the desiccant, which determines the loading design expressed as the percentage of water to be adsorbed by the bed. The capacity decreases as temperature increases.

12.5.1 Operation of Solid-Bed Dehydrator

The system may consist of two-bed (as shown in Fig. 7), three-bed, or multibed operation. In the three-bed operation, if two beds are loading at different stages, the third one would be regenerated. The feed gas in

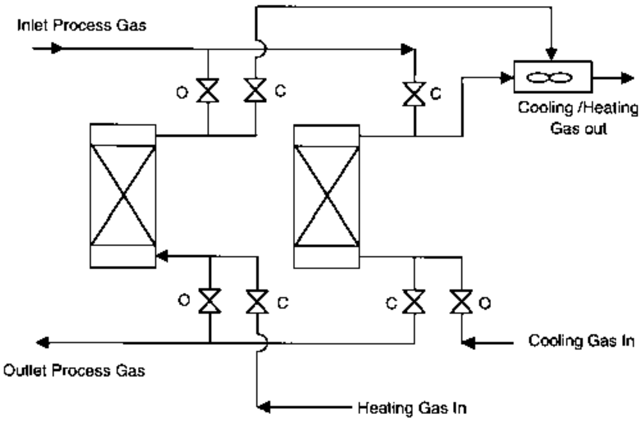

Figure 7 Solid-bed dehydration process.

entering the bed from the top and the upper zone becomes saturated first. The second zone is the mass transfer zone (MTZ) and is being loaded. The third zone is still not used and active. The different saturation progress and representation of different zones is shown in Figure 8. While the bed is in operation, the outlet concentration has very low water concentration (C_s) and the MTZ moves downward. At a certain point, the outlet water content rises to the point that is equivalent to the initial wet gas content (C_0) as if bed is not present. Thus, the beginning of this period is called the breakthrough period (θ_B). This situation is shown in Figure 9.

The operation of the process is controlled by opening valve (O) and closing valve (C), which are shown in Figure 9. After the bed has been used and loaded with water, then it is regenerated by hot gas (say 6 h, as heating time θ_H) and then cooled by switching to cold gas (say for 2 h, θ_C). The approximate temperature profile for inlet gas temperature during heating and cooling of the bed (T_i) to desorb water, and then to cool the bed to prepare it for next cycle and the outlet gas temperature (T_0) in the same period are shown in Figure 10.

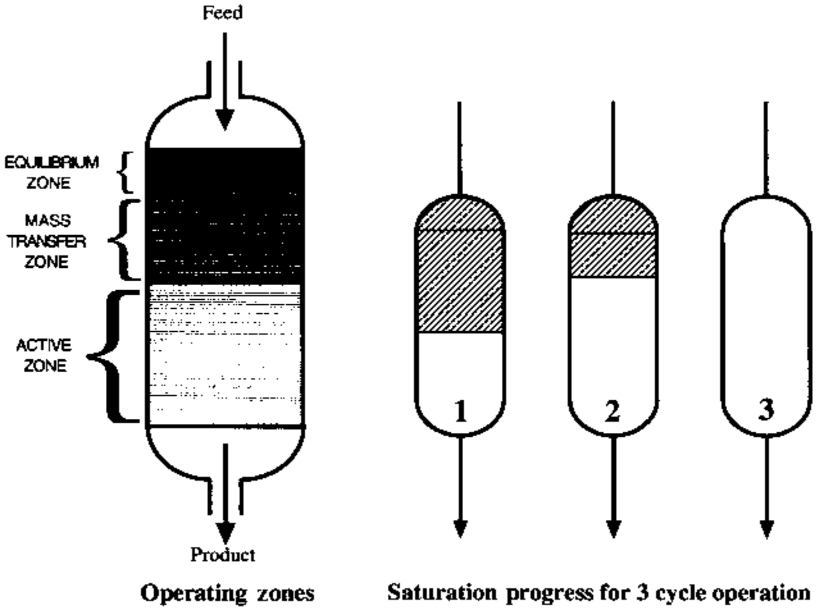

Figure 8 Mode operation.

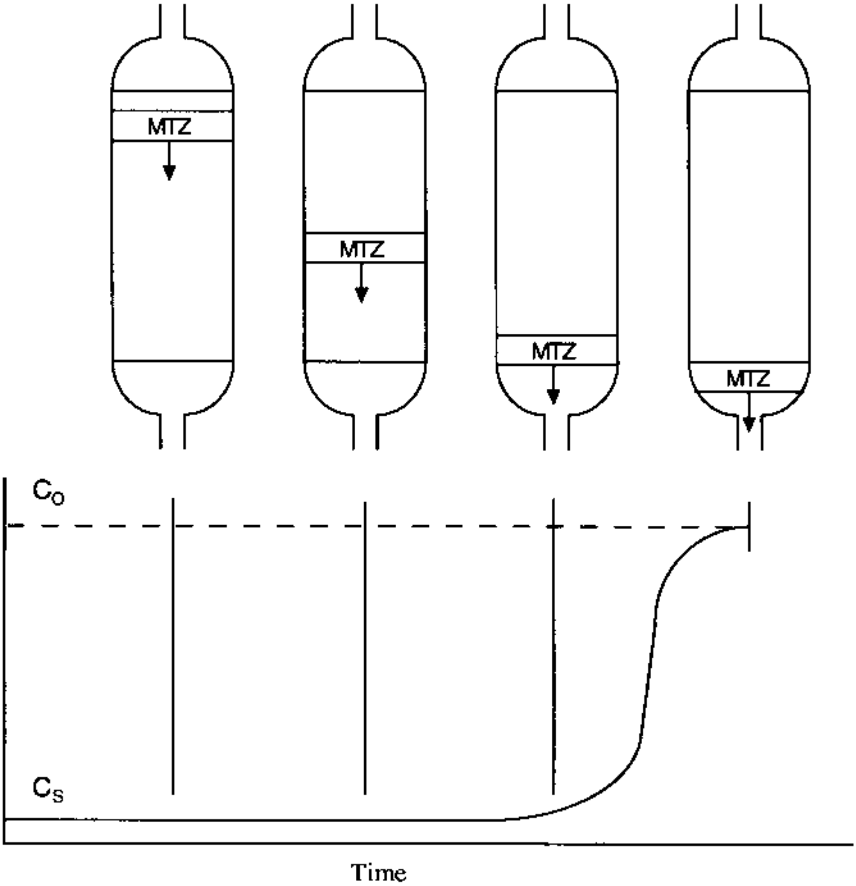

Figure 9 Breakthrough diagram in a fixed bed.

12.5.2 Design of Adsorber for Gas Dehydration

Usually, the given process conditions are gas flow rate V (MMSCF/d), temperature T (°R), gas water content Y_1 (lbs H_2O/MMSCF), desiccant particle size D_P (ft), gas density ρ_g (lbs/ft^3), and operating pressure P (psia). The maximum useful desiccant capacity X_C (lbs $H_2O/100$ lbs desiccant) is given by the vendor it will change if the storage conditions are not good. It should be assumed and then verified in the calculations. The bed operation cycle time is usually assumed as 8 h; thus, one day has three cycles.

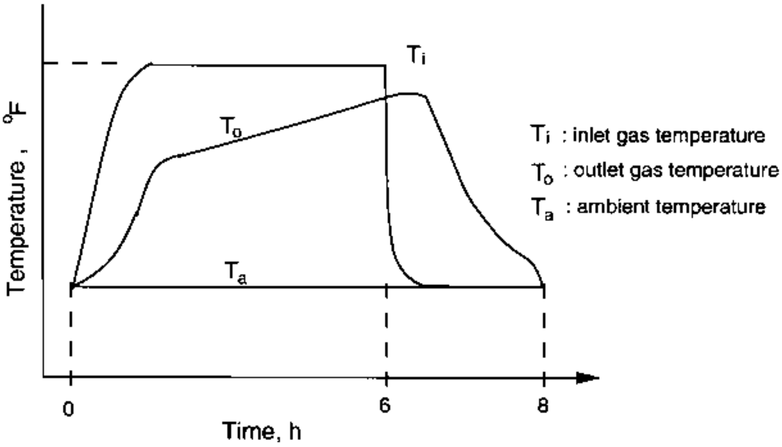

Figure 10 Temperature profile for a two-tower operation.

1. For given gas flow rate V (MMSCF/d) and water content Y_1 (lbs H_2O/MMSCF), calculate the water to be adsorbed from the gas as

$$\frac{\text{Water adsorbed}}{\text{Cycle}} = \frac{VY_1}{3} \tag{9}$$

Assuming three cycles, each cycle is 8 h in one day.

2. If the particle size of the bed is known D_P (ft), calculate the superfacial velocity

$$V_g\left(\frac{\text{ft}}{\text{min}}\right) \quad as \quad V_g = 540\left(\frac{D_P}{\rho_g}\right)^{1/2} \tag{10}$$

where ρ_g is the gas density (lb/ft^3).

3. Calculate the bed diameter d (ft) from

$$d = \left(\frac{25VZ_gT_g}{PV_g}\right)^{1/2} \tag{11}$$

4. Calculate the water loading q (lbs/ft^{2}h) using

$$q = 0.053\left(\frac{VY_1}{d^2}\right) \tag{12}$$

5. Calculate the mass transfer zone height h_z (ft) using

$$h_z = 31.25\left(\frac{q^{0.7895}}{V_g^{0.5506}\mathrm{RS}^{0.2646}}\right) \tag{13}$$

where RS is relative saturation.

6. Calculate the bed length h_B(ft). The bed weight W_B(lb) is calculated using

$$W_B = \frac{\text{Amount of water adsorbed (lbs H}_2\text{O)}}{\text{Capacity of desiccant (lbs H}_2\text{O/lb desiccant)}}$$

$$W_B = \frac{VY_1}{3}X_C^{-1} \tag{14}$$

The desiccant density ρ_B(lbs/ft^3) is used to calculate bed volume V_B(ft^3) using

$$V_B = \frac{W_B}{\rho_B} \tag{15}$$

The bed length h_B(ft) is calculated using

$$h_B = \frac{V_B}{\pi d^2/4} \tag{16}$$

7. The maximum useful desiccant capacity (lbsH$_2$O/100 lbs desiccant) is calculated from

$$X_C = \frac{X'_S(h_B - 0.45\,h_Z)}{h_B} \tag{17}$$

where X'_S is the dynamic desiccant capacity at saturation and is a function of temperature as

$$X'_S = X_S(1.518 - 0.00673\,T_B) \tag{18}$$

X_S is the dynamic capacity at 77°F and T_B(°F) is the bed temperature and can be calculated using gas relative saturation (RS) at 77°F:

$$X_S = 7.407 + 0.09357\,\mathrm{RS} \tag{19}$$

The calculated value of X_C from Eq. (17) is compared with the assumed value and is readjusted so that assumed and calculated values become close.

8. Calculate the breakthrough time θ_b from:

$$\theta_B = \frac{0.01X_C\rho_B h_B}{q} \tag{20}$$

The cycle time θ_C should be greater than the breakthrough time:

$$\theta_C > \theta_b$$

9. Check pressure drop across the bed for a $\frac{1}{8}$-in. particle size of packed material, the pressure drop can be calculated as

$$\frac{\Delta P}{h_B} = 0.056\,\overset{*}{\mu} V_g + 0.0000889 + V_g^2 \tag{21}$$

It is recommended that the pressure drop should be less than 8 psia. The design procedure is illustrated in Figure 11.

Example 5

Given that 35 MMSCF of wet gas is fed into a packed bed of desiccant. The gas molecular weight is 18.0 and its density is 1.5 lbs/ft^3. The operating temperature is 110°F and pressure is 500 psia. The inlet dew point is 100°F and the desired outlet dew point is (1 ppm H_2O).

Solution

Assuming

$$X_C = 0.12$$

$$W_B = \frac{35(90)}{3(0.12)} = 8750$$

$$V_B = \frac{8750}{45} = 195\ \text{ft}^3$$

Assume $\frac{1}{8}$-in. particle size.

$$D_P = \frac{1}{8(12)}$$

$$V_g = 540\left(\frac{1}{8(12)(1.5)}\right)^{1/2} = 45\ \frac{\text{ft}}{\text{min}}$$

$$d = \left(\frac{25(35)(1)(70)}{600(45)}\right)^{1/2} = 3.9\ \text{ft}$$

$$h_B = \frac{V_B}{\pi d^2/4)} = \frac{195(4)}{\pi(3.9)^2} = 16.3\ \text{ft}$$

$$q = 0.053\left(\frac{GW}{d^2}\right) = \frac{35(90)(0.053)}{(3.9)^2}$$

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[*] μ is the gas viscosity in centipoise, Cp.

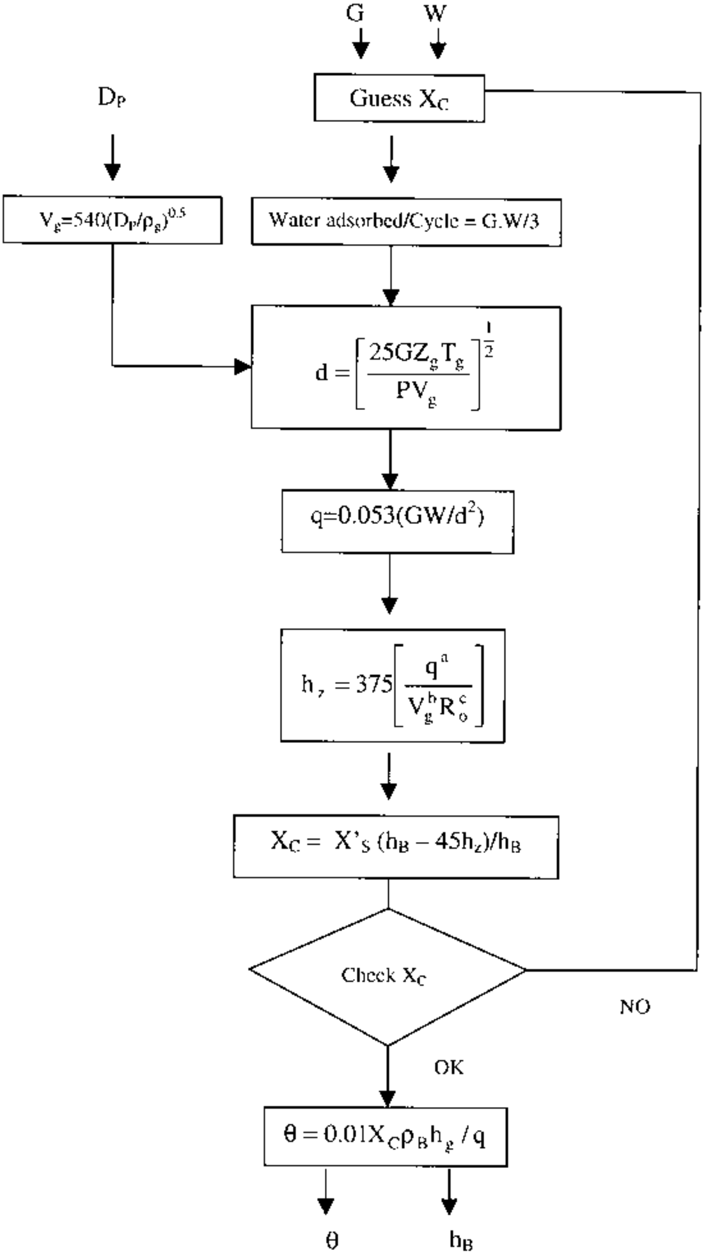

Figure 11 Design procedure of solid dehydrator design.

$$h_Z = 31.25\left(\frac{11^{0.7895}}{(45)^{0.5505}(100)^{0.2646}}\right) = 7.5\,\text{ft}$$

$$X_C = \frac{X'_S(h_B - 0.45\,h_z)}{h_B}$$

$$X_S = 7.407 + 0.09357(100) = 16.764$$
$$X'_S = 16.764[1.518 - 0.00673(100)] = 14.17$$
$$X_C = \frac{14.17[16.3 - 0.45(7.5)]}{16.3}$$
$$= 11.236\,\text{lbs H}_2\text{O}/100\,\text{lbs solid}$$

or

$$X_C = 0.1124\,\text{lbs H}_2\text{O/lb solid (per lb solid)}$$

Initially, we assumed $X_C = 0.12$, which is close enough.

$$\theta_B = \frac{0.01 X_C \rho_B h_B}{q}$$

$$\theta_B = \frac{0.01(11.2)(45)(16.3)}{11} = 7.468$$

So, assuming three cycles each 8 h will be fine.

The pressure drop ΔP (psia) is calculated as

$$\frac{\Delta P}{h_B} = 0.0560\,\mu V_g + 8.89 \times 10^{-5} \rho V_g^2$$

$$\Delta P = 0.0560(0.01)(45) + 8.89 \times 10^{-5}(1.5)(45)^2(16.3) = 4.8\,\text{psi}$$

which is less than 8 psia (recommended).

12.5.3 Regeneration Heat Requirements

The heat required during the heating part of the cycle is

$$Q_H = m_d(T_m - T_a)C_d + m_1(T_m - T_a)C_t + w_1 \Delta H_d + w_1(T_{av} - T_a) \tag{22}$$

$$T_{av} = \frac{T_m + T_a}{2}$$

where

w_1 = water adsorbed/cycle (lbs/cycle)
Q_H = heat required/cycle (Btu/cycle)
M_d, C_d = mass and specific heat of desiccant
M_t,C_t = mass and specific heat of tower
ΔH_d = heat of water desorbed (given) (Btu/lb)
T_m = maximum outlet temperature (°F)
T_a = ambient temperature (°F)

To calculate gas flow rate V_H (lbs/cycle) required to desorb water, you need to raise heating gas to T_H from ambient temperature T_a to supply Q_H.

Thus,

$$Q_H = V_H(T_H - T_{av})C_g$$

where C_g is specific heat of gu; or

$$V_H = \frac{Q_H}{(T_H - T_{av})C_g} \frac{\text{lbs}}{\text{cycle}} \tag{23}$$

The gas heater load, Q_G (Btu/cycle), is

$$Q_G = V_H(T_H - T_a)C_g \frac{\text{Btu}}{\text{cycle}}$$

If the heating period is θ_H, h for each cycle is then

$$Q_G = \frac{V_H(T_H - T_a)C_g}{\theta_H} \frac{\text{Btu}}{\text{h}} \tag{24}$$

where Q_G is the gas heater load and θ_H is the heating time per cycle. Usually, 25% excess load is assumed for losses.

For cooling part cooling requirements, H_C (Btu/cycle),

$$H_C = m_d(T_m - T_a)C_d + m_t(T_m - T_a)C_t \tag{25}$$

and the gas flow rate for cooling, V_C(lbs/cycle),

$$V_C = \frac{H_C}{(T_{av} - T_a)C_g} \frac{\text{lbs}}{\text{cycle}}$$

or

$$V_C = \frac{H_C}{(T_{av} - T_a)C_g\theta_C} \frac{\text{lbs}}{\text{h}} \tag{26}$$

where θ_C is the cooling time, (h/cycle)

Example 6

Find the flow rate of a heating gas at 400°F to desorb 2000 lbs of water from a solid desiccant bed weighing 30,000 lbs and has $C_d = 0.24$ Btu/lb °F. The tower weight is 50,000 lbs and its $C_t = 0.1$ Btu/lb °F. The ambient temperature is 110°F and the maximum outlet temperature is 360°F. The 8-h cycle is operated as 6 h heating and 2 h cooling. Assume $C_g = 0.6$ Btu/lb °F.

Solution

Using Eq. (22), the following results are obtained:

$$T_m = 360, \qquad T_a = 110, \qquad T_{av} = \frac{360 + 110}{2} = 235$$

$$\begin{aligned} Q_H = {} & 30{,}000(360 - 110)(0.24) + 50{,}000(360 - 110)(0.1) \\ & + 2000(1100) + 2000(235 - 110) \\ = {} & 5{,}500{,}000 \text{ Btu/cycle} \end{aligned}$$

The flow rate of heating gas V_H is obtained from Eq. 23:

$$V_H = \frac{5{,}500{,}000}{(400 - 235)(0.6)} = 55{,}555 \,\frac{\text{lbs}}{\text{cycle}}$$

or

$$V_H = 9259 \,\frac{\text{lbs}}{\text{h}}$$

The gas heater load Q_G is obtained from Eq. (24):

$$Q_G = \frac{55{,}555(400 - 110)(0.6)}{6} = 1.6 \times 10^6 \,\frac{\text{Btu}}{\text{lb}}$$

For cooling, use Eq. (25) to find H_C:

$$\begin{aligned} H_C &= 30{,}000(360 - 110)(0.24) + 50{,}000(360 - 110)(0.1) \\ &= 3.05 \times 10^6 \,\frac{\text{Btu}}{\text{cycle}} \end{aligned}$$

The gas flow rate for cooling is calculated from Eq. (26); for 2 h cooling, the value of V_C is found to be

$$V_C = \frac{3.05 \times 10^6}{(235 - 110)(0.6)(2)} = 20{,}233 \,\frac{\text{lbs}}{\text{h}}$$

Additional heat requirements of 25% can be used for each cycle.

12.6 SIMULATION PROBLEM FOR TEG DEHYDRATOR

In a gas field, natural gas is available at a flow rate of 10 MMSCFD and at operating conditions of 900 psia and 80°F. The gas composition is given as follows (mole fractions):

N_2	0.001	C_3H_8	0.0148		
CO_2	0.0284	iC_4H_{10}	0.0059		
H_2S	0.0155	nC_4H_{10}	0.0030	TEG	0.000
CH_4	0.898	iC_5H_{12}	0.001	H_2O	0.000
C_2H_6	0.031	nC_5H_{12}	0.005		

The operating and design conditions of the TEG absorber are:

- The flow rate of the TEG fed to the column is 2 gpm (US gallon) at 900 psia and 120°F.
- Number of trays are 14; with a tray efficiency 50% except the top and bottom trays are 100% each.

SOLUTION

Using HYSYS process simulator licensed by Hypotech, the process flow sheet for the proposed dehydration plant along with material and energy balance calculations for all streams, are produced as presented in Figure 12 and Table 3.

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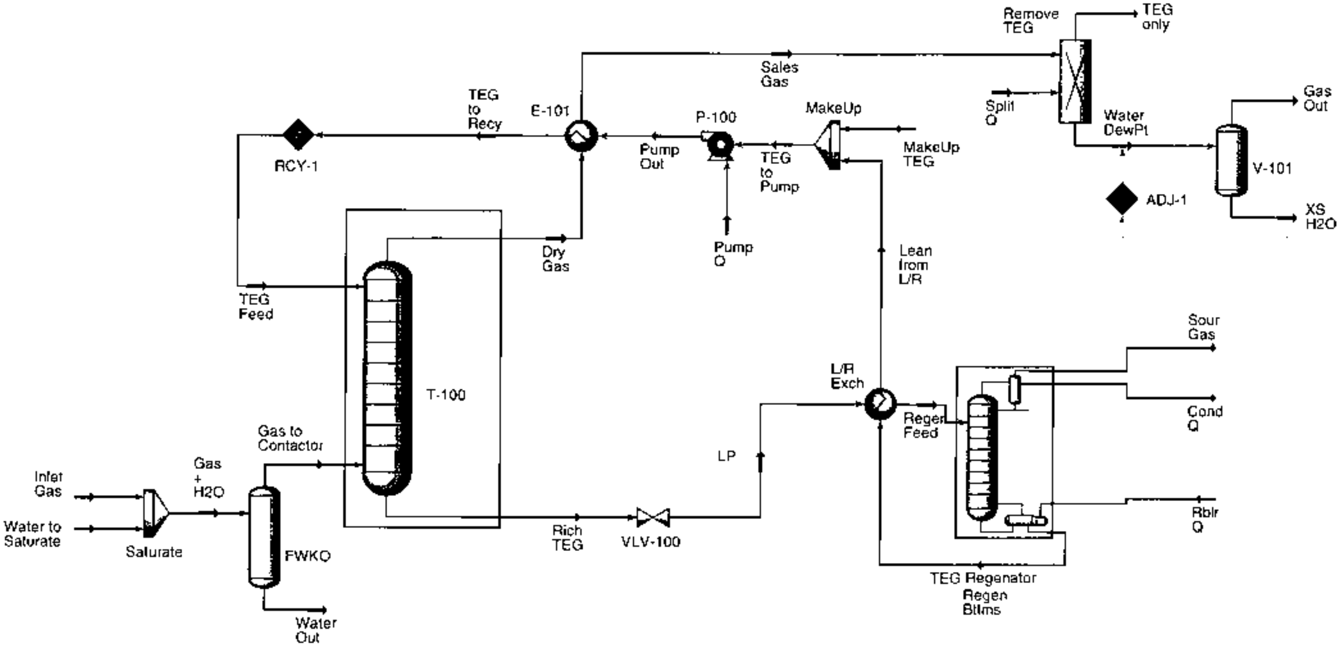


Figure 12 Computer simulation of TEG dehydration.

Table 3 Stream Material and Energy Balance for Flow Sheet in Figure 11

Name		Inlet Gas	TEG Feed	Water to Saturate	Gas + H_2O	Gas to Contactor
Vapour Fraction		1.0000	0.0000	0.5420	0.9999	1.0000
Temperature	(C)	29.44*	48.89*	277.1	29.44*	29.44
Pressure	(bar)	62.05*	62.05*	62.05	62.05	62.05
Molar Flow	(kgmole/h)	498.1*	3.649*	0.4990*	498.6	498.5
Mass Flow	(kg/h)	9183.	512.0	8.989	9192.	9191.
Liquid Volume Flow	(m^3/h)	27.58	0.4543	0.009007	27.59	27.59
Heat Flow	(kcal/h)	−1.018e+07	−6.562e+05	−2.961e+04	−1.021e+07	−1.021e+07
Name		Water Out	Dry Gas	Rich TEG	LP TEG	Regen Bttms
Vapour Fraction		0.0000	1.0000	0.0000	0.0362	0.0000
Temperature	(C)	29.44	31.14	30.17	35.51	204.4
Pressure	(bar)	62.05	62.05	62.05	1.793	1.034
Molar Flow	(kgmole/h)	0.06651	497.9	4.270	4.270	3.648
Mass Flow	(kg/h)	1.200	9177.	525.6	525.6	511.9
Liquid Volume Flow	(m^3/h)	0.001203	27.57	0.4727	0.4727	0.4542
Heat Flow	(kcal/h)	−4527.	−1.017e+07	−6.949e+05	−6.949e+05	−6.071e+05
Name		Lean from L/R	Regen Feed	Sour Gas	MakeUp TEG	TEG to Pump
Vapour Fraction		0.0000	0.0531	1.0000	0.0000	0.0000
Temperature	(C)	145.1	104.4*	101.7	15.56*	145.1
Pressure	(bar)	1.027	1.103*	1.014	1.027	1.027
Molar Flow	(kgmole/h)	3.648	4.270	0.6216	0.0004591	3.649
Mass Flow	(kg/h)	511.9	525.6	13.71	0.06834	512.0
Liquid Volume Flow	(m^3/h)	0.4542	0.4727	0.01854	6.057e−05	0.4543*
Heat Flow	(kcal/h)	−6.283e+05	−6.737e+05	−2.861e+04	−87.51	−6.264e+05

(*Continued*)

Table 3 Continued

Name		Inlet Gas	TEG Feed	Water to Saturate	Gas + H_2O	Gas to Contactor
Name		Pump Out	TEG To Recy	Sales Gas	TEG only	Water Dew Pt
Vapour Fraction		0.0000	0.0000	1.0000	—	1.0000
Temperature	(C)	144.0	48.89*	36.19	—	−13.56*
Pressure	(bar)	62.74*	62.05	61.71	—	62.05*
Molar Flow	(kgmole/h)	3.649	3.649	497.9	0.0001055	497.9
Mass Flow	(kg/h)	512.0	512.0	9177.	0.01584	9177.
Liquid Volume Flow	(m^3/h)	0.4543	0.4543	27.57	1.403e−05	27.57
Heat Flow	(kcal/h)	−6.274e+05	−6.562e+05	−1.014e+07	—	−1.043e+07
Name		Gas Out	XS H_2O			
Vapour Fraction		1.0000	0.0000			
Temperature	(C)	−13.56	−13.56			
Pressure	(bar)	62.05	62.05			
Molar Flow	(kgmole/h)	497.9	0.0005366			
Mass Flow	(kg/h)	9177.	0.009708			
Liquid Volume Flow	(m^3/h)	27.57	9.746e−06			
Heat Flow	(kcal/h)	−1.043e+07	−36.93			

*Specified by user.

REVIEW QUESTIONS AND EXERCISE PROBLEMS

1. When it is necessary to reduce gas pressure in a gas production system, the largest pressure drop should be taken at ______________________ to minimize hydrate formation.
2. To prevent hydrate formation, the following methods may be used:

   (a) ____
   (b) ____
   (c) ____

3. Given that the hydrate formation pressure of a gas at 60°F is 1000 psia, state whether hydrate will form if the gas is under the following conditions:

   (a) $T = 60$°F and $P = 1200$ psia: YES or NO
   (b) $T = 70$°F and $P = 1000$ psia: YES or NO
   (c) $T = 50$°F and $P = 1000$ psia: YES or NO

4. A gas well produces with a high condensate–gas ratio:

   (a) Is it practical to use methanol injection to prevent hydrate formation?
   (b) What is the reason for your answer?

   ____
   ____

5. What are the conditions that must exist for hydrate formation?

   (a) ____
   (b) ____
   (c) ____

6. In a central separation and treatment plant, 100 MMSCFD of gas containing 10% H_2S are collected at 500 psia and 80°F. The gas is first treated using a DEA process to reduce its H_2S content down to 3 ppm. The sweet gas is then compressed to 1500 psia and cooled down to 110°F. The gas is then dehydrated using a glycol dehydration process which reduces its water content down to 7 lbs/MMSCF. Finally, a solid desiccant process is used to further dehydrate (polish) the gas to reduce its water content to less than 1 ppm. (See Figure 10.)

   (i) Assuming that the gas is fed to the compressor at 500 psia and 80°F and that it is saturated with water vapor, determine the amount of free water in the gas after its compression and cooling (1500 psia and 110°F).

(ii) Why is it necessary to compress and cool the gas before the dehydration process?

(iii) Determine the required diameter of the glycol contactor using $C_D = 0.852$ and $d_m = 120\,\mu\text{m}$.

(iv) Determine the required glycol circulation rate in gallons per minute and the reboiler duty assuming the lean glycol concentration to be 98.5%.

(v) Discuss the limitations on the temperatures of the glycol and gas.

(vi) Draw the process flow diagram of the solid desiccant dehydration process used for polishing the gas.

(vii) Determine the volume of the desiccant needed and the diameter and height of the absorber given that the maximum superficial velocity allowed in the tower is 40 ft/min and using a cycle time of 8 h.

13

Recovery, Separation, and Fractionation of Natural Gas Liquids

13.1 INTRODUCTION

The material presented in this chapter includes two parts: the recovery and separation of natural gas liquid (NGL) constituents, and methods of fractionation into finished product streams suitable for sale. In the first part, several alternatives for the separation and recovery of NGL are detailed. They are essentially based on phase change either by using energy separating agent (ESA) or mass separating agent (MSA). Thus, partial liquefaction or condensation of some specific NGL constituents will lead to their separation from the bulk of the gas stream. Total condensation is also a possibility. The role of the operating parameters that influence phase change, hence NGL separation from the bulk of gas, is discussed. The second part of the chapter covers materials on fractionation facilities that are recommended to produce specification quality products from NGL. Types of fractionator with recommended feed streams as well as produced products are highlighted in this part.

13.2 RECOVERY AND SEPARATION OF NGL

13.2.1 Options of Phase Change

To recover and separate NGL from a bulk of gas stream, a change in phase has to take place. In other words, a new phase has to be developed for separation to occur. Two distinctive options are in practice depending on the use of ESA or MSA.

Energy Separating Agent

The distillation process best illustrates a change in phase using ESA. To separate, for example, a mixture of alcohol and water heat is applied. A vapor phase is formed in which alcohol is more concentrated, and then separated by condensation. This case of separation is expressed as follows:

$$\text{A mixture of liquids} + \text{Heat} \rightarrow \text{Liquid} + \text{Vapor}$$

For the case of NGL separation and recovery in a gas plant, removing heat (by refrigeration) on the other hand, will allow heavier components to condense; hence, a liquid phase is formed. This case is represented as follows:

$$\text{A mixture of hydrocarbon vapor} - \text{Heat} \rightarrow \text{Liquid} + \text{Vapor}$$

Partial liquefaction is carried out for a specific cut, whereas total liquefaction is done for the whole gas stream.

Mass Separating Agent

To separate NGL, a new phase is developed by using either a solid material in contact with the gas stream (adsorption) or a liquid in contact with the gas (absorption). These two cases are represented in the following subsections.

13.2.2 Parameters Controlling NGL Separation

A change in phase for NGL recovery and separation always involves control of one or more of the following three parameters:

- Operating pressure, P
- Operating temperature, T
- System composition or concentration, x and y

To obtain the right quantities of specific NGL constituents, a control of the relevant parameters has to be carried out:

1. For separation using ESA, pressure is maintained by direct control. Temperature, on the other hand, is reduced by refrigeration using one of the following techniques:
    - (a) Compression refrigeration
    - (b) Cryogenic separation; expansion across a turbine
    - (c) Cryogenic separation; expansion across a valve

2. For separation using MSA, a control in the composition or the concentration of the hydrocarbons to be recovered (NGL); y and x is obtained by using adsorption or absorption methods.

Adsorption provides a new surface area, through the solid material, which entrains or "adsorbs" the components to be recovered and separated as NGL. Thus, the components desired as liquid are deposited on the surface of the selected solid, then regenerated off in a high concentration; hence, their condensation efficiency is enhanced. About 10–15% of the feed is recovered as liquid.

Adsorption is defined as a concentration (or composition) control process that precedes condensation. Therefore, refrigeration methods, may be coupled with adsorption to bring in condensation and liquid recovery. *Absorption*, on the other hand, presents a similar function of providing a surface or "contact" area of liquid–gas interface.

The efficiency of condensation, hence NGL recovery, is a function of P, T, gas and oil flow rates, and contact time. Again, absorption could be coupled with refrigeration to enhance condensation.

To summarize the above, a proper design of a system implies the use of the optimum levels of all operating factors plus the availability of sufficient area of contact for mass and heat transfer between phases.

13.2.3 Selected Separation Processes

In this section a brief description is given for the absorption, refrigeration, and cryogenic (Joule–Thomson turbo expansion) processes recommended to separate NGL constituents from a gas stream. Details are illustrated in the corresponding flow diagrams. Adsorption process, on the other hand, has been covered in Chapter 12.

Absorption Process

The absorption unit consists of two sections: the absorption and regeneration as illustrated in Figure 1. An upflow natural gas stream is brought in direct contact, countercurrently with the solvent (light oil in the kerosene boiling range) in the absorber. The column—a tray or packed one—operates at about 400–1000 psia and ambient or moderately subambient temperatures. The rich oil (absorbed NGL plus solvent) is directed to a distillation unit to separate and recover the NGL, whereas the lean oil is recycled back to the absorber. In addition to natural gasoline, C_3/C_4 could be recovered as well in this process. Provision is made to separate ethane from rich oil using a deethanizer column.

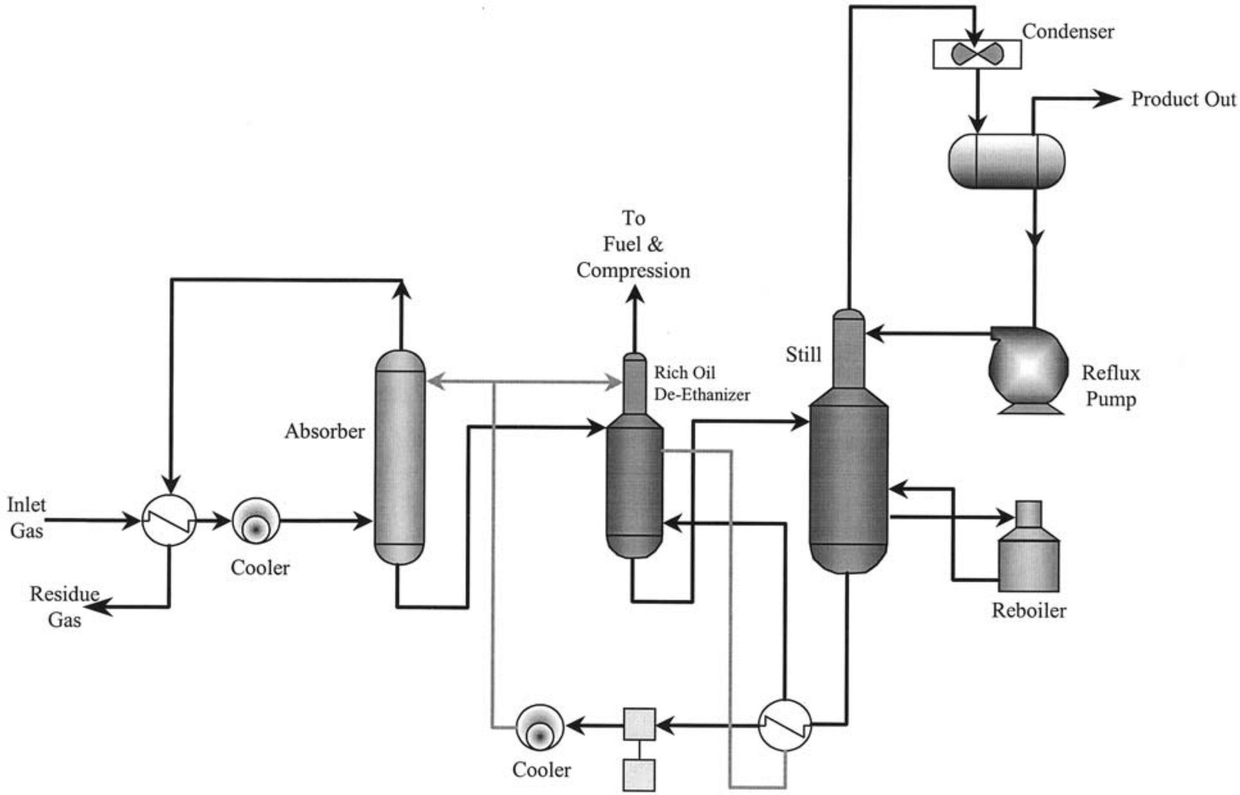


Figure 1 Separation of NGL by absorption.

Ethane recovery, however, is quite small. This process is being phased out.

Refrigeration Process

The production of NGL at low temperature is practiced in many gas processing plants in order to condense NGL from gas streams. As indicated in Figure 2, using nontoxic and noncorrosive refrigerants to chill the feed natural gas to a temperature between 0°F and −40°F using a low-level one-component refrigerant system provides external refrigeration. When using a high-level cascade refrigerant system, a much lower temperature in the range of −100°F to −150°F is reached. Liquids are separated from the residue gas at multiple temperatures and then fractionated into final products. Ethane recovery is a strong function of the operating temperatures as is explained next.

The following operating conditions are important in the separation of NGL constituents: Two main objectives are usually targeted when the desired temperature of the gas–liquid stream leaving the chiller is specified: (1) dew point control (dpc) of the hydrocarbons and (2) liquid recovery. Now, if the dpc of the hydrocarbon is the primary target, then the temperature of the gas–liquid stream is set at about 6–10°C below the desired dew point temperature. On the other hand, if the liquid recovery is the main objective, then condensing the least amount of nonsalable components should be achieved. In other words, condensing methane is not cost-effective.

As far as the operating pressure for the refrigeration system, the recommended operating pressure for maximum liquid recovery is set between 400 and 600 psia. Condensation of methane increases with higher pressure; therefore, optimum pressure must minimize the total cost of the system. In general, separation is carried out at a pressure corresponding to the sales gas pressure simply to eliminate the cost of gas recompression. To summarize the above, for a given selected separation pressure the corresponding operating temperature is chosen based on the type of product:

- If the liquid product is to be sold as "crude oil," then the separation temperature is between 0°C and 5°C.
- If the liquid product contains propane as the lightest component, temperature is about −30°C to −18°C. In this case, temperature depends on whether absorption or adsorption is combined with refrigeration.
- If the operating temperature is set below −30°C, a cryogenic range of ethane recovery is encountered.

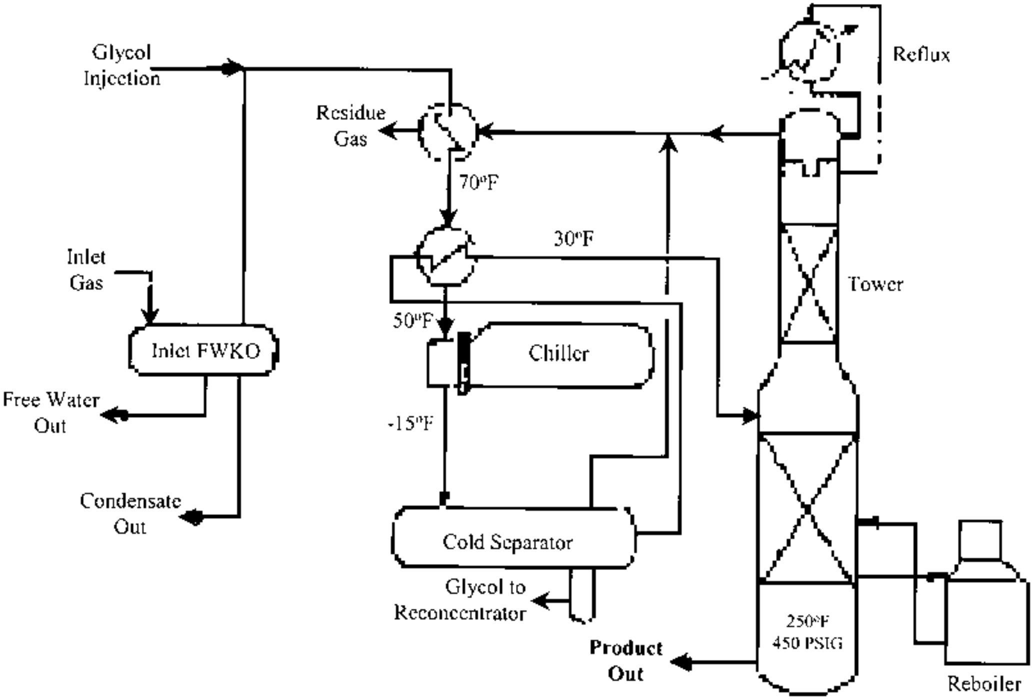


Figure 2 Separation of NGL using refrigeration plant.

Cryogenic Processes

Natural gas liquid could be separated from natural gas using two approaches based on cryogenic expansion (autorefrigeration):

- An expander plant produces refrigeration to condense and recover the liquid hydrocarbons contained in the natural gas by using a turboexpander. In this process, the enthalpy of the natural gas is converted into useful work, behaving thermodynamically as an approximate isentropic process.
- Expansion across the valve will lead to a similar result. However, the expansion is described in this case as "isenthalpic."

Temperatures produced by turboexpansion are much lower than those of valve expansion.

A schematic presentation for the turboexpansion process is presented in Figure 3. The process operates at -100°F to -160°F and 1000 psia. The process represents a new development in the gas processing industry. Increased liquid recovery (especially ethane) is an advantage of this process.

Figure 4 illustrates the condensation process using ethane/propane, followed by demethanization to produce NGL as a final product. Figure 5, on the other hand, presents a typical gas plant for the recovery and separation of NGL.

A concise comparison among the three options (absorption, refrigeration, and cryogenic separation processes) recommended for NGL recovery is given by Abdel-Aal [1].

13.3 FRACTIONATION OF NGL

13.3.1 Goals and Tasks

In general, and in gas plants in particular, fractionating plants have common operating goals:

1. The production of on-specification products
2. The control of impurities in valuable products (either top or bottom)
3. The control in fuel consumption.

As far as the tasks for system design of a fractionating facility, these goals are as follows:

1. Fundamental knowledge on the process or processes selected to carry out the separation; in particular, distillation.

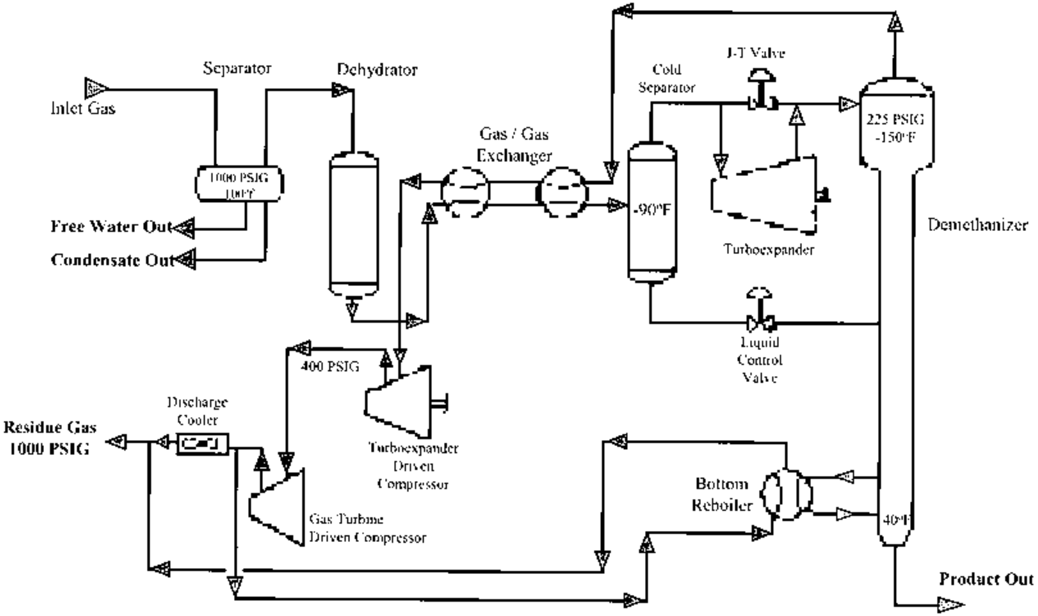


Figure 3 Cryogenic separator of NGL.

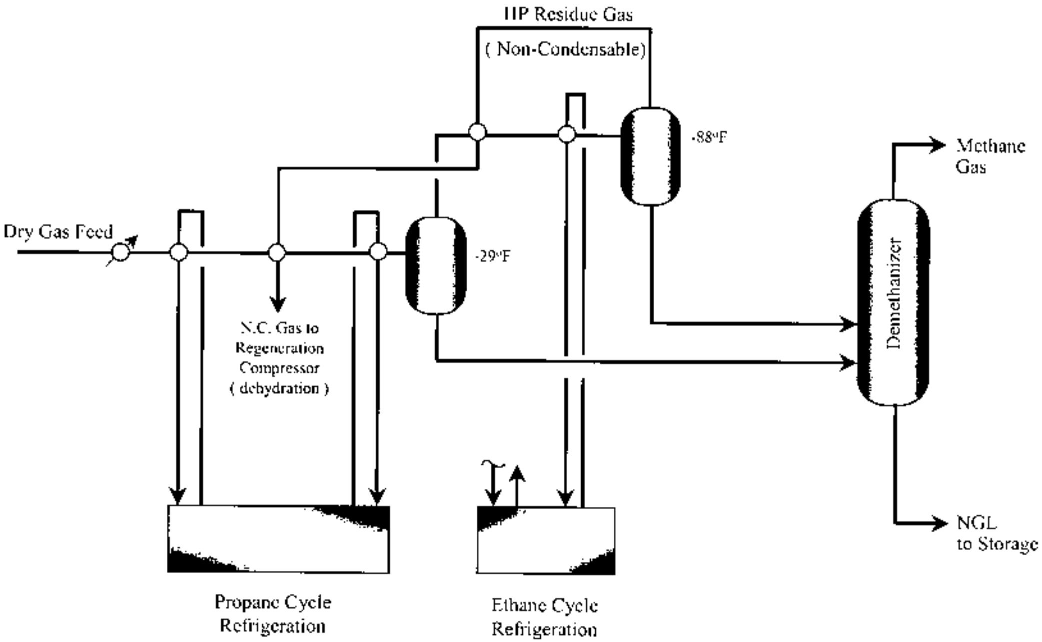


Figure 4 NGL condensation/demethanization.

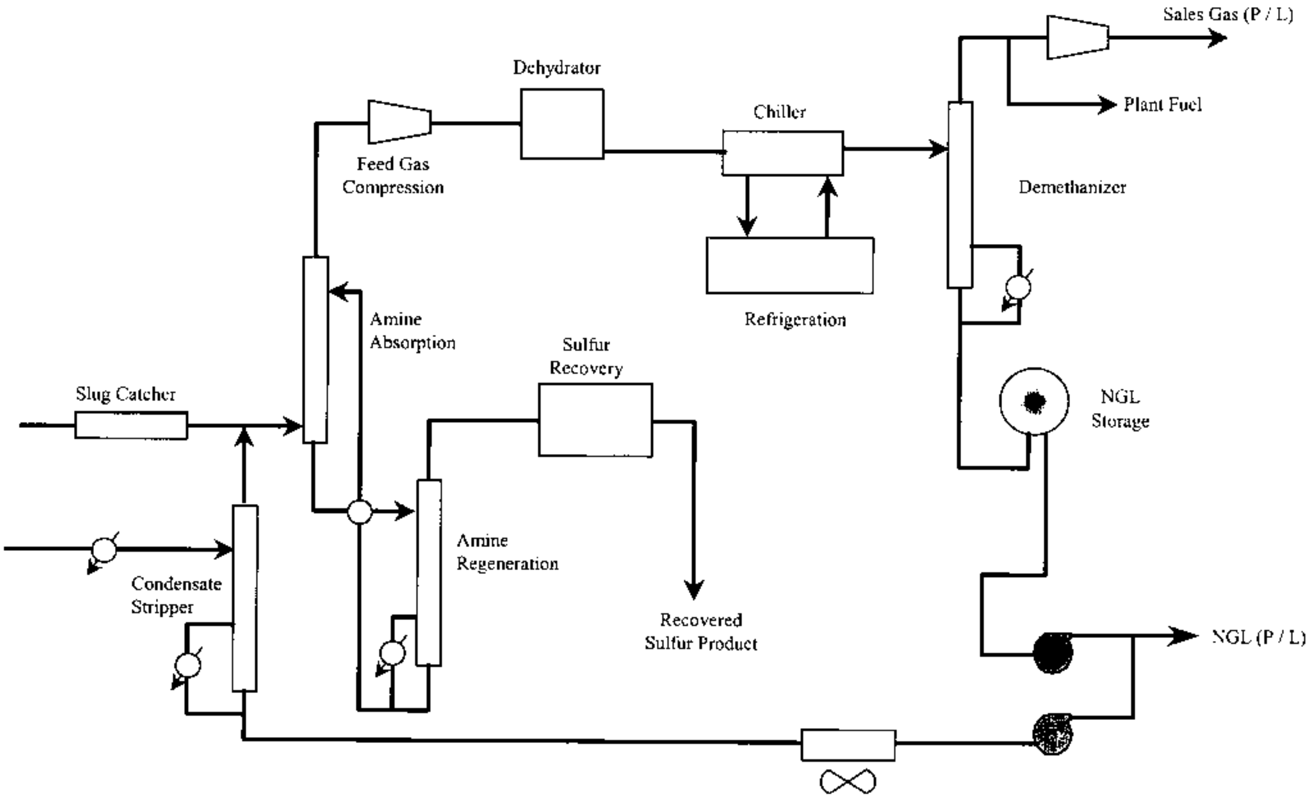


Figure 5 A typical gas plant of NGL recovery.

2. Guidelines on the order of sequence of separation (i.e., synthesis of separation sequences).

13.3.2 Fundamentals of Distillation

For separation to take place, say by distillation, the selection of an exploitable chemical or physical property difference is very important. Factors influencing this are as follows:

1. The physical property itself
2. The magnitude of the property difference
3. The amount of material to be distilled
4. The relative properties of different species or components; purity required
5. The chemical behavior of the material during distillation and its corrosiveness

A measure of the ease of separation of one component A from another B is known as the separation factor, SF, defined as

$$\mathrm{SF} = \frac{(C_A/C_B)_{\text{Top product}}}{(C_A/C_B)_{\text{Bottom product}}}$$

where C is the concentration. A high value of SF means an easy separation. A good example is the separation of salt from seawater by "evaporation." Here, the value of SF is found, by intuition, to be infinity, because we are separating water (volatile component) from salt (nonvolatile).

Consider a raw hydrocarbon stream to be fractionated, which may contain the following hydrocarbons:

Selected Hydrocarbons with Corresponding Boiling Point

Component	Boiling point °F at atmospheric pressure
Ethane	– 128
Propane	– 44
Isobutane	11
N-Butane	31
Pentanes Plus	82–250

From the above data, it is evident that in order to separate a mixture of NGL, say propane plus, as a bottom product, then there must be a difference in boiling point between the top product (ethane) and the bottom product. This difference is an indication of the degree of difficulty of separation or the value of SF. As an equilibration separation process, SF should be much greater than unity for ethane to concentrate in the top and propane plus to concentrate in the bottom.

A difficult separation implies the following:

1. Higher number of distillation trays, which affects column size
2. Higher reflux ratio, which influences pump size and power consumption
3. Additional reboiler heat duty, which influences reboiler size and energy consumption

13.3.3 Distillation Processes and Types of Fractionators

The separation of NGL may require different modes of distillation as well as other methods of separation techniques. A summary-up of different distillation processes and other separation methods is shown in Figures 6 and 7 respectively.

Fractionators of different types are commonly used in gas plants:

Type of fractionator	Feed	Top product	Bottom product
Demethanizer	C_1/C_2	Methane	Ethane
Deethanizer	LPG	Ethane	Propane plus
Depropanizer	Deethanizer bottoms	Propane	Butanes plus
Debutanizer	Depropanizer bottoms	Butanes (iso + n)	Natural gasoline (pentanes plus)
Deisobutanizer	Debutanizer top	Isobutane	Normal butane

Control of the following key operating variables will ensure efficient results of fractionation operations [2]:

1. Top tower temperature, which sets the amount of the heavy hydrocarbons in the top product. This is controlled by the reflux ratio. Increasing the reflux rate will decrease this amount. The reader should observe that reflux liquid is produced as a result of overhead condensation of vapors. For columns using total condensers, such as depropanizers and debutanizers, all vapors

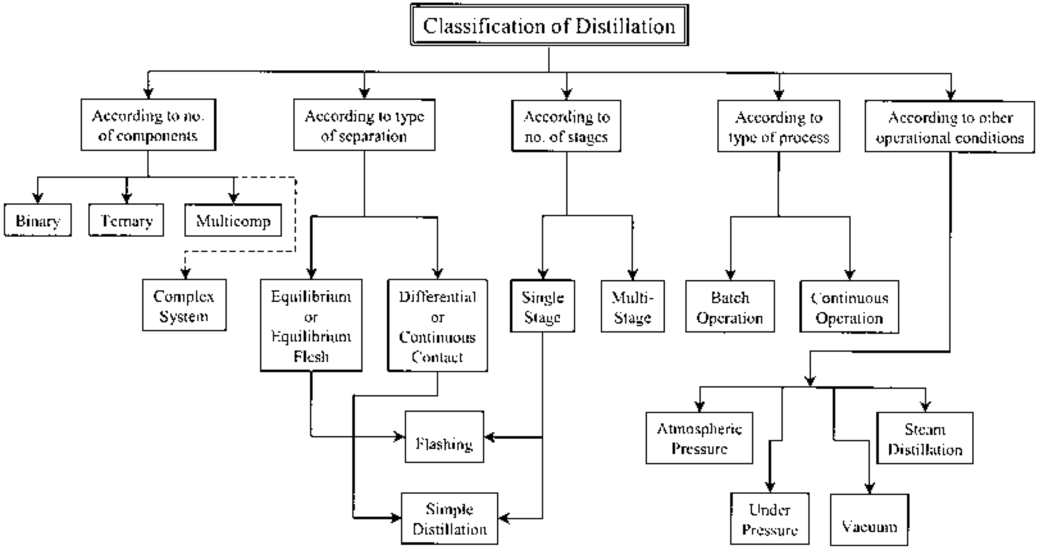


Figure 6 Classification of distillation processes. (From Ref. 1.)

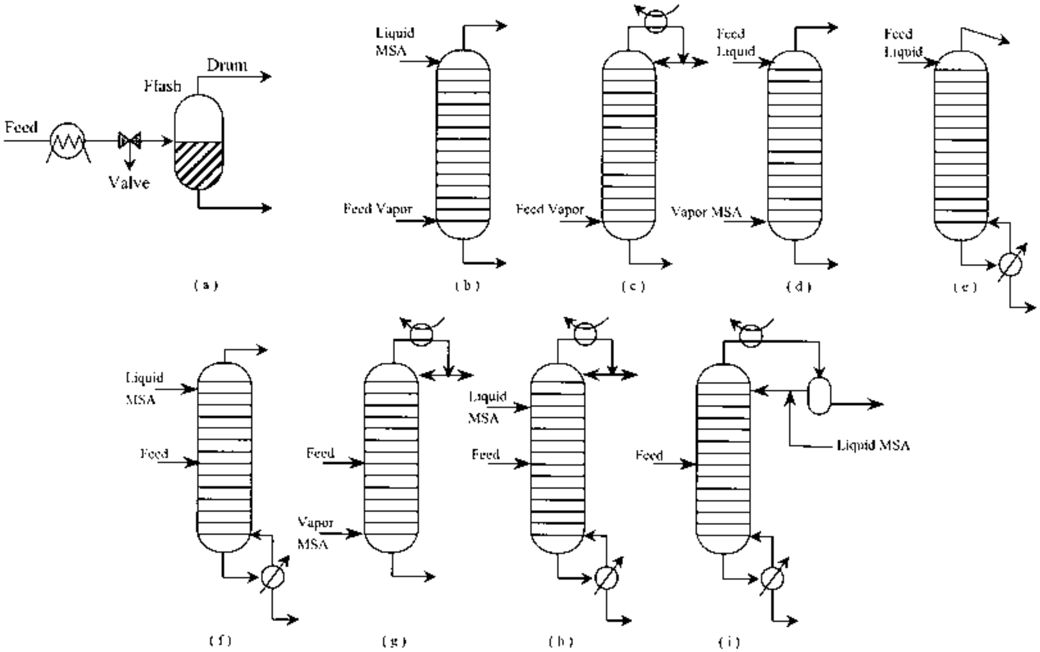


Figure 7 Separation operations related to distillation. (a) Flash vaporization or condensation; (b) Absorption; (c) Rectifier; (d) Stripping; (e) Reboiled stripping; (f) Reboiled absorption; (g) Refluxed stripping; (h) Extractive distillation; and (i) Azeotropic distillation. (From Ref. 3.)

are condensed to produce reflux and liquid product. On the other hand, for columns employing partial condensers, such as deethanizers, product is produced as vapor (ethane).

2. Bottom reboiler temperature, which sets the amount of light hydrocarbons in the bottom product. Adjusting the heat input to the reboiler controls this.
3. Tower operating pressure, which is fixed by the type of condensing medium (i.e., its temperature). Product quality is not affected, to a great extent, by changing the operating pressure.

13.4 SYNTHESIS OF SEPARATION (DISTILLATION) SEQUENCES

In the separation and the recovery of NGL constituents from a gas stream, it is required to find the optimal arrangement of the separation steps that will be both economically and technically feasible.

In this regard, solutions are reached through some rules or heuristics as explained next [4].

13.4.1 Cost of Separation (Fractionation)

The cost of fractionation is influenced by the following:

1. The quantity or feed input to be processed, load (L)
2. The boiling point difference between two components (Z), defined also as the property difference between the two species on each side of the separation breaking point

One can simply relate the cost of separation to these two variables by the following relationship: cost of separation is a function of L/Z, or

$$\text{Cost of separation} = k\left(\frac{L}{Z}\right)$$

Because Z is fixed for a given mixture, our main objective is to select an optimal arrangement of separation sequence that will minimize L, thus, the cost of separation.

13.4.2 Rules (Heuristics) and Examples

In this subsection, some examples are presented along with the relevant rules that apply. Some important formulas are presented first:

$$\text{Number of separating columns} = \text{Number of components} - 1 \tag{1}$$

The number of sequences S is related to the number of components c by the relationship

$$S = \frac{[2(c-1)]!}{c!(c-1)!} \tag{2}$$

Example 1 (Introductory Example)

Find the possible technical sequences for separating a mixture of BTX (benzene, toluene, xylene).

Solution

First, applying Eq. (1), the number of columns to be used is two. Similarly, solving Eq. (2),

$$\begin{aligned} S &= \frac{[2(2)]!}{(3 \times 2 \times 1)(2 \times 1)} \\ &= \frac{4 \times 3 \times 2 \times 1}{3 \times 2 \times 2} = 2 \end{aligned}$$

The number of feasible sequences is found to be 2. However, in Figure 8, four sequences are shown; but not all of them are feasible. Both arrangements (b) and (c) are excluded because the former does not produce pure products and the latter employs three columns. Thus, arrangements (a) and (b) are both feasible and equivalent.

Statement of Rules

Rule 1: All other things being equal, aim to separate the more plentiful components early.

Rule 2: Difficult separations are best saved last.

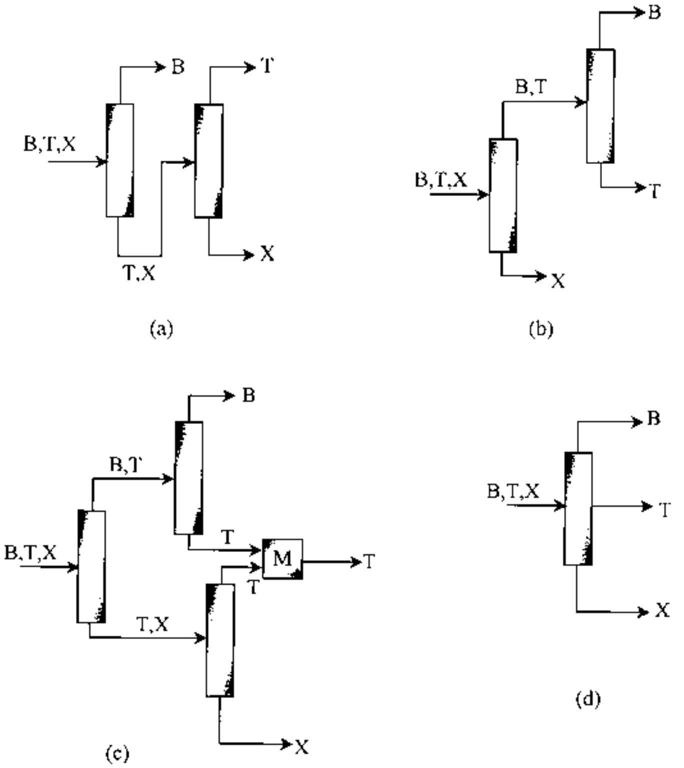


Figure 8 Suggested sequences for separating B.T.X. mixture (Example 1).

Rule 3: When using distillation or similar schemes, choose a sequence that will finally remove the most valuable species or desired product as a distillate, all other things being equal.

Rule 4: During distillation, sequences that remove the components one by one in column overheads should be favored, all other things being equal.

Rule 5: All other things being equal, avoid excursions in temperature and pressure, but aim high rather than low.

Example 2

Suppose we want to separate a mixture of four components: components 1–4 into pure products. Assume that they are present in the feed in equal amounts: D_1, D_2, D_3, and D_4.

Solution

Let us calculate the value of S by applying formula (2). S is found to be 5; the five schemes are shown in Figure 9. Now, in order to select the best scheme, the total load of all sequences have to be calculated:

$$\begin{aligned}\text{Total load of a sequence} &= \text{sum of individual loads of all} \\ &\quad \text{components undergoing separation} \\ &= \text{sum of } L_i\end{aligned}$$

Because $L_i = D_i n$, where n is defined as the number of columns or separators component i, goes through before final separation,

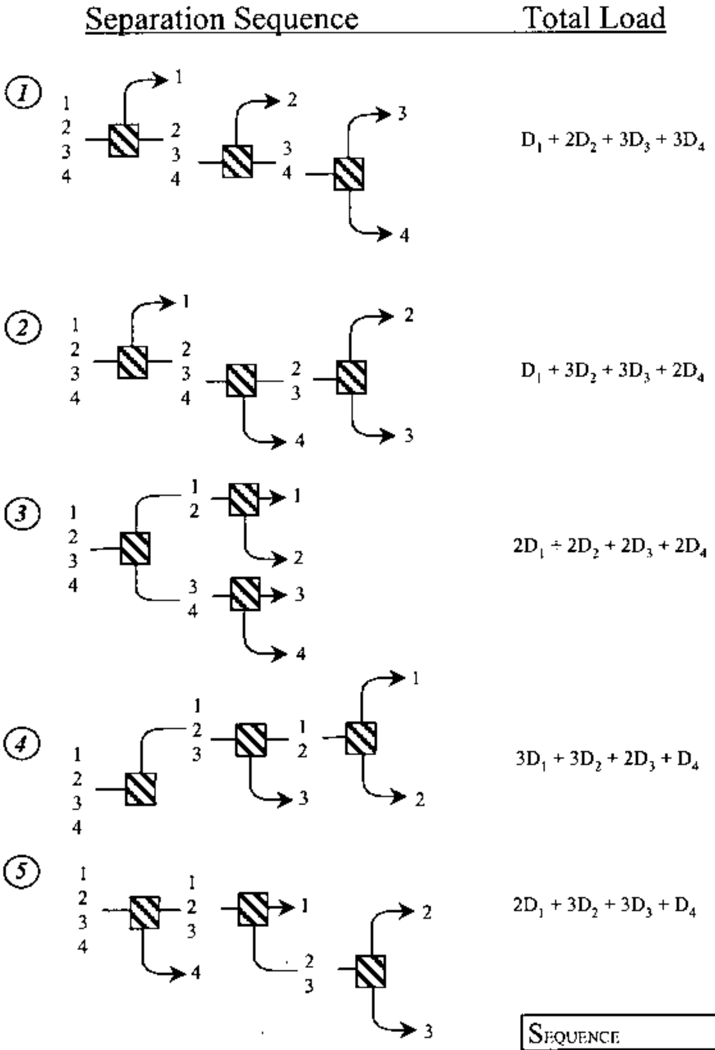
$$\text{Total load of a sequence} = \text{Sum of } D_i n \tag{3}$$

Calculations are carried out as shown in Figure 9. It can be concluded that sequence 3 has less load (8D).

Example 3

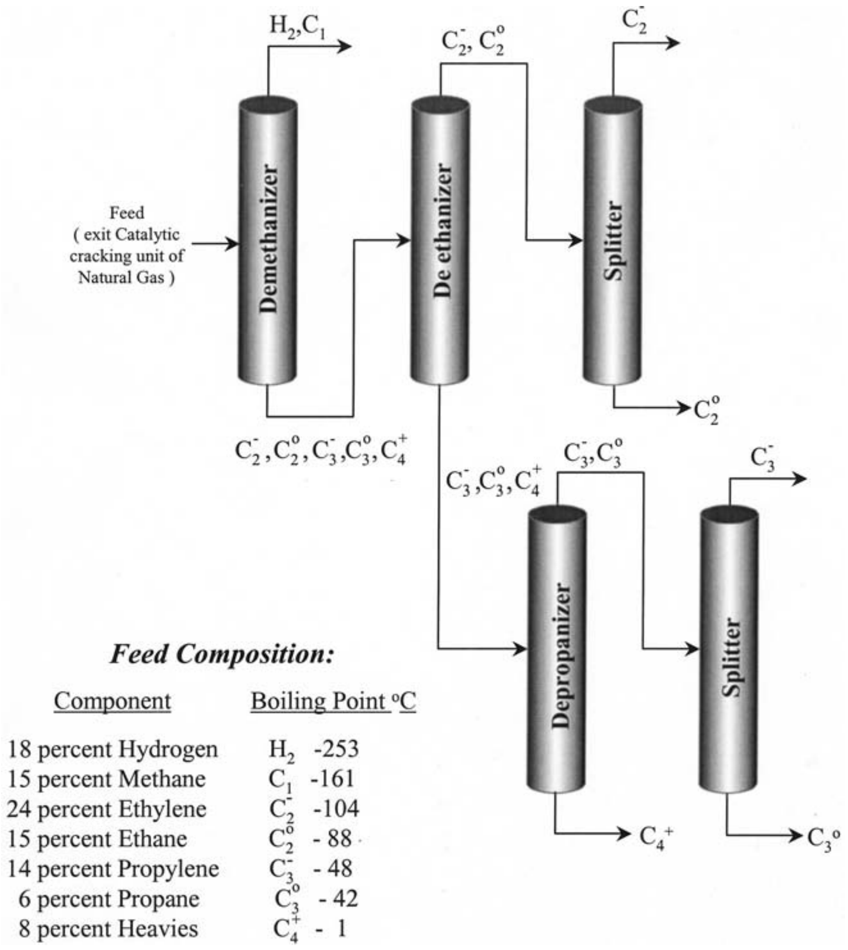
This example is a direct industrial application for some of the rules or heuristics presented in the course of separation of NGL constituents. It deals with ethylene and propylene manufacture. Figure 10 describes a sequence of distillation separation processes for a gas mixture produced by a catalytic cracking plant of natural gas. The plant built by Sinclair produces 500 million lbs/year of ethylene. Ethylene and propylene are the valuable products formed by catalytic cracking of the hydrocarbons in the natural gas. The following observations are cited in accordance with the heuristics stated:

1. *Difficult separation is last*: Because of the close boiling points of propane and propylene, separation between them is kept the very last (splitter). The next most difficult separation is between ethane and ethylene. Again, it is kept to a last position (splitter).
2. *Favor overhead removal in plentiful quantities*: In the first distillation column, the demethanizer separates the volatile components hydrogen and methane (18% and 15%, respectively).
3. *Remove valuable products as distillates*: Both ethylene and propylene are separated as top products. This ensures that the materials do not experience discoloration and separate in pure form.



SEQUENCE	TOTAL LOAD
1	9D
2	9D
3	8D
4	9D
5	9D

Figure 9 Solution of Example 2.



Feed Composition:

Component	Boiling Point °C	
18 percent Hydrogen	H_2	-253
15 percent Methane	C_1	-161
24 percent Ethylene	C_2^-	-104
15 percent Ethane	C_2^o	- 88
14 percent Propylene	C_3^-	- 48
6 percent Propane	C_3^o	- 42
8 percent Heavies	C_4^+	- 1

Figure 10 Sequences of distillation columns to separate products in the manufacture of ethylene and propylene.

13.5 NGL PRODUCTS AND THEIR STABILIZATION

Listed are some commercial finished products that are obtained by fractionation of NGL stream and sold in a gas plant:

1. Ethane product
2. Propane product
3. Butanes product
4. Pentanes and heavier (natural gasoline) product.

Field stabilization is required for liquid products recovered as explained before, in order to remove the large quantities of methane and other volatile hydrocarbons that remain locked in the condensed liquid at high pressure.

Stabilization methods are described in Chapter 7.

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REVIEW QUESTIONS AND EXERCISE PROBLEMS

1. To recover and separate NGL from a bulk of gas stream, a change in phase must occur. What are the two main options that bring in a change in phase that lead to separating NGL?
2. A change in phase for NGL recovery, thus its separation from the gas stream always involves the control of one or more of the following parameters:

   (a) operating pressure
   (b) acid gas content
   (c) operating temperature
   (d) water vapor content
   (e) system composition of hydrocarbons.

3. For separation using energy separating agents (ESA),______is maintained by direct control.______, on the other hand, is reduced by refrigeration using one of the three techniques: ______, ______, or ______.
4. For separation using mass separating agent (MSA), a control in the ______ of the hydrocarbons to be recovered as NGL is obtained by using ______ or ______.
5. In the absorption process for the recovery of NGL, the up-flow gas stream is brought in direct contact counter-current with the solvent which is ______. The rich oil is made up of ______ plus ______.

6. Refrigeration methods for NGL recovery imply using ______ to chill the feed (natural gas) to a temperature between 0 and −40°F.
7. Name the two well-known approaches of cryogenic processes that lead to the separation of NGL; that is "auto-refrigeration."