

SECTION 16

Hydrocarbon Recovery

Gas processing covers a broad range of operations to prepare natural gas for market. Processes for removal of contaminants such as H_2S , CO_2 and water are covered extensively in other sections of the Data Book. This chapter will cover the processes involved in recovering light hydrocarbon liquids either for sale when their value as liquids is higher than their value as gas components or they must be removed to avoid condensation. The equipment components included in the processes described are covered in other sections of the Data Book. This section will bring those components together in process configurations used for liquid production.

INTRODUCTION

The recovery of light hydrocarbon liquids from natural gas streams can range from simple dew point control to avoid liquid formation to deep ethane extraction. The desired degree of liquid recovery has a profound effect on process selection, complexity, and cost of the processing facility.

The term NGL (natural gas liquids) is a general term which applies to liquids recovered from natural gas and as such refers to ethane and heavier products. The term LPG (liquefied petroleum gas) describes hydrocarbon mixtures in which the main components are propane, iso and normal butane, propylene and butylenes. Typically in natural gas production olefins are not present in LPG.

Typically, modern gas processing facilities produce a single ethane plus product (normally called Y-grade) which is often sent offsite for fractionation and processing. Whether accomplished on-site or at another facility, the mixed product will typically be fractionated to make products such as purity ethane, ethane-propane (EP), commercial propane, isobutane, normal butane, mixed butanes, butane-gasoline (BG), and gasoline (or stabilized condensate). The degree of fractionation and the liquid products is market and geographically dependent.

Early efforts in the 20th century for liquid recovery involved compression and cooling of the gas stream and stabilization of a gasoline product. The lean oil absorption process was developed in the 1920s to increase recovery of gasoline and produce products with increasing quantities of butane. These gasoline products were, and still are, sold on a Reid vapor pressure (RVP) specification. Vapor pressures such as 10 or 12 psia are common specifications for gasoline products. To further increase production of liquids, refrigerated lean oil absorption was developed in the 1950s. By cooling the oil and the gas with refrigeration, the absorber vapor outlet is leaner and propane product can be recovered. With the production of propane from lean oil plants, a market developed for LPG as a portable liquid fuel.

In lieu of using lean oil, refrigeration of the gas can be used for propane and heavier component recovery. The use of straight refrigeration typically results in a much more economical processing facility than using lean oil. The chilling of the gas can be accomplished with mechanical refrigeration, absorption refrigeration, expansion through a J-T valve, or a combination. In order to achieve still lower processing temperatures, cascade

refrigeration, mixed refrigerants, and most significantly turbo-expander technologies have been developed and applied. With these technologies, recoveries of liquids can be significantly increased to achieve deep ethane recoveries. Early ethane recovery facilities targeted about 50 % ethane recovery. As processes developed, ethane recovery efficiencies have increased to well over 90% in well integrated facilities.

In some instances heavy hydrocarbons are removed to control the hydrocarbon dew point of the gas and prevent liquid from condensing in pipeline transmission and fuel systems. In this case the liquids are a byproduct of the processing and if no market exists for the liquids, they may be used as fuel. Alternatively, the liquids may be stabilized and marketed as condensate.

GAS COMPOSITION

The gas composition has a major impact on the economics of NGL recovery and the process selection. In general, gas with a greater quantity of liquefiable hydrocarbons produces a greater quantity of products and hence greater revenues for the gas processing facility. Richer gas also entails larger refrigeration duties, larger heat exchange surfaces and higher capital cost for a given recovery efficiency. Leaner gases generally require more severe processing conditions (lower temperatures) to achieve high recovery efficiencies and incur a higher cost per unit of liquid product.

Gases are typically characterized by the gallons per thousand cubic feet of recoverable hydrocarbons in the gas. This is commonly expressed as "GPM." GPM was traditionally meant to apply to propane and heavier components but is often used to include ethane. The GPM of a gas can be calculated as shown in Example 16-1.

The other major consideration in the evaluation of NGL recovery options is the specification of the residue sales gas. Sales specifications are usually concerned with a minimum Higher Heating Value (HHV) of the gas, but in some instances the maximum HHV can also be a consideration. The calculation of HHV is covered in Section 23 and in more detail in GPA Standard 2172, "Calculation of Gross Heating Value, Relative Density, and Compressibility Factor for Natural Gas Mixtures from Compositional Analysis." In addition, for some gas sales the maximum and minimum Wobbe Number of the gas may be specified. For more information on the calculation of Wobbe Number, See Section 1 definitions.

Removal of liquids results in gas "shrinkage" and reduction of the HHV. This shrinkage represents a loss of revenue for the gas sales which must be considered in the economics of an NGL recovery plant. In general, sales gas specifications set the minimum HHV at 950-1000 BTU/scf. Thus, if any components such as nitrogen or CO_2 are present in the gas, sufficient ethane and heavier components must remain in the gas to meet the heating value specification. If little nitrogen or CO_2 is present in the gas, the recovery level of the ethane and heavier components is then limited by markets, cost of recovery, and gas value. The

calculation of HHV and shrinkage cost is illustrated in Example 16-1.

Example 16-1 — Find the GPM of the gas mixture in Fig. 16-1. Find the HHV of the feed gas and the HHV of the residue gas with the following NGL recovery efficiencies: C_2 – 90%, C_3 – 98%, iC_4/nC_4 – 99%, C_5+ – 100%. What is the shrinkage cost at \$4/MMBTU?

Solution Steps:

Solution is shown in Fig. 16-1.

From Fig. 23-2 obtain the gal/lb mole for each of the components. Multiply the mole fraction of each component (mol% / 100) by its gal/lb-mol value and divide by 379.49 scf/lb-mol to get gallons per cubic feet of gas. Multiply this value by 1,000 scf/Mscf to get gallons per thousand cubic feet of gas (GPM). The total GPM from Fig. 16-1 is 3.11.

For the recoveries specified the net gal/day and residue composition can be found as shown in Fig. 16-1. In order to compute the HHV of the two streams, the HHVs of each component are found in Fig. 23-2. Multiplying the individual HHVs by the mole % gives a total HHV of 1115.02 for the feed gas and 971.24 for the residue gas.

The shrinkage volume can be found by the difference of the volume of the feed gas times the HHV and the volume of residue gas times its HHV. This volume is then multiplied by \$4/MMBTU to get the shrinkage value of the NGLs.

$$\begin{aligned} \text{Shrinkage Value} &= \$4/\text{MMBTU} \cdot [(330 \cdot 1115.02) - \\ &\quad (295.862 \cdot 971.24)] = \cdot \$4/\text{MMBTU} \cdot 80606 \text{ MMBTUs} = \\ &\quad \$322,424/\text{day} \end{aligned}$$

The value of the NGLs in \$/gal versus the value of the components in the residue gas in \$/gal or the “spread” between these values is the primary economic criteria for NGL recovery project evaluations.

FIG. 16-1
Solution to Example 16-1

GPM CALCULATION							
Component	Feed Gas Mole %	Gal/Mole	Available		Estimated Recovery %	Net Gal/Day	Residue Gas Mole %
			GPM	Gal/Day			
N ₂	1.000						1.115
CO ₂	3.000						3.346
C ₁	85.000						94.808
C ₂	5.800	10.119	1.55	510363	90	459327	0.647
C ₃	3.000	10.424	0.82	271938	98	266499	0.067
IC ₄	0.700	12.384	0.23	75383	99	74629	0.008
NC ₄	0.800	11.936	0.25	83035	99	82205	0.009
IC ₅	0.300	13.855	0.11	36144	100	36144	0.000
NC ₅	0.200	13.712	0.07	23848	100	23848	0.000
C ₆ + (Set as NC ₆ for example)	0.200	15.566	0.08	23072	100	26072	0.000
Total	100.000		3.11	1027783		969724	100.000
MMSCFD	330.000						295.862
SHRINKAGE CALCULATION							
Component	Feed Gas Mole%	Residue Gas Mole %	HHV BTU/scf	Feed Gas BTU/scf	Residue Gas BTU/scf		
N ₂	1.000	1.115	0.0	0.00	0.00		
CO ₂	3.000	3.346	0.0	0.00	0.00		
C ₁	85.000	94.808	1010.0	858.50	957.56		
C ₂	5.800	0.647	1769.7	102.64	11.45		
C ₃	3.000	0.067	2516.2	75.49	1.68		
IC ₄	0.700	0.008	3252.0	22.76	0.25		
NC ₄	0.800	0.009	3262.4	26.10	0.29		
IC ₅	0.300	0.000	4000.9	12.00	0.00		
NC ₅	0.200	0.000	4008.7	8.02	0.00		
C ₆ + (Set as NC ₆ for example)	0.200	0.000	4756.0	9.51	0.00		
Total	100.000	100.000		1115.02	971.24		
MMSCFD	330.000	295.862					

DEW POINT CONTROL

Retrograde condensation has long been known to occur at reservoir conditions. Recognition that it also occurs in typical processing conditions was an early result of computer calculations using equations of state to predict vapor-liquid behavior. The phenomenon is illustrated in Fig. 16-2 showing dew point calculations for a gas stream leaving a separator at 100°F and 1000 psia. These dew point curves show that as the pressure is reduced, liquid is formed. The heavier the hydrocarbon, the more the dew point temperature increases as the pressure is lowered. The cricondentherm of the dew point curve is primarily determined by the presence of the heaviest component in the gas rather than just the total quantity of the heavy component in the feed gas so accurate determination of the heaviest components is important in establishing the phase envelope.

When gas is transported in pipelines, consideration must be given to the control of the formation of hydrocarbon liquids in the pipeline system. Condensation of liquid is a problem in metering, pressure drop and safe operation. Condensation of liquid can also be a major problem due to two-phase flow and liquid slugging.

To prevent the formation of liquids in the system, it is necessary to control the hydrocarbon dew point below the pipeline

operating conditions. Since the pipeline operating conditions are usually fixed by design and environmental considerations, single-phase flow can only be assured by removal of the heavier hydrocarbons from the gas.

Low Temperature Separation (LTS)

Several methods can be used to reduce the hydrocarbon dew point. If sufficient pressure is available, the removal can be accomplished by expansion refrigeration in an LTS unit. The expansion refrigeration system uses the Joule-Thomson effect to reduce the gas temperature upon adiabatic expansion. This temperature reduction results in not only hydrocarbon liquid condensation but also water condensation. The water is generally removed as hydrates in this process, melted and removed. Thus, the process can actually accomplish dew point control of both water and hydrocarbon in a single unit.

Fig. 16-3 shows one example of an LTS system. The high pressure gas may first go through a heater but this is often not needed, depending on the gas conditions. The gas then enters the heat exchanger coil in the bottom of the separator where the gas is cooled by exchange with the condensed liquid and by melting the hydrates formed. Any water or condensate produced at this point is removed in the high pressure separator. The gas from the separator is then heat exchanged with the

FIG. 16-2
Typical Low Pressure Retrograde Condensation Dew Point Curves²

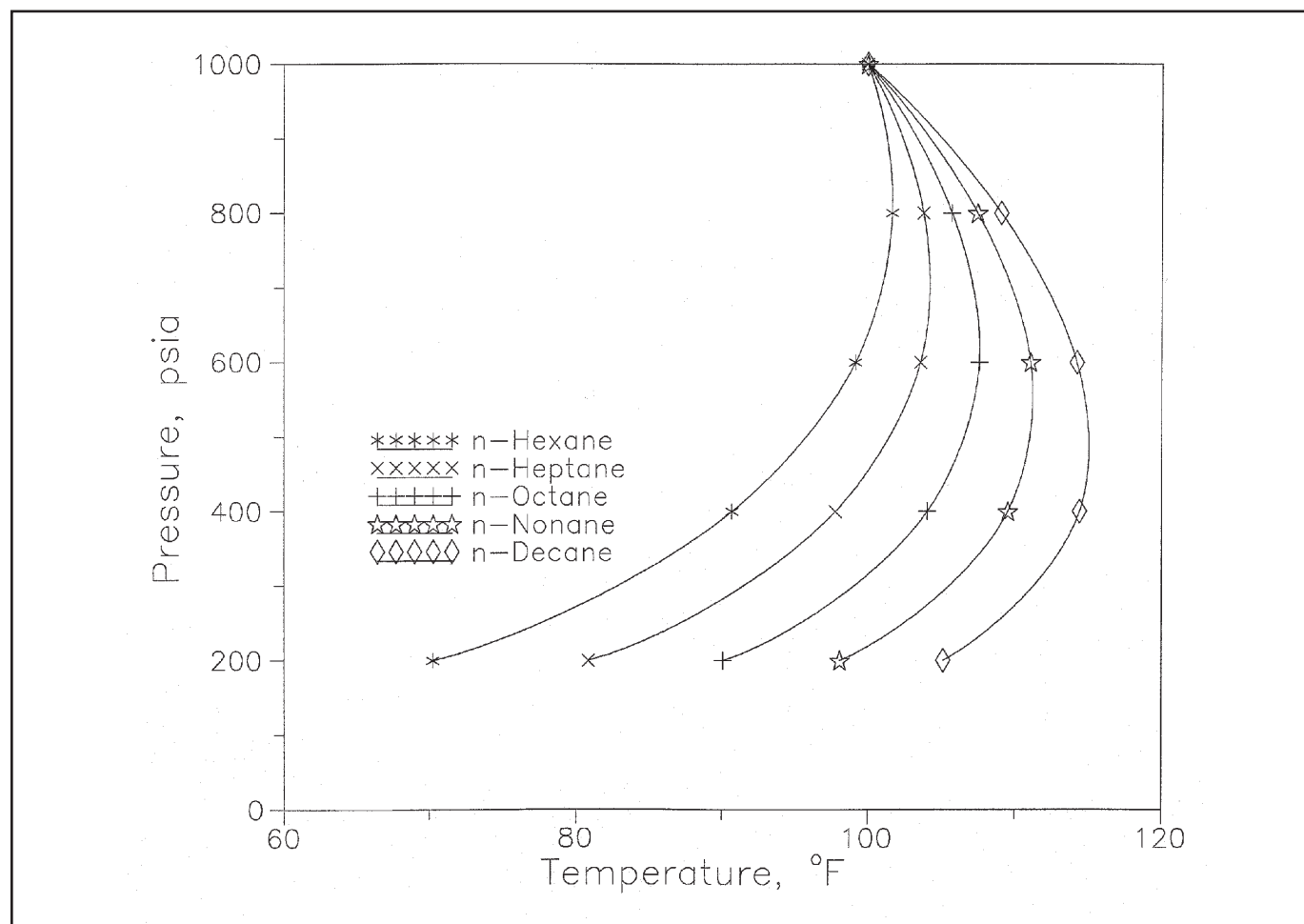


FIG. 16-4
Low-Temperature Separation System with Glycol Injection and Condensate Stabilization³

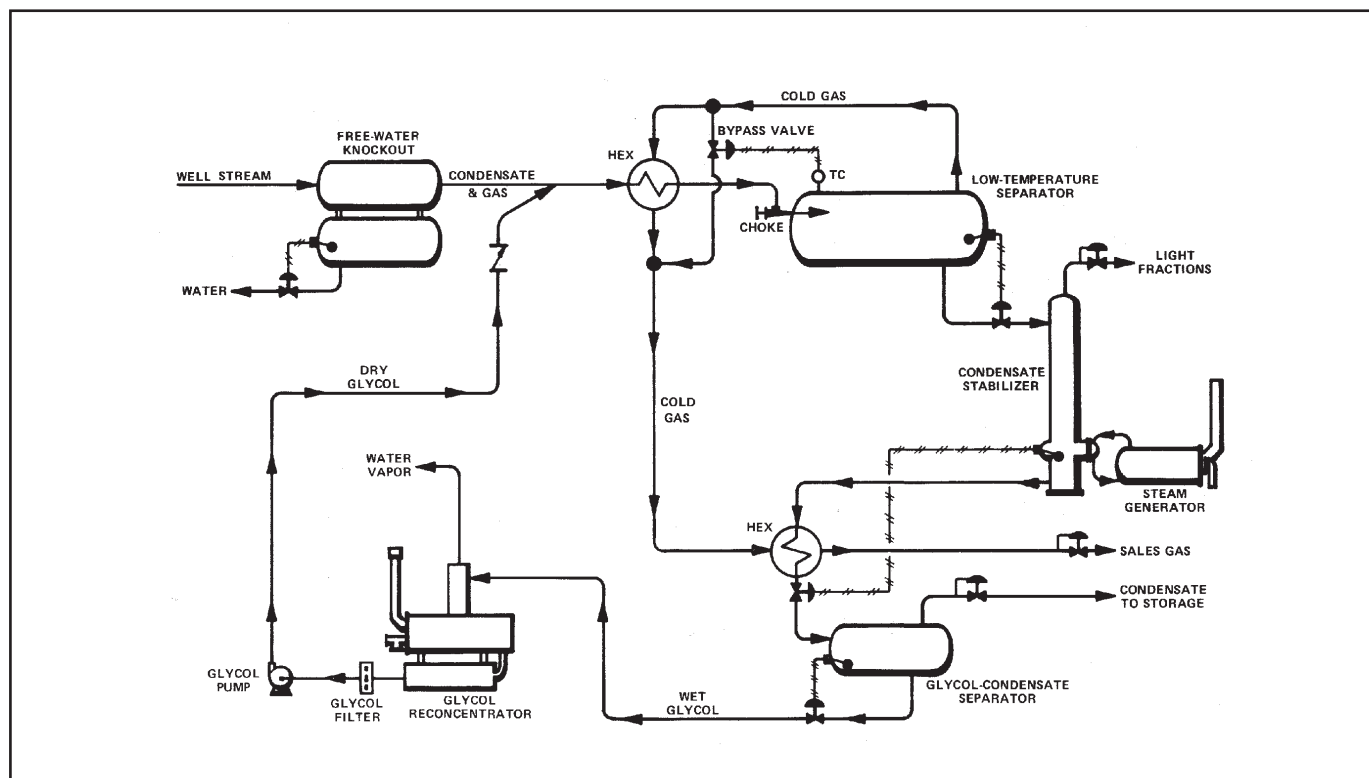
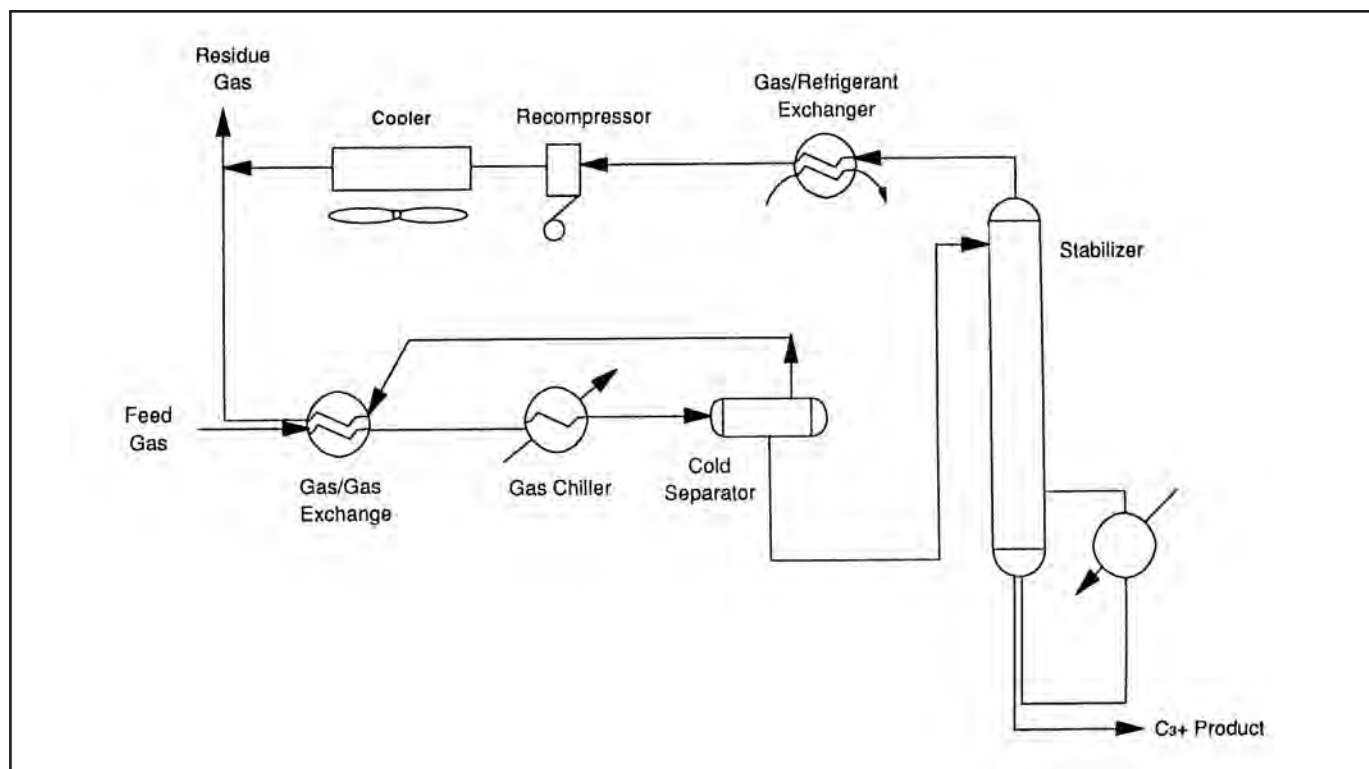


FIG. 16-5
Straight Refrigeration Process



Stabilization

One of the problems in using dew point control units of both expansion LTS and mechanical refrigeration systems is the disposition of the liquids removed. The liquids must be stabilized by flashing to lower pressure at high temperature or by using a stabilization column. When the condensate is flashed to a lower pressure, light hydrocarbons are liberated which can often be used in a fuel gas system.

The stabilization column can produce a higher quality and better controlled product. The condensate stabilizer is usually a top feed column which runs at a reduced pressure from the cold separator and has a reboiler to produce a specified vapor pressure product. The overhead vapor is either sent to fuel as shown in Fig. 16-4 or recompressed and combined with the sales gas as shown in Fig. 16-5. The column contains either trays or packing to provide necessary mass transfer for stabilization of the liquid feed. After stabilization, the product is cooled and sent to storage.

Compact Separation Designs

Compact process configurations have been commercialized to take advantage of gas expansion for liquid separation. Each of these processes use static equipment to achieve the desired separation and are focused on replacing Joule-Thomson expansion valves and/or turbo-expanders.

One of these processes which is currently used in several operations is the Twister[®] technology. This process (Fig. 16-6) uses a supersonic nozzle in which the pressure is reduced via isentropic expansion such that liquid is formed. Saturated feed gas is passed across a vane ring which induces a swirling motion where after the swirling gas is expanded in a Laval-type nozzle causing pressure and temperature to drop while the flow becomes supersonic. The centrifugal motion forces the liquid droplets to agglomerate at the wall from where it is drained from the apparatus. The dry vapor stream is then decelerated

in a diffuser duct to pipeline velocity, causing the pressure to increase to 70–80% of the feed pressure. Ample testing has shown that the Twister process has a high isentropic efficiency while separating almost all of the condensed liquid.

This technology is used for water and hydrocarbon dewpoint control in both onshore and offshore locations, especially where space and weight are at a premium. Since the expansion cooling in Twister is close to isentropic, more valuable NGL liquids are produced compared to a JT LTS process. A Twister based dew pointing process can often operate without glycol or methanol systems for hydrate suppression.

Another process uses a vortex tube device to affect the separation. The vortex tube is based on the Ranque-Hilsch tubes developed in the 1940s. These tubes have been used as laboratory devices and small scale coolers. The working principle of these devices is the same. (Fig. 16-7) A gas is injected tangentially through a nozzle into the center of the tube where it expands to a low pressure. The gas flows cyclonic to the far end of the tube. During this flow, two temperature zones are formed, a warm zone near the wall and a cooler zone near the center. At the end of the tube the center gas is deflected and returns along the tube through an orifice near the inlet nozzle. The tube is therefore capable of producing two gas streams at different temperatures. The cold gas is at a temperature below that achievable with an isenthalpic expansion. If the two outlet streams were to be mixed, the combined temperature would be equal to the temperature achieved by the isenthalpic expansion. Thus the vortex tube performs the same function as the Joule-Thomson valve but produces a lower outlet gas temperature for a portion of the stream. This apparatus could have application where gas pressure drop is available, dewpoint control is needed, and the warm and cool gases are recombined after liquid removal. This technology is very space efficient and can be attractive, especially for offshore processing applications.

FIG. 16-6
Concept of the Twister Process

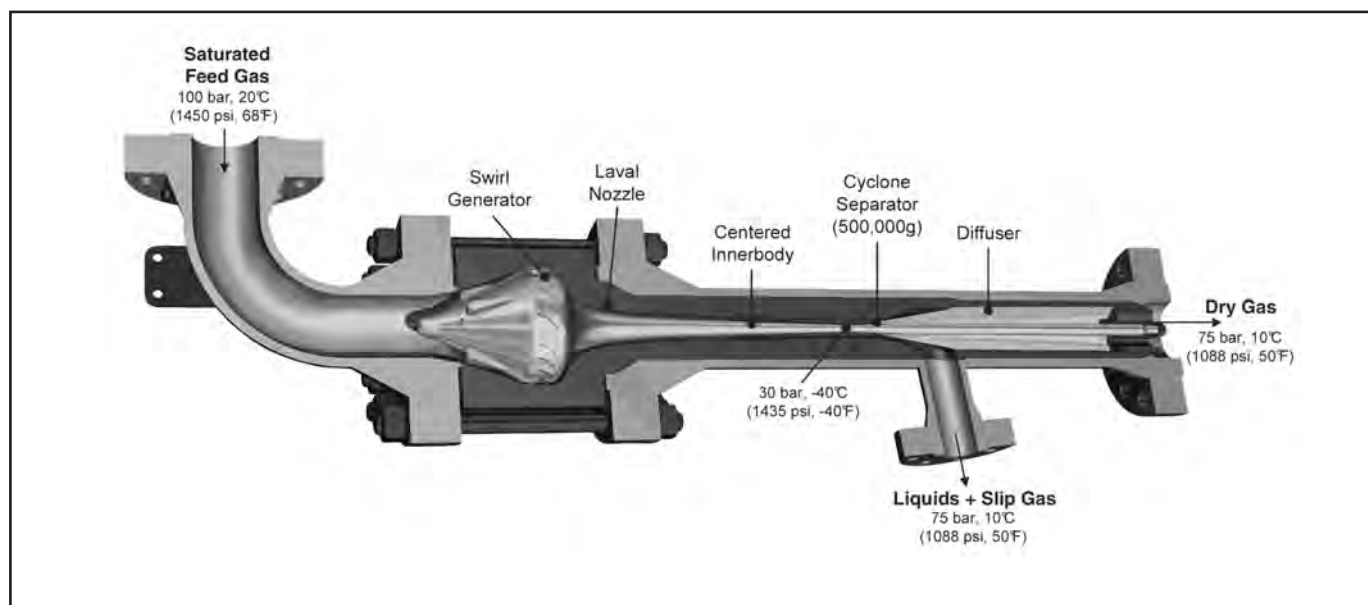
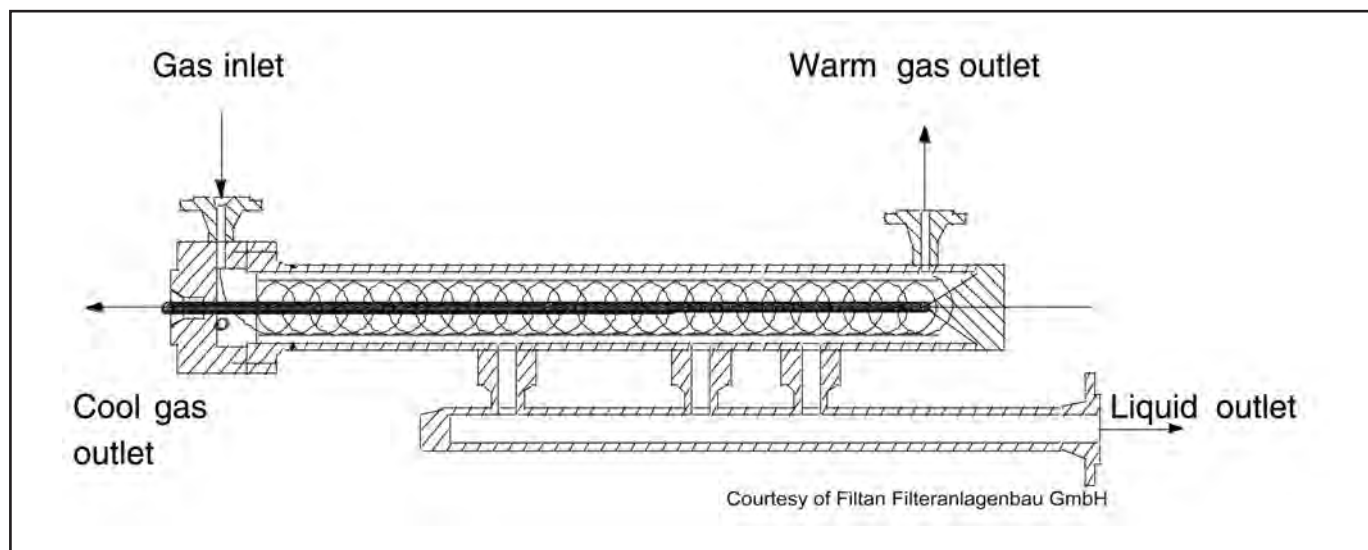


FIG. 16-7
Basic Design of a Vortex-Tube Device



Membrane Conditioning

Rubbery heavy hydrocarbon selective membranes are another recent process development for use in fuel as conditioning. These membranes are different from conventional cellulosic membranes used for CO₂ removal from natural gas as described in Section 21. These rubbery membranes preferentially permeate the heavy hydrocarbons through the membrane over methane, which is counterintuitive. The reason for this “reverse permeation” behavior is that the rubbery membranes separate gases based on solubility differences as compared to diffusion rate differences in the molecules.

This separation difference allows these membranes to be used for a variety of applications in the natural gas processing and treating areas. The main application of these rubbery membranes has been in remote area fuel gas conditioning to decrease BTU value of raw fuel gas. This process allows the raw gas to meet required engine specifications, thereby allowing operators to run compression driven by gas engines and move the gas from the remote location to the central processing facilities. One version of the membrane will also remove CO₂/H₂S as well as the heavy hydrocarbons. One company, MTR has dozens of units installed in various offshore, onshore remote locations, shale gas production areas and early production facilities all over the world⁴ (See Fig. 16-8).

Other commercial applications of the rubbery membranes are in nitrogen rejection (to reduce N₂ content in natural gas), and for LPG/NGL recovery from gas streams. In the nitrogen rejection application, N₂ is the least permeable component in the rubbery membrane, so the membrane can be used to reduce the N₂ concentration in the pipeline gas, but in this case the methane rich gas will be in the low pressure permeate along with the heavier hydrocarbons. Therefore, the pipeline gas in this case would have a high BTU recovery, while rejecting the nitrogen in the residue gas.

In LPG recovery, the fact that the heavy hydrocarbons permeate the membrane, can be effectively used to increase the NGL content in the condensing loop, thereby allowing NGL recovery at higher condensing temperature. NGL recovery mem-

brane processes are most suitable for flare gas applications in which NGL recovery is combined with power generation from the clean residue gas produced by the rubbery membrane.

FIG. 16-8
Commercial Conditions for Rubbery Membranes

Type of Rubbery Membrane	Pressure Max & (Range) Psig	Flow rate Max & (Range) MMscfd
Standard Hydrocarbon selective (SR Membranes)	1200 psig (50 – 1200 psig)	120 (0.1 – 120)
Polar Hydrocarbon selective membranes (Polaris Membranes)	1000 psig (50 – 1000 psig)	10 (0.2 – 10 MMSCFD)

Rotary Valve Fast Cycle PSA for Fuel Gas Conditioning

A fast cycle Pressure Swing Adsorption (PSA) developed by Xebec has been used successfully to condition gas.⁵ This process selectively adsorbs different gases by adsorbents when subjected to varying pressures. Inlet gas flows up through a vessel filled with a set of adsorbents at relatively high pressure. The heavier hydrocarbons (C₂+), inerts (CO₂ & N₂) and water are preferentially retained, while methane is not. Before the bed becomes saturated, it is switched out of adsorbing mode, and the pressure is reduced. The lower pressure releases the impurities from the adsorbent as exhaust gases. The adsorbent bed is then re-pressurized and is ready to repeat the cycle.

To ensure a continuous supply of product gas, the process uses multiple adsorbent beds, which are cycled in a sequential manner through identical operating cycles. The system contains slowly rotating, multi-port, selector valves, which are central to the operation of the unit. These valves are used to create a very efficient, rapid PSA cycle, which results in smaller equipment packages with reduced capital and installation costs than a traditional PSA would require.

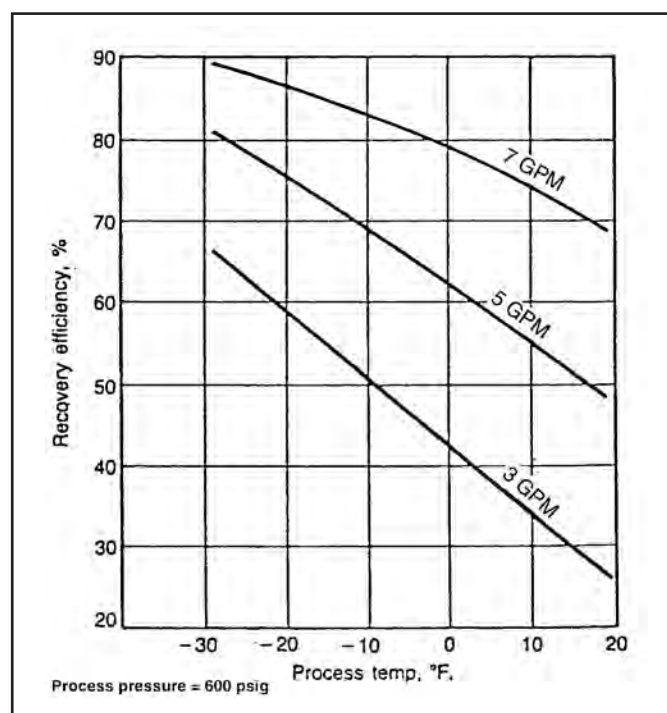
STRAIGHT REFRIGERATION NGL RECOVERY

The straight refrigeration process is quite flexible in its application to NGL recovery. As outlined in the previous section, the process can simply be used for dew point control when modest liquid recovery is needed or desired. Alternatively, the process can be used for high propane recovery and, in the case of rich gases, for reasonable quantities of ethane recovery. The recovery level is a strong function of the feed gas pressure, gas composition and temperature level in the refrigeration chiller. Fig. 16-9 shows curves for estimating the recovery achievable as a function of temperature and gas richness for a given processing pressure. (GPM in this figure is propane plus.) Generally speaking, higher recovery efficiencies can be achieved with richer feed gas. The straight refrigeration process is typically used with a glycol injection system. This configuration is limited in the temperature of operation by the viscosity of the glycol at the lower temperatures. Also, refrigeration is typically provided by propane refrigeration which is limited to -44°F refrigerant at atmospheric pressure and thus a processing temperature of about -40°F . In order to go lower in processing temperature, upstream dehydration and alternative refrigeration systems must be considered.

Fig. 16-10 illustrates the ethane recovery efficiency which can be expected. As with propane recovery, for a given temperature level, higher extraction efficiency can be achieved with richer gas. However, ethane recovery of over 30% can be achieved from a gas as lean as 3 GPM (C_3^+). Fig. 16-11 illustrates the effect of gas pressure on plant performance in propane plus recovery operation.

FIG. 16-9

Propane Plus Recovery Efficiency⁵



Refrigeration Process Alternatives

There are many variations in the straight refrigeration process. Fig. 16-12 illustrates four of the most common variations. In the first scheme the gas is cooled against the residue gas and

FIG. 16-10

Ethane Plus Recovery Efficiency⁶

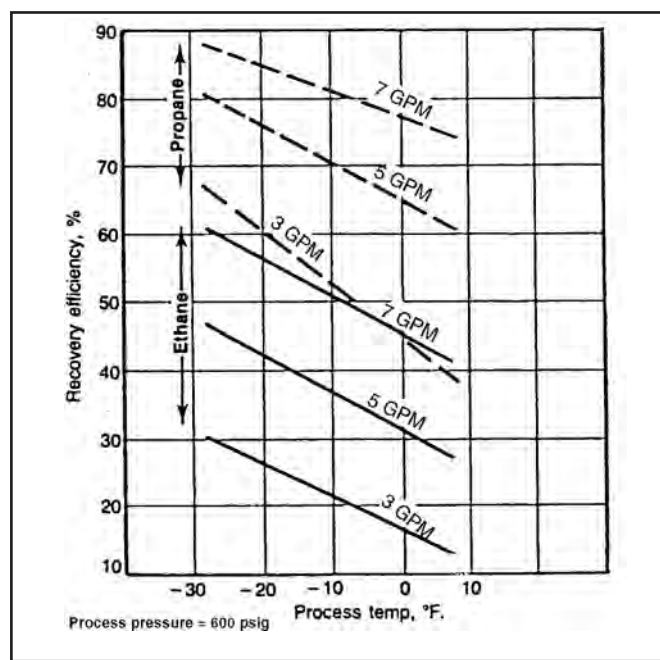


FIG. 16-11

Effect of Gas Conditions on Propane Recovery⁶

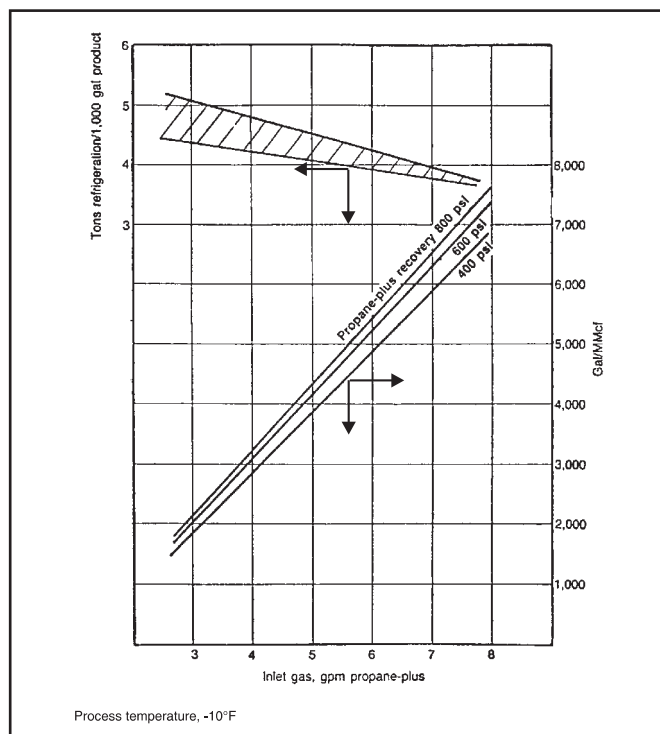
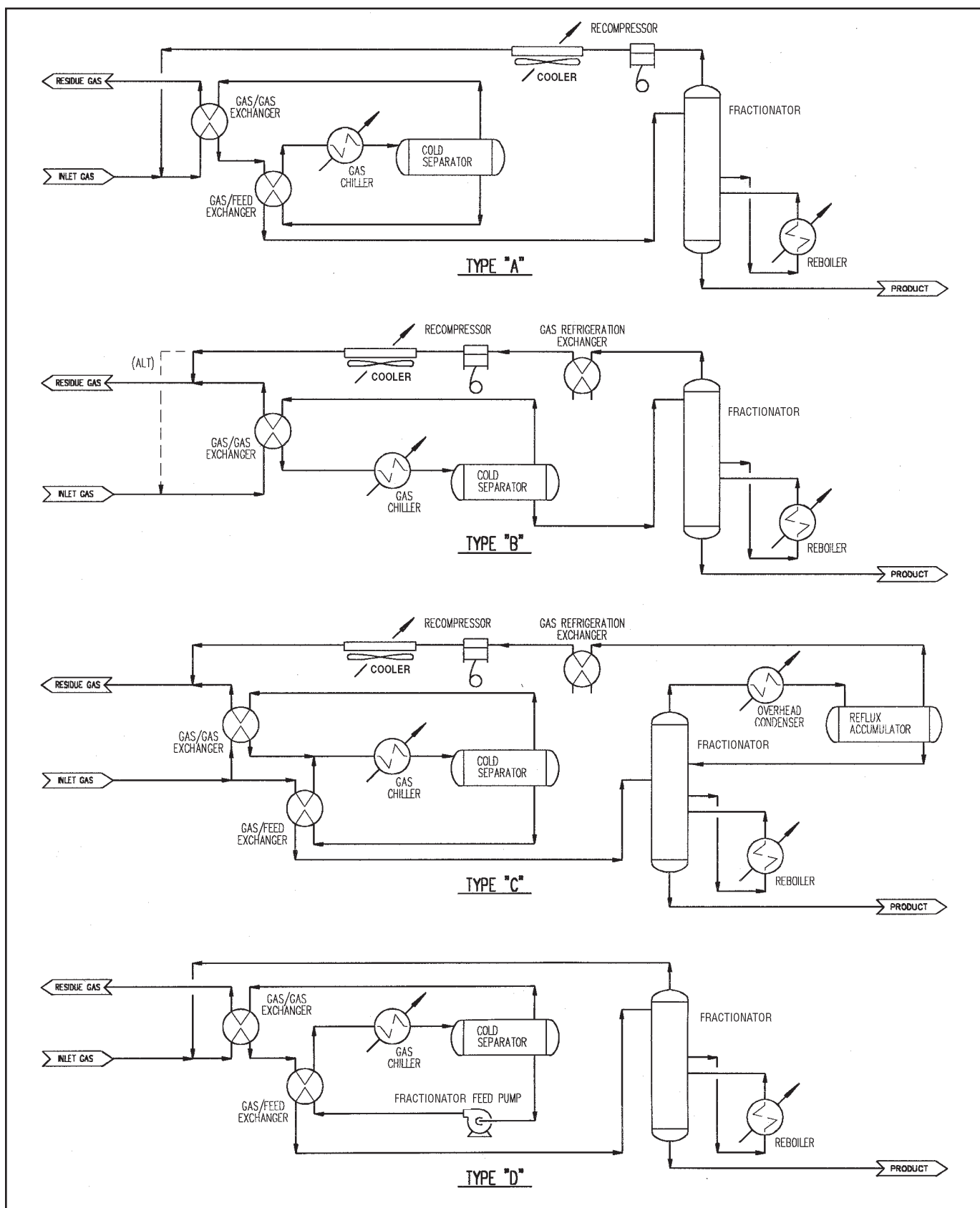


FIG. 16-12
Refrigeration Process Alternatives⁷



the cold separator liquid before being chilled with refrigeration. This scheme uses a top-feed fractionator with the overhead being recompressed and recycled to the inlet. The use of the liquid /feed gas exchanger helps reduce the chiller load. In this case, the residue gas from the cold separator has a dew point of the cold separator operating temperature.

The second scheme also uses a top-feed fractionator, but the cold separator liquid is fed directly to the fractionator. This fractionator operates with a lower overhead temperature which justifies exchange with the refrigeration system. The overhead after being warmed is recompressed and blended with the residue gas from the cold separator. In this configuration the fractionator overhead usually raises the residue gas dew point somewhat. The cold separator temperature must be set to ensure that the desired dew point specification of the combined stream is achieved.

The third process uses a refluxed fractionator. This type of design usually has the highest liquid recovery efficiency, but has a higher cost due to the overhead reflux system. The fourth variation can be used where the cold separator liquid can be pumped and the stabilizer run at an elevated pressure. This can eliminate the need for a recompressor.

Any one or a combination of the following conditions:

- Higher separator pressure
- Richer gas
- Recovery limited to propane-plus

will lead to higher recycle/recompressor rates. This results in more refrigeration horsepower, more recompressor horsepower,

more fractionator heat, and larger equipment. These conditions favor the second and third schemes of Fig. 16-12.

Any one or a combination of the following conditions:

- Lower separator pressure (around 600 psig)
- Leaner gas (below 3 GPM C₃+
- Recovery includes ethane

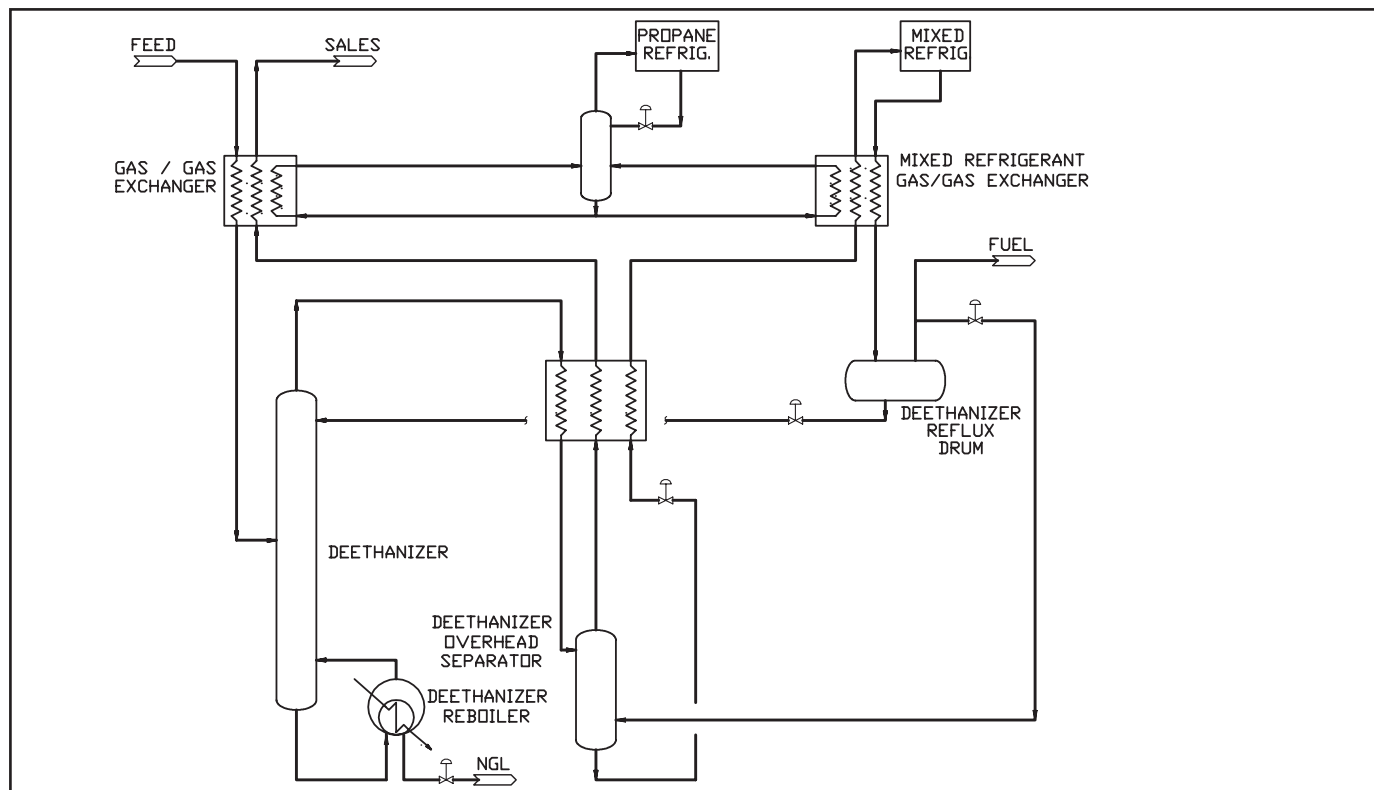
will lead to lower recycle/recompressor rates. These conditions favor the first scheme in Fig. 16-12, or the fourth scheme if the separator pressure is not higher than 400-450 psig. Separator pressure below 400 psig, especially with lean gas, will result in poor product recovery.

Regardless of the exact configuration employed, the capacity of the specific refrigeration system varies directly with refrigerant condensing temperature and evaporating temperature. Condensing temperature is set by the condensing medium available at the plant site, and the process chiller temperature is set by the refrigerant evaporating temperature. Refrigerant horsepower requirements vary with condensing and evaporating temperatures. Lower condenser temperature and higher evaporating temperature require lower horsepower per unit of refrigeration required. For a given refrigeration load, horsepower and condenser duties can be found in Section 14 for a variety of refrigerants.

IPORSM Process

The Lummus Technology/Randall Gas Technologies licenses the IPORSM process, a refrigeration based process, recovers 99.8+% C₃ from gas streams with essentially no C₂ recovery.

FIG. 16-13
IPORSM Process



This design incorporates a closed loop propane refrigeration cycle coupled with an open loop ethane mixed refrigeration cycle. This enables lower temperature operation and provides a reflux stream to the fractionation column, the combination of which produces high recovery for relatively low power consumption.

The process, shown in Fig. 16-13, can operate over a wide range of pressures, depending on the feed conditions and required recoveries, and is well suited to handling rich feed gas. The optimum feed pressure is 250 to 550 psig. The process has an outlet pressure about 25 psi lower than the inlet pressure, so recompression will be required to meet higher residue gas pressure battery limits.

The use of ethane-based refrigeration has been demonstrated to provide efficient high liquids recovery, particularly when combined with simultaneous heat and mass transfer equipment.

LEAN OIL ABSORPTION NGL RECOVERY

Absorption is the physical process where a vapor molecule of a lighter hydrocarbon component will go into solution with a heavier hydrocarbon liquid (nonane, decane and heavier) and be separated from the gas stream. The process can be operated at ambient temperatures if only the heavier NGL products are desired. A refrigerated system enhances the recovery of lighter hydrocarbon products such as ethane and propane; however, ethane recovery is limited to less than 50% to avoid having to remove excessive amounts of methane in the Rich Oil Demethanizer, which would lead to excessive solvent circulation. The absorbing fluid (lean oil) is usually a mixture of paraffinic compounds having a molecular weight between 100 and 200 assuming limited ethane recovery.

Lean oil absorption processes have the advantage that the absorber can operate at essentially feed gas pressure with minimal loss of pressure in the gas stream which exits the process. Plants, whether ambient or refrigerated, are constructed of carbon steel. This type of process was used from the early part of the 20th century and plants are still in use today. However,

most lean oil plants have been shut down or replaced with more modern straight refrigeration or turboexpander process plants. The lean oil process requires large processing equipment with excessive energy requirements. Lean oil absorption units are still used in many refinery operations.

Process Considerations

The desired composition of the lean oil is determined by the absorber pressure and temperature. The optimum molecular weight lean oil is the lowest weight oil which can be retained in the absorber with acceptable equilibrium losses to the residue gas. Lean oil absorption plants operating without refrigeration will require a higher molecular weight oil, usually in the 150-200 molecular weight range. Refrigerated lean oil absorption systems can operate with an absorbing medium as low as 100 molecular weight with proper design.

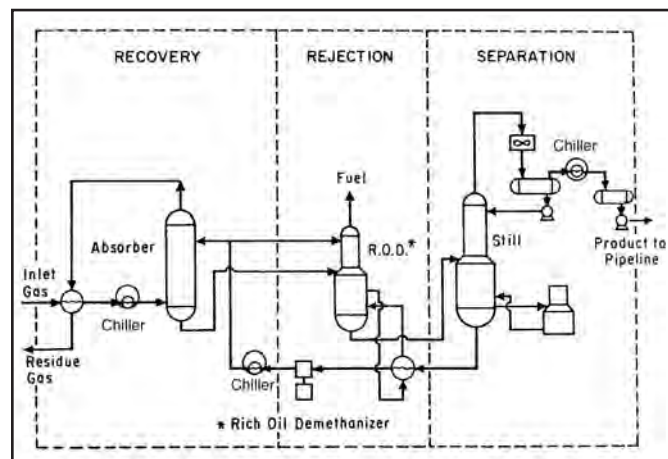
Since the absorption is on a molar basis, it is desired to contact the gas stream with the maximum number of moles of lean oil to maximize the recovery of products from the gas. However, the circulation rate is in units of volume, e.g., gallons per minute. Therefore, a plant designed to circulate a heavier molecular weight oil can circulate more moles of oil with the same equipment if the molecular weight is lowered.

Many absorption oil recovery plants designed to originally operate at ambient temperatures have been modified to include a refrigeration system that allows both the lean oil and the gas to be chilled before entering the absorber. The reduced temperature increases the absorption and allows circulation of less oil of lower molecular weight because the vaporization rate into the residue gas is reduced. Oil is also lost with the NGL product. Oil losses with the product can be minimized by improving fractionation in the lean oil still. Many refrigerated lean oil absorption plants can recover enough heavy ends from the gas stream to offset oil losses from the absorber, thereby making its own absorption oil.

If the gas stream contains compounds that cause the absorption oil molecular weight to exceed design, a lean oil stripper can be used on a side stream of circulating lean oil to remove the heavy components. It is important to maintain the molecular weight of the absorption oil at the design value because the circulating equipment, heat exchangers, and distillation process are designed to utilize a particular molecular weight fluid.

FIG. 16-14

Refrigerated Lean Oil Absorption⁸



Refrigerated Lean Oil

Fig. 16-14 shows a typical refrigerated lean oil absorption process. The actual equipment configuration changes with different gas feeds and product recoveries.

Raw gas enters the plant inlet separator upstream of the main process where inlet liquids are separated. The gas then enters a series of heat exchangers where cold process gas and the refrigerant reduce the feed gas temperature. This reduction in temperature results in condensation of the heavier hydrocarbons in the inlet gas.

The gas is then fed to the bottom of the absorber where it flows upward countercurrent to the lean oil which is introduced at the top of the column. The lean oil has also been chilled to aid in NGL absorption. This column has trays or packing which increase the contact of the gas and lean oil. The lean oil physically absorbs the heavier hydrocarbons from the gas. The lighter components stay in the gas and leave the top of the absorber. The oil and absorbed hydrocarbons leave the bottom of the absorber as "rich oil."

The rich oil flows to the Rich Oil Demethanizer (ROD) where heat is applied to the rich oil stream to drive out the lighter hydrocarbons which were absorbed. Some of the cold lean oil is also fed to the top of the ROD to prevent loss of desirable NGLs from the rich oil.

The rich oil from the ROD is then fed to a fractionation tower or "still." The still is operated at a low pressure and the NGLs are released from the rich oil by the combination of pressure reduction and heat addition in the still. The operation of the still is critical to the overall plant operation as this is not only the point where the desired product is produced, but the lean oil quality from the bottom of the column is important in the absorption of NGLs in the absorber. The refrigeration required for the oil and gas chilling and the heat inputs to the ROD and still are the key parameters which must be controlled to operate a lean oil plant efficiently.

LOW TEMPERATURE NGL RECOVERY PROCESSES

Dew point control and mechanical refrigeration systems are intended for applications where enough liquefiable hydrocarbons must be removed from the gas stream such that the gas can be transported without liquids formation and/or where some of the heavier components must be removed in order to meet a maximum BTU content specification. In these cases, the goal is to meet the natural gas stream sales specification, not recovery of liquids. In cases where the liquefiable hydrocarbons present in a gas stream are more valuable as a liquid product than as the BTU value they contribute to the gas, more effective methods of liquids recovery are desired. The lean oil processes described in the last subsection were the original technique used for this objective, but as already noted, this technology has been superseded by more cost effective cryogenic expansion process methods, which are covered in the following sub-sections.

In order to condense and separate more ethane and propane and thus achieve higher NGL recovery levels than are possible with dew point control or propane-based refrigeration alone, more gas cooling must be provided, which means cryogenic temperatures (below -50°F) are required. In order to achieve these temperatures, a combination of pressure expansion and chilling is used. The two types of cryogenic expansion processes in use today are:

1. Adiabatic J-T expansion across a control valve
2. Isentropic expansion through a turboexpander
3. Variants of each system above with supplemental mechanical refrigeration

The turboexpander processes are the most widely used; however, J-T expansion type processes are sometimes used for lower recovery levels in unattended or remote locations, and usually for gas flow rates less than 20 MMscfd. Turboexpander type processes are used for higher ethane and propane recovery levels with larger gas inlet flow rates where the cost of the additional equipment is easily justified by the value of the additional liquids recovered.

One limitation on the recovery of ethane and heavier products is the effect of liquids extraction on the BTU content of the residue gas. The presence of inerts in the feed gas may limit the amount of ethane that can be removed and still meet the minimum BTU heating value specification for the residue gas.

Another consideration is the liquids recovery level that is justifiable for a given gas composition and plant location. If ethane can be recovered at a good price differential over its value as a fuel in the residue gas, then the additional compression and complexity of a high ethane recovery facility may be justified. If there is no local market for ethane, or no way to transport high vapor pressure product to market, then only propane and heavier components may have an economic incentive for recovery. The plant may be designed for initial operation in a propane recovery mode, with future conversion to ethane recovery if or when the local demand for ethane justifies the expense of additional compression. Many plants in North America are designed for dual mode operation in which the operator can select high ethane recovery or ethane rejection, while still maintaining high propane recovery. These plants can be switched, during normal operation, as often as necessary between ethane recovery and ethane rejection, in response to changes in the processing margin. Historically, recovery of propane and heavier components is almost always justified.

Propane Plus NGL Recovery Compared to Ethane Plus NGL Recovery

Ethane is more volatile than the propane and heavier components in a natural gas stream so generally deeper cooling is required to condense the ethane fraction than is needed to condense only the propane and heavier components. The temperatures needed for ethane recovery are thus colder than those needed for propane recovery regardless of the process design. This means that more compression power is required for high ethane recovery than for propane recovery only because of the higher pressure differential required to produce lower temperatures across the cryogenic sections and/or refrigeration system.

The colder operating temperatures for ethane recovery mean that more of the plant will be constructed of stainless steel or aluminum than may be needed for a facility which only recovers propane. Another consequence of the colder temperatures is that the allowable concentration of CO_2 in the feed to avoid CO_2 freezing will be less for an ethane recovery design than for a propane recovery design.

As stated previously, many process designs can operate in either an ethane recovery mode or in an ethane rejection (propane recovery) mode, as long as the equipment design temperatures are acceptable for both modes and the compression power is sufficient to support the desired ethane recovery level. An ethane recovery design will not be as efficient for propane recovery compared to a propane recovery specific design. It is, however, possible to combine a very high efficiency ethane recovery design with a very high efficiency propane recovery design, if this need is identified during the design phase.⁹

There is always an incentive to minimize the horsepower, emissions, and capital cost for the facility for a given recovery level of ethane and/or propane. A simple J-T expansion type process may have very good economic performance for a given location, flow rate and product price, but it will have low process recovery efficiency. This technology is therefore normally restricted to relatively small plants. Typical J-T cryogenic process configurations are described in the next sub-section. A good understanding of J-T operation will help in understanding the expander plant designs in the subsequent sub-sections. Turboexpander plants are operated in a J-T mode before the turboexpander is started up and when the turboexpander is unavailable. Examples are provided later showing the difference in product recovery for operation in J-T mode compared to turboexpander mode.