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AM-16-65      Screening of Options for LPG Recovery from  
Refinery Fuel Gas Streams

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# Screening of Options for LPG Recovery from Refinery Fuel Gas Streams

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## 1 Abstract

One potential means to increase refinery margins is to recover liquefied petroleum gas (LPG) from refinery fuel gas. The options for LPG recovery are many and varied. The refinery fuel gas stream(s) that will be processed must be identified, and there may be multiple streams to consider. The processing technology to be used for LPG recovery must also be selected. Options for LPG recovery from refinery fuel gas include cryogenic processes (turboexpander and Joule-Thompson variants), straight refrigeration processes, lean-oil processes, membrane processes or membrane process enhancements, adsorption, and others. This paper presents results of an analysis of LPG recovery from fuel gas, and includes comparisons of capital cost, major operating cost, and approximate payback periods for an example oil refinery. The paper illustrates the options available, the choices that have to be made, and the economic effects of LPG recovery for these choices.

Selection of streams from which to recover LPG is an important aspect of LPG recovery. The primary factor that usually determines a stream's suitability for economical LPG recovery is the concentration of desirable recoverable liquid (e.g., propane, butane, etc.) contained in the stream. Some streams may contain components, e.g., propylene, that affect whether or not they should be considered a candidate for LPG recovery, or they may contain species that require special treating prior to LPG recovery. The specifications for the LPG product may also influence which streams are included or excluded from an LPG recovery scheme. The paper considers these factors and demonstrates how the economics are affected.

## 2 Background

Refinery off-gas streams that are burned as fuel often contain significant concentrations of LPG components, e.g., propane, butanes, and other hydrocarbons such as propylene, butylenes, heptanes, etc. Refiners may consider the recovery of an LPG product stream from their refinery fuel gas streams as a way of increasing refinery margins. Trimeric conducted an economic feasibility and technology screening study for the recovery of a mixed LPG stream for some example fuel gas streams at a refinery; the key results of this study are provided in this paper.

Many refineries maintain a fuel gas header that supplies various processes with fuel for burners and other usages; the gas in the header is often supplied from off-gases from a number of refinery process units, and may be supplemented by purchased natural gas. A key issue in the

consideration of recovering LPG from the refinery fuel gases is the differential price or margin between the value of the LPG products on an energy content (BTU) basis, and the purchase price of natural gas on a BTU basis. This is important because LPGs recovered from the fuel gas represent loss of heating value from the fuel gas, which, if the refinery is not long on fuel gas, must then be made up by additional natural gas purchases. The margin for LPG recovery at a refinery may change frequently; LPG prices generally move at a discount to crude prices whereas the cost of natural gas varies with its own supply and demand situation. Currently, with very low crude prices, the LPG margins are likely to be relatively low.

Most refineries have a significant number of gas streams that feed the fuel gas system; additionally, a refinery may have more than one fuel gas header that may operate at different pressures or have other differences. Gas streams in the refinery will have varying concentrations of desired LPG components, such as propane and butane, as well as components that must be limited in LPG products, such as olefins. Part of the consideration and development of an LPG recovery strategy is to evaluate the source fuel gas streams individually, and determine if it makes sense to segregate some streams. The simplest implementation of LPG recovery would usually be to apply the recovery process to the existing mixed fuel gas stream in the fuel gas header(s), without any stream segregation. If it is not desired to recover LPG from the existing mixed fuel gas stream in the header (because, for example, the mixed stream is too lean or contains high concentrations of undesirable components), then LPG recovery could be applied to specific streams. The installation of new piping systems would be required to accommodate the segregation of fuel gas constituent streams for LPG recovery.

A number of technologies are available for recovering LPGs from refinery fuel gases. The technology selection may be dependent upon the characteristic of the refinery fuel gas stream, the volume of the stream, the richness (amount of recoverable LPG) of the gas, and the product specification desired. A section of this document will present some of the technologies often used for LPG recovery, and how the technology choice is influenced by both the characteristics of refinery fuel gases and the needs of the refiner.

Raw LPG streams from recovery processes will usually be primarily a mixture of the C3 and C4 hydrocarbons, and will also include some ethane and C5+ components. Often the raw LPG stream can be fed to existing LPG recovery (saturated gases plant, or Sats Gas plant for short) distillation trains at the refinery, which may include sweetening and deethanizer, depropanizer, and debutanizer columns. This paper focuses on the production of the raw LPG stream, rather than its subsequent treatment. Although ethane can also potentially be recovered from fuel gas streams along with the C3+ (depending upon the technology chosen), part of the basis of this project was that the additional cost of recovering the ethane would not be merited.

### **3 Technologies for LPG Recovery from Refinery Gas Streams**

Propane and higher hydrocarbons are frequently recovered from natural gas streams, refinery gas streams, and other gas streams. Many technologies have been applied to the recovery of LPG components from gas streams in various industries. This section summarizes the LPG recovery processes considered in the screening study.

There are a number of commercially viable approaches to recovering LPGs from the fuel gas to meet the necessary marketing specifications. The ones considered in this paper are:

- Straight Refrigeration
- Lean Oil Absorption
  - Ambient Cooling
  - Refrigeration Cooling
- Cryogenic Processing
  - Joule-Thomson Expansion variants
  - Turboexpander variants
- Other technologies, such as membranes and adsorption

Each of these processes has advantages relative to the others for certain applications. A brief description of the technologies is given in the following subsections.

### *3.1 Straight Refrigeration*

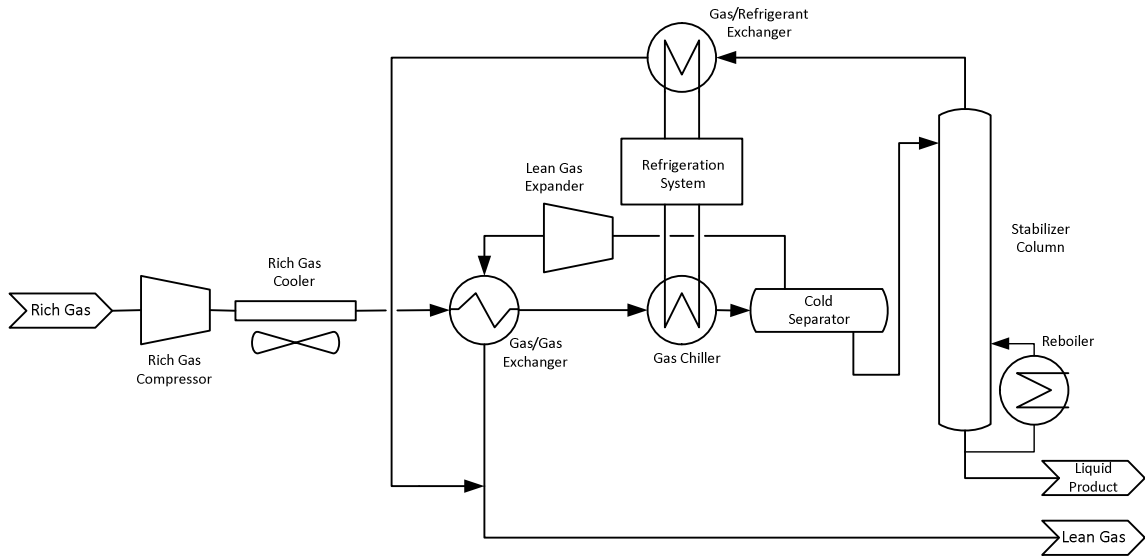
Figure 3.1 shows a schematic of the basic straight refrigeration process for LPG recovery, as it might be applied to low-pressure refinery fuel gases. The basic approach is to compress, cool, and then chill the rich gas further to preferentially condense the components of interest (e.g., C3+). The condensed liquid is then separated from the remaining lean (contains few recoverable LPGs) gas, and the condensed liquid is fed to a stabilization distillation column. The stabilization column strips light compounds from the liquid product, with the lights from the top of the stabilization column subsequently recombined with the lean fuel gas. An important feature typical of straight refrigeration is that the full gas stream does not enter the stabilization (distillation) column. Only the relatively small condensed liquid stream enters the column, and the column is usually (but not always) designed for stripping only, with a relatively small boil-up ratio. Because of this, the distillation column in straight refrigeration processes is usually of a relatively small diameter compared to the diameter that would be required were the entire feed gas stream to enter the column. The straight refrigeration process recovers less of some of the LPG components (e.g., C3) than an equivalent process in which the entire gas stream is distilled, but the capital cost saving with straight refrigeration may merit the reduced LPG recovery. Although C3 and C4 recovery in the straight refrigeration process may be significantly lower than 100%, C5 and heavier components are relatively easy to recover at high efficiency.

A number of variations of the straight refrigeration process are available. Some of these variations are described in the GPSA Engineering Data Book<sup>1</sup>. The various configurations for straight refrigeration represent different ways of tuning LPG fractional recovery, purity of product, and refrigeration duty, and balancing capital cost with recovery efficiency. When applied to typical low-pressure fuel gas streams at refineries, compression of the feed gas is required to operate these systems at typical propane or propylene refrigeration temperatures (e.g., -40°F). Given that refineries will usually not desire the lean fuel gas to be returned to the header at the higher pressures at which these units operate, expansion of the lean gas back to the fuel gas header pressure can be used to recover some work.

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<sup>1</sup> Gas Processors Suppliers Association, Engineering Data Book, 13<sup>th</sup> Ed., 2012, pg. 16-9.

**Figure 3.1: Basic Straight Refrigeration Process for Liquids Recovery from Fuel Gas**



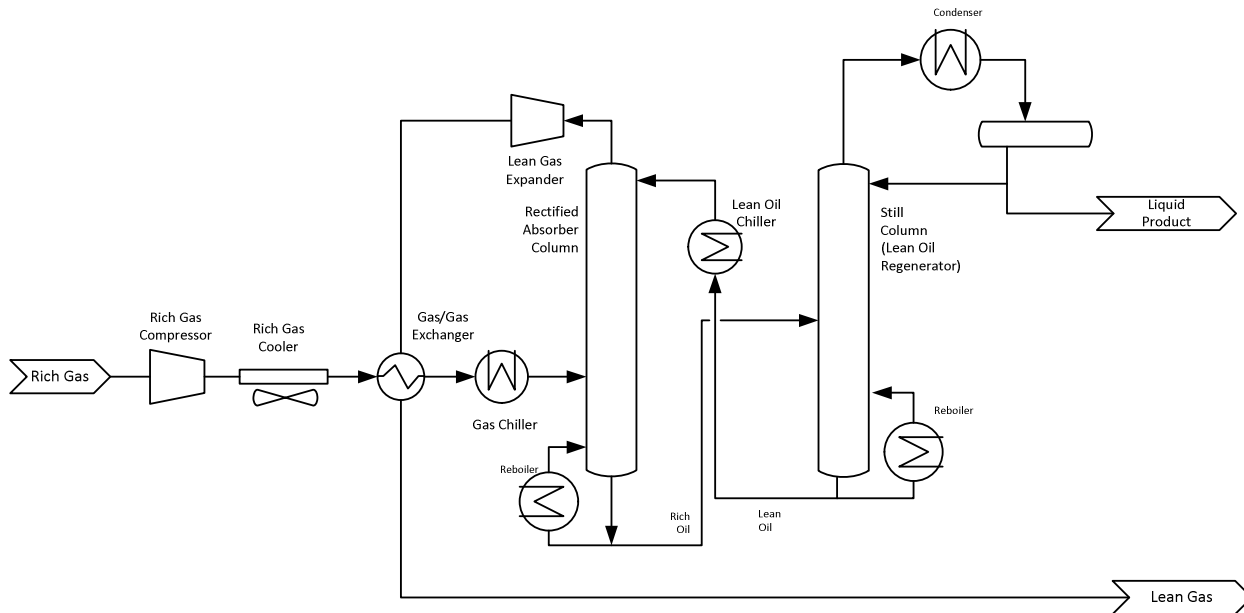
### 3.2 Lean Oil Absorption

The absorption of LPG components from fuel gases using a heavier hydrocarbon liquid stream is another widely practiced technology. Most refineries can identify an existing stream that would have the appropriate properties to act as an absorbent working fluid. Operation of the absorption process at reduced temperatures (using refrigeration) is often done to enhance the recovery of the lighter portions of the target LPG components, although operation at approximately ambient temperature has been commonly practiced.

A schematic of a lean oil absorption process that might be applied to the treatment of low-pressure-refinery fuel gas is shown in Figure 3.2. In the process shown in the figure, the rich fuel gas must be compressed to a moderate pressure, cooled, and contacted with the cooled/chilled lean oil in a rectified absorber column. The lean oil absorbs LPG components preferentially in the column, but also will absorb a smaller fraction of the ethane and methane. The rectified absorber column has a reboiler that serves to vaporize most of the ethane and methane absorbed by the oil and causes it to exit the column overhead with the lean gas. The rich oil – containing most of the LPG components from the gas – from the bottom of the absorber column is then regenerated in the still column. The LPG components are removed as the overhead liquid product of the still column, and the regenerated lean oil is returned to the rectified absorber column. The lean gas from the top of the rectified absorber column is then returned to the fuel gas header pressure; this can be done through an expander, if desired, to provide some of the rich gas compression work.



**Figure 3.2: Basic Lean Oil Absorption Process for Liquids Recovery from Fuel Gas**



There are many variations of the lean oil absorption process, one of the main variations being between refrigerated and non-refrigerated processes. If a relatively light oil is used as the absorbent, it may be necessary to operate the process in the refrigerated mode in order to avoid losses of some of the oil by evaporation in the absorber, additionally refrigeration may be required to increase recovery of propane to the desired level.

### 3.3 Cryogenic Processes

Low-temperature processes for LPG recovery typically involve the use of turboexpansion and/or Joule-Thomson valve (JT-valve) throttling to create low-temperature conditions required for high recovery percentages of LPGs (and ethane, if desired). Operation of the process in this mode can result in higher recovery of propane compared to straight refrigeration or lean oil absorption processes. Whether a JT-valve or a turboexpander is specified to handle the pressure drop of the main gas feed stream is a significant issue, and this choice is strongly influenced by the composition of the feed gas.

The expansion of a gas through a JT-valve is approximated as an isenthalpic process, and will only produce a colder temperature in the expanded gas compared to the feed if the JT coefficient of the gas is positive (at the operating condition). While many components that are present in refinery fuel gases have positive JT coefficients, a notable exception that has a negative JT coefficient (at the temperatures and pressures at which processing would normally occur) is hydrogen. Therefore, refinery gases that contain high fractions of hydrogen may not be effectively treated in low-temperature processes that rely on JT-valves for autorefrigeration. Hydrogen is frequently a large fraction of fuel gases, originating from the various common

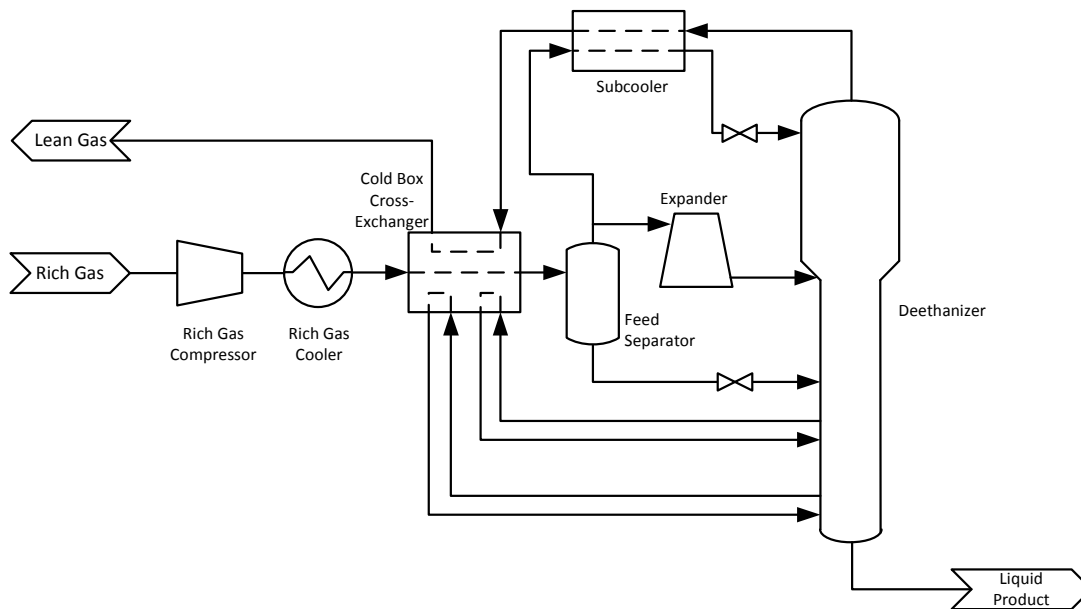


hydrotreating unit operations; therefore the use of JT-expansion-based processes may only apply to a limited number of refinery fuel gas cases.

The expansion of a gas through a turboexpander can be approximated as an isentropic process because work is extracted from the gas as it expands. The negative JT coefficient of hydrogen does not prevent the gas from cooling upon expansion due to the extraction of work from the stream during the expansion. Therefore, when considering refinery gases that contain a high fraction of hydrogen, usually JT-valve-based processes would not be considered, but turboexpansion processes are valid for consideration.

Trimeric simulated a turboexpansion based process for the recovery of LPG from refinery fuel gas streams, based loosely on the Gas Subcooled Process<sup>2</sup>, assuming it to be representative of the turboexpansion family of processes. Figure 3.3 illustrates the process flow scheme assumed for this study. Many other configurations of low temperature distillation for recovery of LPG from fuel gas streams using turboexpansion are possible.

**Figure 3.3: A Cryogenic Process for Liquids Recovery from Fuel Gas**



In this process, the feed gas is compressed, cooled via cross exchange, and liquids separated from the gas phase. Both phases are then throttled to a lower pressure, with most of the gas phase passing through an expander that is used to recover work that can be used to partially satisfy the compression requirement, while another part of the gas is further cooled/condensed by cross exchange with the column overhead and fed to the top of the column. The column operates at temperature and pressure conditions such that the reboiler duty is supplied by cross exchange with the feed, and the liquid feed to the top of the column is supplied by the throttled/chilled feed stream.

<sup>2</sup> *ibid*, pg. 16-14.

### 3.4 Other Processes

Other processes have been developed for the recovery of LPG components from gas streams, and only a few of the predominant technologies were considered in detail in this study. Other technologies that can potentially be applied include membranes and adsorption. Some types of membranes can be used to enrich gas streams in heavier components, and other membranes can be used to separate and recover a hydrogen-rich product stream.

Often membranes are used in conjunction with refrigeration-based processes, serving to enrich the gas to make the condensation of the liquids easier. Although the use of membranes to enrich feed gases is a viable option, it was not considered for application in this study since the gas feed was considered sufficiently rich without membrane treatment.

Adsorbents can be used to preferentially adsorb hydrocarbons from gases. Short-cycle temperature/pressure swing processes may be used to recover LPG components from gases, although these processes are typically more applicable to the recovery of C<sub>5</sub>+ hydrocarbons and for hydrocarbon dew point control. Adsorption-based processes for LPG recovery are probably not competitive with the other processes considered in detail here for this fuel gas LPG recovery application.

## 4 Targeting Gas Streams for LPG Recovery

The first step in considering fuel gas streams for LPG recovery is to determine whether or not the recovery of LPG from fuel gases has significant economic incentive; specifically, the value of the recoverable LPG components per BTU should be compared to the cost of the additional supplemental natural gas that would be required to replace the BTU value of the recovered LPG. Once it has been established that the potential incremental value of the LPG is significant, the fuel gas streams in the refinery can be considered.

Gas richness (recoverable LPG content) is typically characterized in terms of the gallons of recoverable liquid products per thousand standard cubic feet of gas, or GPM. Gases containing below 2-3 GPM propane and higher hydrocarbons are generally considered to be lean in LPG's. Gas containing above 5-7 GPM propane and higher hydrocarbons is typically referred to as rich gas.

The concentration of LPG components in the mixed gas stream in the fuel gas header, as well as the concentrations in each of the major fuel gas streams that feed the header, should be evaluated. Gas streams richer in LPG components are usually more economical to apply LPG recovery to versus leaner streams. It may be desired to consider segregation of the leaner constituent gas streams from richer constituent streams, and subject only the richer streams to LPG recovery; this may result in the need to modify piping and controls associated with the fuel gas header system. Additionally, some refinery gas streams originate from H<sub>2</sub>S-containing unit operations, such as hydrotreaters. Sour refinery gases are often combined and processed in a central H<sub>2</sub>S-removal operation, such as an amine scrubber. Integration of any new LPG recovery with existing sweetening systems may also be an important consideration. Full consideration of the issue of whether to treat the existing comingled fuel gas stream as-is in the header versus the

modification of the fuel gas header system to segregate the richer and lean constituent gas streams is very dependent upon the situation at the specific refinery, and must be defined on a case-to-case basis.

Another important factor in the consideration of fuel gas streams for LPG recovery is the presence of components that could result in the final LPG products being off-spec. For example, the HD-5 propane specification requires that the concentration of propylene in the propane product be no more than 5 volume%. Because any propylene in the recovered LPG will likely end up in the propane product, the ratio of propane to propylene must be high enough to prevent violation of the propane product specification.

Water content of the gas stream can also be important, depending upon the type of technologies being considered. The application of refrigerated or lower temperature (cryogenic) technologies requires low water content of the feed gas. Very low temperature processes may require that the feed gas be dried using molecular sieves, while other refrigeration-based technologies operating at somewhat warmer temperatures may only require protection against hydrate formation by the use of glycol or methanol injection into the inlet gas stream as it is chilled.

In an example refinery case, fuel gas from two mixed headers, along with two fuel gas streams from individual units were considered for LPG recovery. Characteristic data from the four streams are given in Table 4.1.

**Table 4.1. Characteristic Data of Fuel Gas Streams Considered for LPG Recovery.**

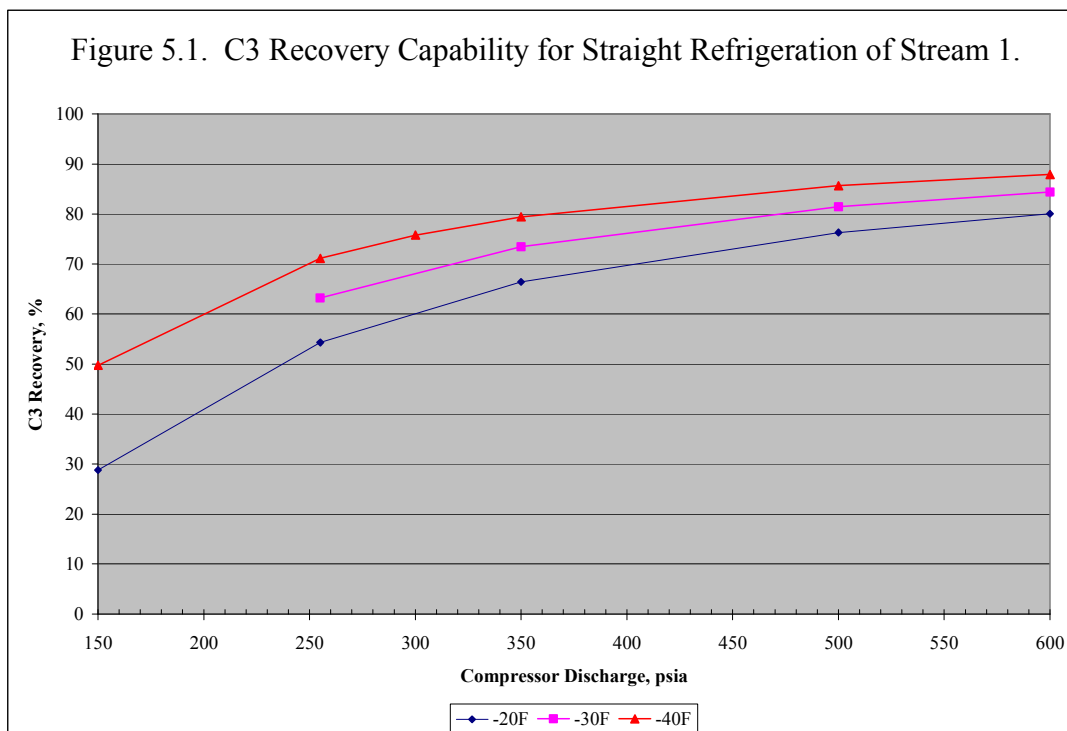
		<b>Stream 1</b>	<b>Stream 2</b>	<b>Stream 3</b>	<b>Stream 4</b>
		Mixed LP Header	Mixed HP Header	Separate Unit 1	Separate Unit 2
<b>Flowrate</b>	MMscfd	5	3	1	1.6
<b>Pressure</b>	psia	80	130	200	240
<b>Richness</b>	gal C3+/MSCF	5	5	11	5
<b>Composition (partial), mole percent</b>					
Hydrogen		55	75	37	70
Oxygen		0.3	0	0.4	0.8
H <sub>2</sub> S		0.01	<0.01	<0.01	<0.01
Methane		12	5	5	5
Ethane		14	4	19	4
Propane		10	5	24	5
Propylene		0.1	0	0.2	0
C4's		5	6	11	11
C5+		1	4	3	2
Water		1	0.7	0	0

## 5 Technical and Economic Evaluation

### 5.1 Straight Refrigeration

Straight refrigeration systems can be designed to use refrigeration operation temperatures that are easily obtained with common refrigeration systems (e.g., propane), and also do not require the use of special metallurgy. A range of -20 to -40°F was selected as the basis for this evaluation because this temperature range is typically used to achieve reasonably high LPG recovery in natural gas processing applications for relatively rich gas streams. Propane was arbitrarily selected as refrigerant since it is a very common refrigerant for straight refrigeration processes and it is commonly operated down to about -40°F or slightly lower. As an alternate to propane, there are Freon replacements which would perform comparably; also propylene, which is often produced at refineries, can also be used as a refrigerant to achieve similar, or marginally lower, refrigeration temperatures. A temperature of -50°F is the limit for Charpy impact-tested carbon steel; going to lower temperatures requires upgraded metallurgy and the associated higher capital cost.

Figure 5.1 shows a graph of the predicted propane recovery that can be expected from Stream 1 through the range of operating temperatures considered, with the compression outlet pressure varied from 150 to 600 psia, using the type of system shown in in Figure 3.1. For this gas stream, based on some preliminary economics based on compression cost, it was estimated that compression to about 300 psia and chilling to -40°F was a reasonable basis for the recovery of LPG from Stream 1. This case results in the recovery of about 75% of the propane and 93% of the combined propane and heavier material.



Cases using straight refrigeration were considered where only the low pressure (Stream 1) gas was considered, where the combined high pressure and low pressure streams (1 and 2) were included, and where all of the streams were treated together. Different compression pressures were also considered as part of this analysis.

Process simulation, early-phase equipment design techniques, and factored cost estimates were used to generate capital costs (~2008 basis) for the straight refrigeration systems for the different configurations.

Operating costs considered included compression (refrigeration) power, and the purchase of additional natural gas to replace the lost BTU value of the recovered LPGs. The product LPGs were valued based on typical values for the separated products.

Based on consideration of the operating costs and increased revenue, a payback period for the capital cost was estimated. The results of the study for straight refrigeration processes are shown in Table 5.1.

**Table 5.1. Economic Evaluation of Straight Refrigeration Cases.**

Parameter	Units	Case 1 Stream 1 only	Case 2 Stream 1 only	Case 3 Stream 1 + 2	Case 4 All Streams
Gas flow rate	MMscfd	5	5	8	10.5
C3+ content of inlet gas	gpm/Mscf	5	5	5	5
Compression Pressure	psia	300	365	320	300
Recovery of C3 and C4s	%	76 / 93	80 / 95	75 / 93	77 / 94
CAPEX	MM\$	8.4	8.8	10.8	12.4
Compression/Refrig OPEX	MM\$/y	0.23	0.26	0.31	0.34
Natural Gas Cost	MM\$/y	5.3	5.5	8.7	13.3
TOTAL OPEX	MM\$/y	5.6	5.8	9.0	13.6
Gross Revenue	MM\$/y	8.0	8.3	13.6	21.0
Net Revenue	MM\$/y	2.4	2.5	4.6	7.3
Total Treating Cost	\$/Mscf	3.9	4.1	3.8	4.2
Payback	years	3.5	3.6	2.3	1.7

The economics in Table 5.1 do not include some non-process operating costs, such as maintenance, additional operations labor required, or other incremental overheads. However, the payback period for the recovery for these cases is relatively low and even if the additional cost factors were included, the payback period would still be favorably low.

## 5.2 Lean Oil Absorption

For the evaluation of lean oil absorption to the recovery of LPGs for this application, a source of lean oil must be identified. Although the lean oil will be continually regenerated and reused in the absorption process, some purge and makeup will occasionally be required, so it is most convenient at a refinery to use an existing refinery stream. For this work, a 54 degree API gravity naphtha stream having a 200 to 350°F boiling range was chosen as a convenient representative stream that is available in refineries. This stream is considered to be lighter than

optimum for use in lean oil absorption in the desired absorption temperature range of 50 to 100°F since fairly high oil losses would be experienced, which would be reflected in lower overall C5+ recoveries. A higher molecular weight and boiling range refinery stream would have been more suitable for the example operation if one had been available.

As with the straight refrigeration processes, liquid recoveries with a lean oil process increase with decreasing temperature and increasing operating pressure; but there is one additional variable: increasing lean oil circulation rate. The combination of these three factors determines recovery of light hydrocarbons absorbed from the gas by the lean oil. (Another variable, lean oil molecular weight, was not considered since this study considered the above defined lean oil exclusively.) Although a refrigerated lean oil process with oil and gas temperatures well below 0°F would have been possible, at some point the refrigeration effect would dominate, negating the benefits of the lean oil itself. The lowest temperature considered for chilling the oil or gas was +45°F which is above the hydrate-formation temperature of either the gas or lean oil; this operating temperature basis results in the capability to recover LPG from the gas without dehydration (other than simple condensing).

The basic lean oil process that was considered in the economic evaluation is shown in Figure 3.2, and it consists of an absorber and still column for regeneration. Compressor discharge pressures were varied from 150 to 350 psia, lean oil and gas temperatures from 90°F (achieved with cooling water) down to 45°F (achieved by either mechanical refrigeration or by expanding the residue gas down to LP fuel system pressure) and lean oil circulation rates from 25 to 146 gpm.

Cases considering various operating conditions, and feed fuel gas streams were considered by simulating a flowsheet similar to Figure 3.2. Optimization variations included dual turboexpanders in series on the absorber overhead residue gas stream for improved energy efficiency. Methanol injection was assumed upstream of the first turboexpander for hydrate inhibition. A summary of the simulation material balance results for two lean oil absorption cases are shown in Table 5.2.

Process simulation, early-phase equipment design techniques, and factored cost estimates were used to generate capital costs (~2008 basis) for the lean oil absorption systems for the different configurations. Operating costs considered included compression power, the purchase of additional natural gas to replace the lost BTU value of the recovered LPGs, still reboiler heating duty, and methanol injection for hydrate inhibition. The product LPGs were valued based on typical values for the separated products.

Based on consideration of the operating costs and increased revenue, a payback period for the capital cost was estimated. The results of the economic evaluation for Lean Oil Absorption are shown in Table 5.3.

Similar to the straight refrigeration cases, the economics in Table 5.3 do not include some non-process operating costs, such as maintenance, additional operations labor required, or other incremental overheads. Consideration of these factors would result in the payback period being nominally higher.

**Table 5.2: Summary Material Balance Information for Lean Oil Cases**

Case Description	Case 5 Stream 1 & 2	Case 6 All Streams
Compressor disch P, psia	345	345
Absorber Inlet gas T, F	45	45
Lean oil rate, gpm	102	117
Lean oil T, F	45	45
Rich Oil Contents	BPD	BPD
H <sub>2</sub>	8.68	12.23
O <sub>2</sub>	0.13	0.28
C1	17.43	21.92
C2	163.67	223.22
C3	366.51	553.43
C3=	2.66	3.49
C4s	309	521
C5+	152	194
Lean oil losses	9	12
Product Recovery	%	%
C3s	88	89
iC4	100	100
nC4	100	100
iC5+ (net)	94	94

**Table 5.3. Economic Evaluation of Lean Oil Absorption Cases.**

Parameter	Units	Case 5 Stream 1 & 2	Case 6 All Streams
Gas flow rate	MMscfd	8	10.5
C3+ content of inlet gas	gpm/Mscf	5	5
CAPEX	MM\$	15	17
Compression and Refrig OPEX	MM\$/y	0.4	0.4
Natural Gas Cost	MM\$/y	9.8	15.0
TOTAL OPEX	MM\$/y	10.1	15.4
Gross Revenue	MM\$/y	14.8	22.7
Net Revenue	MM\$/y	4.6	7.2
Total Treating Cost	\$/Mscf	4.6	4.9
Payback	years	3.2	2.3

### 5.3 Cryogenic Processing

For the evaluation of cryogenic processing for LPG recovery from the fuel gas streams, Trimeric simulated the process scheme shown in Figure 3.3 and estimated capital and operating cost parameters based on the simulations. Inlet gas is compressed to 790 psia, cooled to 120°F with air, then 90°F with water. The gas must then be dehydrated by a molecular sieve system prior to entering the cold box.

The inlet gas is cooled in the cold box to about -1°F and then is chilled to -30°F by a propane refrigeration system. The partially condensed -30°F stream is separated with the cold liquid expanded across a valve, then fed to the cold box and then serving as the bottom feed to the fractionator. The -30°F vapor is split into two streams; the larger 90% fraction is expanded across a turboexpander and then fed as the middle feed to the fractionator. The smaller 10% fraction exchanges heat with cold fractionator overhead and then is fed to the top of the fractionator.

Overhead from the column is fed to two exchangers in series (the overhead exchanger mentioned above and then the cold box). Leaving the cold box, this stream is at about 70°F. It is then expanded to the desired fuel system pressure across a second turboexpander. It then re-enters the cold box as the 3rd cold stream and is heated to approximately 70°F prior to being directed into the LP fuel system. Power recovered from the 2 turboexpanders is used to partially offset the inlet gas compression demand.

It was found that by compressing the inlet gas to 790 psia and expanding to 190 psia, 95% of the C3s and close to 100% of the C4+ could be recovered. Note that the low operating temperature (down to -120°F) sections of the plant would require significant use of expensive stainless steel metallurgy.

Capital and operating cost parameters were estimated using similar methods to the other technologies. Operating and capital costs are shown in Table 5.4.

**Table 5.4. Economic Evaluation of Cryogenic Processing Cases.**

Parameter	Units	Case 7 Stream 1	Case 8 Stream 1 & 2
Gas flow rate	MMscfd	5	8
C3+ content of inlet gas	gpm/Mscf	5	5
CAPEX	MM\$	20.1	27.1
Compression and Refrig OPEX	MM\$/y	0.32	0.43
Natural Gas Cost	MM\$/y	6.2	9.9
Total OPEX	MM\$/y	6.5	10.3
Gross Revenue	MM\$/y	9.1	15.2
Net Revenue	MM\$/y	2.6	4.9
Total Treating Cost	\$/Mscf	5.7	5.5
Payback	years	7.6	5.5

The economics in Table 5.4 do not show some non-process operating costs, such as maintenance, additional operations labor required, or other incremental overheads. Consideration of these factors would result in the payback period being nominally higher.



## 6 Conclusions

The comparison of the economic analysis of the technologies considered yielded a clear bias for the straight refrigeration process for the recovery of LPGs from these gas streams. The economic viability of recovering LPGs from the fuel gas streams increases as more of the candidate fuel gas streams that were considered in this study are included in the recovery process; however, the practical issues associated with making these streams available for recovery (e.g., new piping systems, modifications to other systems) are not included in the economic analysis.

A summary of the economic estimates for each recovery technology is shown in Table 6.1. In this table, only the cases are given in which gas stream 1 and gas stream 2 are both subjected to LPG recovery together in the same system, so these cases should be directly comparable.

**Table 6.1. Economic Summary for LPG Recovery (Stream 1 + Stream 2) using Three Technologies.**

Parameter	Units	Straight Refrigeration Case 3	Lean Oil Absorption Case 5	Cryogenic Processing Case 8
Gas flow rate	MMscfd	8	8	8
C3+ content of inlet gas	gpm/Mscf	5	5	5
CAPEX	MM\$	10.8	15	27.1
Compression/Refrig OPEX	MM\$/y	0.31	0.4	0.43
Natural Gas Cost	MM\$/y	8.7	9.8	9.9
TOTAL OPEX	MM\$/y	9.0	10.1	10.3
Gross Revenue	MM\$/y	13.6	14.8	15.2
Net Revenue	MM\$/y	4.6	4.6	4.9
Total Treating Cost	\$/Mscf	3.8	4.6	5.5
Payback	years	2.3	3.2	5.5

Although the economic analysis did not include all possible cost factors – insofar as that the costs associated with maintenance, overheads, additional operations staff, etc., were not included in the economic analysis – the overall outcome indicated that the recovery of LPG from these fuel gas streams was an economically viable proposition, given the values that were assumed for power, fuel, and the product LPGs.