LNG Production from Natural Gas

Submitted for the Partial fulfillment of

BACHELOR OF TECHNOLOGY
(Applied Petroleum Engineering)

(Session: September 2005 to May 2009)

Submitted to

University of Petroleum & Energy Studies, Bidholi, Dehradun

Under the able guidance of: Mr. Rajeshwar Mahajan

Submitted By:

Karmendra Mohan Singh

&

Sonit Balyan

University of Petroleum & Energy Studies, Bidholi, Dehradun

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CERTIFICATE

This is to certify that the dissertation report on **"LNG Production from Natural Gas"** completed and submitted to the University of Petroleum & Energy Studies, Dehradun by Mr. Karmendra Mohan Singh & Mr. Sonit Balyan in partial fulfillment of the requirement for the award of degree of Bachelor of Applied Petroleum Engineering is a bonafide work carried out by them under my supervision and guidance.

To the best of my knowledge and belief the work has been based on investigation made, data collected and analyzed by them and this work has not been submitted elsewhere for a degree.

Rajeshwar Mahajan
11-05-08.

Mr. Rajeshwar Mahajan

Assistant Professor

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U.P.E.S.

ACKNOWLEDGMENT

This project bears imprints of many people.

First and foremost, we are thankful to University of Petroleum & Energy Studies for giving us opportunity to carry out our major project on **"LNG Production from Natural Gas"**.

We are indebted to our mentor, ***Mr. Rajeshwar Mahajan*** **(Assistant Professor, College of Engineering)**, for his constant support and guidance during this study. We would like to express our sincere gratitude for his patience and encouragement throughout this work, without which this work would not have been possible.

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At last we would like to thank all of our colleagues for making our stay at UPES memorable.

Karmendra Mohan Singh

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ABSTRACT

Natural gas is becoming one of the most important primary energy sources for the 21st century. This is due to large reserves throughout the world and because it is relatively clean fuel. LNG is natural gas cooled until it condenses into a clear liquid. LNG is stored at -162° Celsius (-260°F). At this temperature LNG takes up far less space about 1/600th of its original volume as a gas. LNG can be stored safely in insulated tanks built to withstand its cold temperature. While LNG is reasonably costly to produce, advances in technology are reducing the costs associated with the liquefaction and regasification of LNG. Because it is easy to transport, LNG can serve to make economical those stranded natural gas deposits for which the construction of pipelines is uneconomical. World LNG movement today is about 120 billion cubic meters out of a total natural gas trade of 450 billion cubic meters per annum.

As a part of this project, Scenario of LNG liquefaction plants today and future projections will be discussed. LNG liquefaction is one of the most important part of the LNG value chain and it occupies a substantial portion of LNG value costs. For many years there was absolutely no problem in choosing the liquefaction process for a new LNG plant. The only choice was C_3/MR process. This process has been implemented repeatedly with improvements. The liquefaction systems described are:

- Propane Mixed refrigerant Process (C_3/MR)
- Liquefin Process and Cascade Process
- Dual Mixed Refrigerant Process (DMR)

The selection of best technology is a complex affair, extremely sensitive to a number of design parameters that requires careful attention. The refrigerants which can be used are:

- Methane
- Liquid nitrogen
- Ethylene
- Propane

Base load natural gas liquefaction plants have always been designed along the train concept. While some plants consist of single train, most LNG facilities have several parallel process trains. Large trains can be feasible and cost-effective. The train concept has evolved because of market demand, delivery and shipping logistics and mechanical limitations on the size of critical equipment, such as mechanical drivers, compressors and heat exchangers. Each train usually includes gas treating, dehydration, mercury removal, refrigeration, heavy hydrocarbon removal and LNG production. While the trains are supplied by outside utilities such as electric power, cooling water, instrument air, each is an independent process plant. A single train can be shut

down for a period of months without affecting the operations of adjacent trains. Many factors considered when choosing optimal train size for an LNG project are

- Gas deliverability from the field
- Market demand and LNG delivery build up profile
- Overall optimization of production, storage and shipping logistics
- Available proven equipment size
- Potential capital cost savings

As heat exchanger is the main part of the liquefaction cycle so it size should be optimum. So in this project we have designed heat exchanger for the pre-cooling and liquefaction for a certain feed gas.

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List of Abbreviations

tcm: trillion cubic meters

bcm: billion cubic meters

LNG: Liquefied Natural Gas

CNG: Compressed Natural Gas

SCM: Standard Cubic Meters

NG: Natural Gas

MTPA: Million Tons per Annum

ppm: Parts Per Million

Btu: British thermal unit

Scf: Standard Cubic Feet

Kcal: Kilo Calorie

btoe: Billion tons of oil equivalent

MMTPA: Million Metric Tons per Annum

TCF: Trillion Cubic Feet

tpy: Tons per year

MCHE: Main Cryogenic Heat Exchanger

MMscfd: Million Metric Standard Cubic Feet per Day

MMscmd: Million Metric Standard Cubic Meters per Day

CHAPTER 1

Introduction

1.1) Natural Gas:

Natural gas is a combustible mixture of hydrocarbon gases consisting primarily methane, with low concentrations of other hydrocarbons, water, carbon dioxide, nitrogen, oxygen and some sulfur compounds. In 2005, natural gas accounted for 23.5% of the global primary energy balance, in third position after oil (36.4%) and coal (27.8%) and before hydro electricity (6.3%) and nuclear energy (5.9%). Since the 1980s, the amount of proven reserves has kept rising and was, at the end of 2005, estimated at around 175.2 tcm, which equals 60 years of global consumption. Much of this gas is located in regions far from consuming markets. Worldwide production of natural gas is estimated as being around 2763 bcm.The annual increase was 1.5% in 1995 and 2.5% in 2005.

It is preferred because:

- It is a colorless, shapeless and odorless in its pure form.
- It is a homogeneous fluid of low density & viscosity.
- Natural gas is combustible, and when burned it gives off a great deal of energy.
- Natural gas is clean burning and emits lower levels of potentially harmful byproducts into the air .

1.2) Composition of Natural Gas:

Table1.1 Typical Composition of Natural Gas

Methane	CH_4	70-98%
Ethane	C_2H_6	
Propane	C_3H_8	0-20%
Butane	C_4H_{10}	
Carbon Dioxide	CO_2	0-8%
Oxygen	O_2	0-0.2%
Nitrogen	N_2	0-5%
Hydrogen Sulphide	H_2S	0-5%
Rare gases	Ar,He,Ne,Xe	Traces

1.3) Natural Gas Transportation Methods:

- Cross border pipelines
- LNG
- CNG by ship

1.3.1) Transportation cost Comparison:

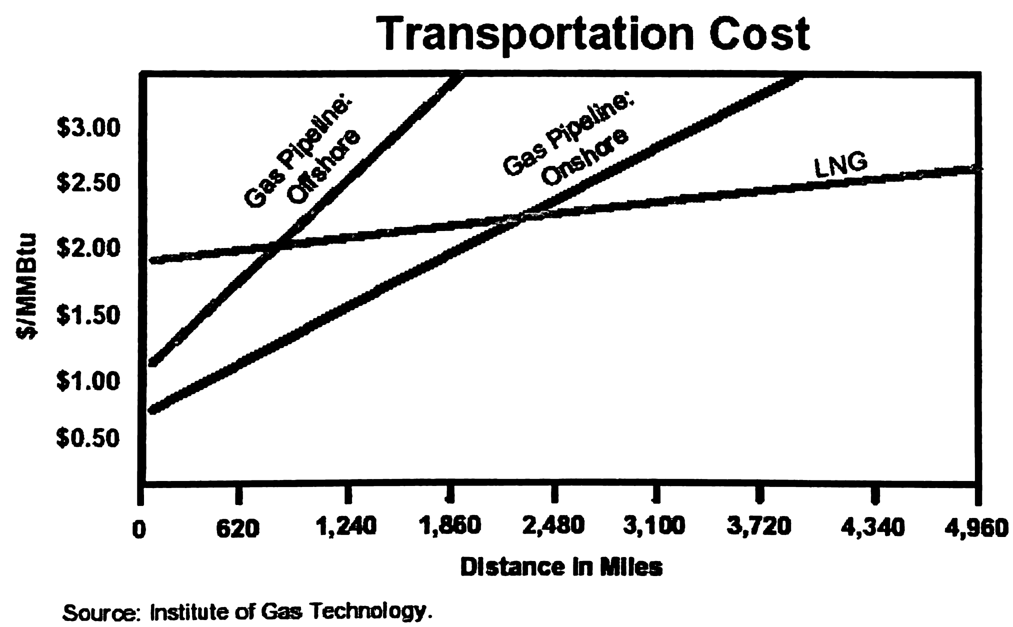

Figure 1.1 Transportation cost of natural gas by various modes

The main difficulty in conveying gas to the markets is its transport over long distances. Liquefaction makes it possible to market important gas reserves located in remote areas, far away from consumer countries, as it reduces the volume by 600 times of natural gas(1 m^3 LNG = 600 SCM of NG). Hence, since its inception in the 1960s, LNG has grown significantly and makes a strong contribution to meeting the world's energy needs it now represents nearly a quarter of worldwide gas exports and worldwide production is expected to more than double from 2004 to 2010 and reach more than 300 million tons per annum (MTPA). Therefore, LNG is a fast expanding energy vector and, although currently it is commonly evaporated and compressed into long-distance pipelines, new solutions are bringing LNG directly to off-pipeline customers, thus opening a new segment of application. LNG technology is constantly evolving to meet the increasing demand and requires the use of increasingly complex technology.

1.4)Liquefied Natural Gas:

When natural gas is cooled to a temperature of approximately -161°C at an atmospheric pressure, it condenses to Liquefied Natural Gas (LNG). Liquefaction reduces the volume by approximately 600 times, thus making it more economical to transport over long distance across the countries.

1.4.1) Composition of LNG:

Natural gas is composed primarily of methane but may also contain ethane, propane and heavier hydrocarbons. Small quantities of nitrogen, oxygen, carbon dioxide, sulfur compounds and water may also be found in natural gas.

The liquefaction process for LNG requires natural gas free from impurities like Co2, sulfur compounds, water vapor, mercury etc

Table 1.2 LNG Product Specification

Component	Mole %
Nitrogen	1.0 (max)
Butanes+	2.0 (max)
Pentanes+	0.1 (max)
Carbon dioxide	50 ppm (vol) (max)
Hydrogen Sulphide	5 mg/ Sm^3 (max)
Total Sulfur	30 mg/Sm^3 (max)
Aromatic	2 ppm (vol) max
Water	0.1 ppm (vol) max
Heating value	1040 to 1150 Btu/ Scf, or 9000 to 10,000 kcal/ Scm

1.4.2) Properties of LNG:

- Extremely low temperature - 260°F (-162°C).
- LNG is nontoxic.
- Odorless and colorless: LNG looks like boiling water. When exposed to atmospheric temperatures and pressure, it vaporizes to about 600 times its liquid volume.
- Vapor Dissipation : As the vapor warms to - 160°F (- 107°C), it becomes lighter than air and will dissipate
- LNG weighs less than one-half that of water, actually about 45% as much.
- When vaporized it burns only in concentrations of 5% to 15% when mixed with air.

1.4.3) Need for LNG:

LNG has been the world's fastest growing energy option over the last two decades. Its market has quadruplicated in size since 1980. A total of 143 bcm of gas was carried by ship in 2001 which is 25% of world's gas export. The remaining 75% natural gas export has been by pipeline.

The world's proven reserves of natural gas are abundant and estimated to be more than 155.8 tcm equivalent to approximately 140 btoe. This quantity is almost three quarters of the world's proven oil reserves. The geographical distribution of natural gas reserves is very different from that of oil. Whereas more than three quarters of world's oil reserves are located in OPEC countries, equivalent figure for gas is only about 1/3rd.

1.4.4) The drivers for the use of LNG:

- Growing demand for fossil fuel
- Versatility of usage
- Environmental awareness-Natural gas being the cleanest fossil fuel, its application is getting global preference over other fossil fuels.
- Security of supply-LNG contracts are normally of long term with commitment for high investment on either side of the LNG value chain i.e., liquefaction, transportation and re-gasification.
- Pricing and delivery flexibility-LNG can be delivered at any of the re-gasification terminal. Pricing can also be flexible as per agreed formulae.
- Safer than the nuclear energy-Worldwide perception is that nuclear energy contains enormous risk that is unacceptable to most of the industrialized nations.
- Growth rate of energy consumption world wide-The World primary energy (comprising commercially traded fuels) consumption during the year 2005 was registered at 10537.1

MTOE with a growth rate of 2.7% over the previous year. Similarly during the year 2006, energy growth has been 2.4%.Same trend continued.

- Serving large number of customers connected to the gas supply grid

1.5) LNG Value Chain:

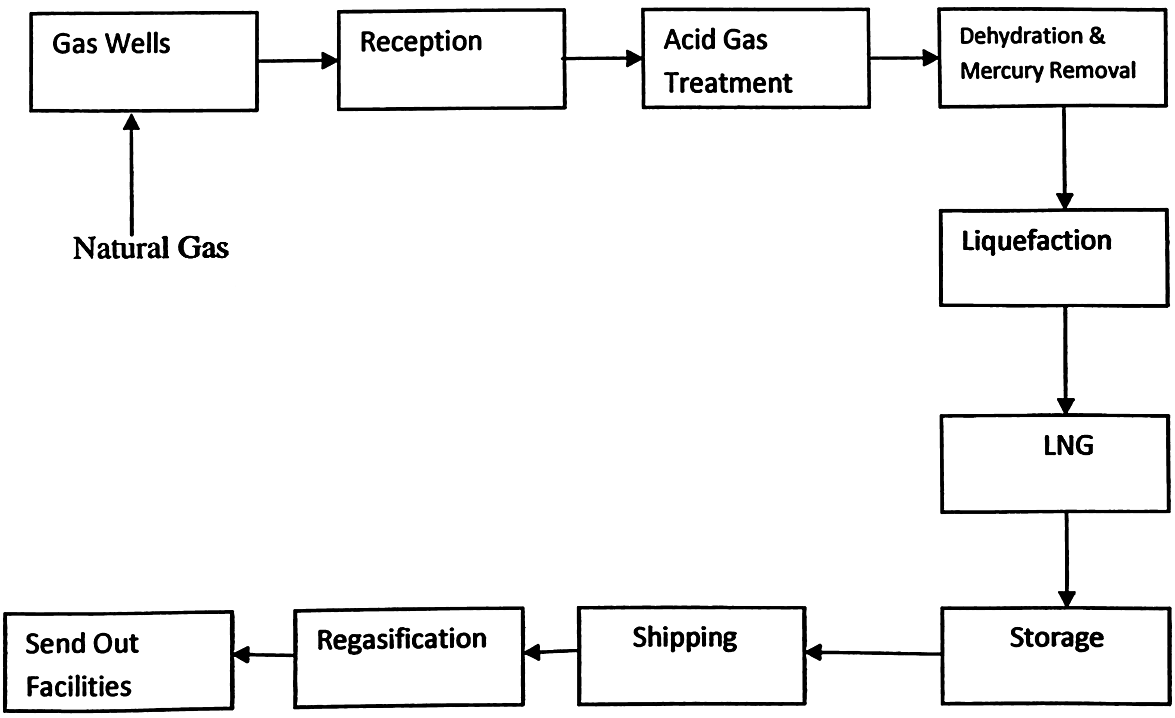

Figure1.2 LNG Value Chain Block Flow Diagram

- **Exploration and Production:** To find natural gas and start production of the gas for delivery to gas users. In the present case the LNG liquefaction plant. For a normal LNG plant (5MMPTPA capacity), the proven reserves of gas shall have be 5-8 TCF before LNG liquefaction project is initiated.

- **Gas Processing:**In the context of discussions on LNG value chain, the gas produced from the field is entirely dedicated to LNG liquefaction facility. The normal gas treatment activities such as dehydration, gas sweetening (removal of sour gas), CO_2 removal, mercury removal (if necessary), and higher hydrocarbon removal have to be suitably located between the gas producing facilities and the LNG liquefaction plant. The economic consideration coupled with system safety, the following will be a desirable combination.

Table 1.3 Various gas processing operations and their preferred locations.

S.NO.	Process	Location
1	Dehydration	Near gas producing field.
2	Water treatment & disposal	Near gas producing field.
3	Gas sweetening	Preferably at liquefaction plant.
4	CO_2 removal	Preferably at liquefaction plant.
5	Mercury etc. removal	Preferably at liquefaction plant.
6	Higher hydrocarbon removal and condensate treatment	Preferably at liquefaction plant.

The feed gas is delivered at high pressure from upstream gas fields via trunk lines and various unit operations carried out in field processing. The first unit operation is the physical separation of the distinct phases, which are gas and possibly liquid hydrocarbon, liquid water, and solids. The temperature and pressure of the gas stream dictates whether liquid hydrocarbon and/or water are present. Phase separation usually occurs in a pressure vessel provided for that purpose. Gas leaves from the top of the vessel and liquid from the bottom. If a three-phase separator is needed, special internal element are required to permit separation of the water and hydrocarbon liquids. Solids fall to the bottom and must be removed by special techniques. Mist extractors-wire mesh, vanes, or baffles are sometimes used to reduce liquid entrainment in the gas stream.

The next in processing is treating, if necessary, to remove the acid-gas components hydrogen sulfide and carbon dioxide; H_2S is toxic and both are corrosive. Hydrogen sulfide removal must be essentially complete, while the extent of carbon dioxide removal depends on the intended use for the gas. If the gas is to be cooled to cryogenic temperatures (less than, say, -100°F, CO_2 removal to a few tenths of a percent may be required to prevent formation of solid CO_2.

Dehydration is necessary to prevent formation of gas hydrates, which may plug processing equipment or pipelines at high pressure, even at temperatures considerably higher than 32°F (0°C). The two principal dehydration methods are glycol contacting and solid-desiccant adsorption. Methanol or glycol also can be injected to prevent hydrate formation. Gas containing considerable amount of liquefiable HC (ethane, propane and heavier) produces condensate upon cooling or compressing and cooling. In some cases the potential NGL are sufficiently valuable. Normally, condensate products in a central facility rather than in the field. Recovered condensate may have to be stabilized by partial removal of dissolved gaseous components to obtain a liquid product.

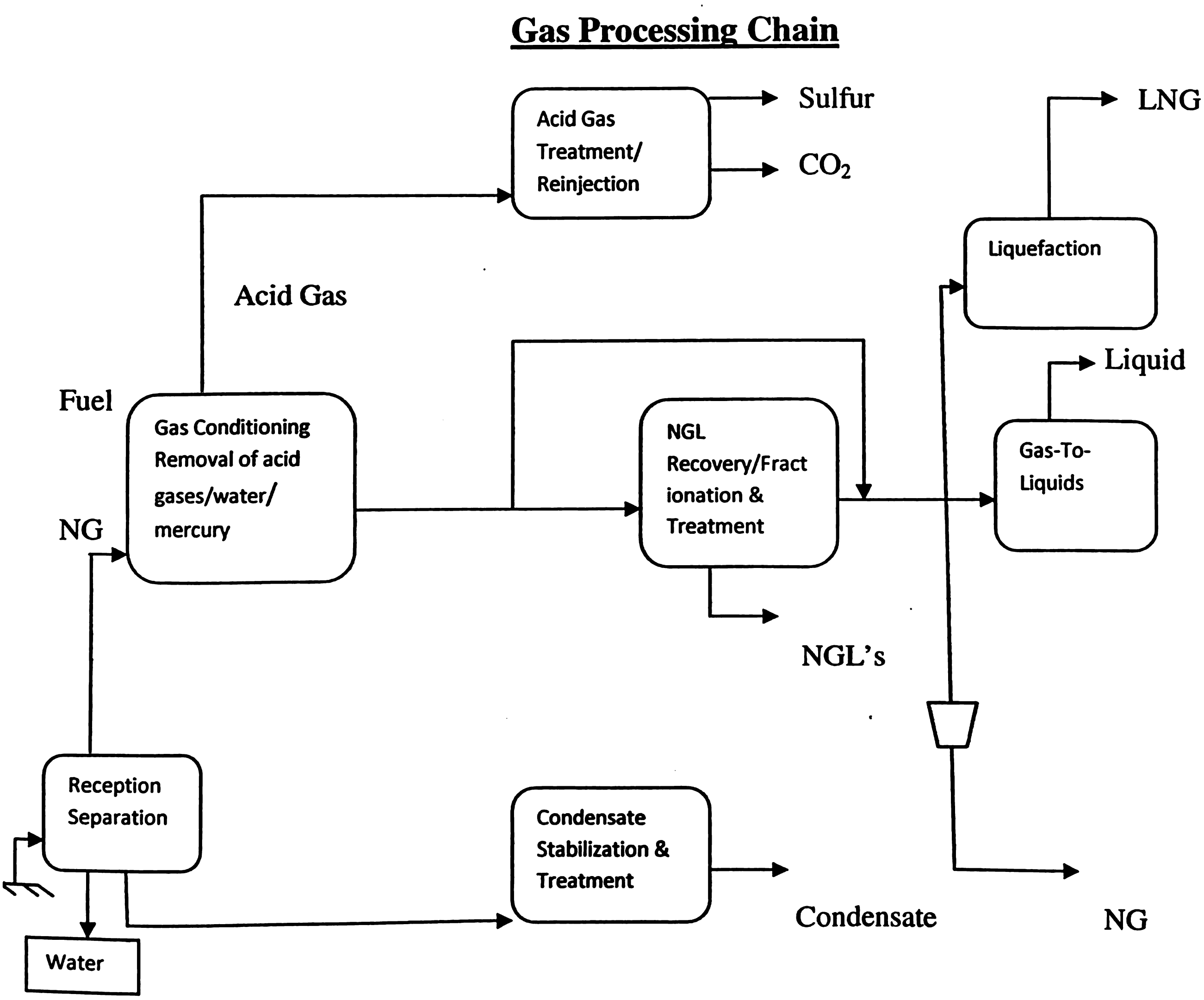

Figure 1.3 Gas processing chain

An example of a LNG plants overall flow scheme, and the main process units and supporting utilities, is shown in the figure. Refrigerant streams to separate heavier hydrocarbons then cool the dry sweet gas. The remaining gas is made up mainly of methane and contains less than 0.1 mol% of pentane and heavier hydrocarbons.

Allowable limit of impurities in LNG feed gas is given below:

Table 1.4 Allowable limits of impurities in LNG.

Impurity	Allowable limit	Limiting criterion
Carbon dioxide	< 50 ppm (vol)	Solubility
Water	< -73°C dew point	Solubility
Hydrogen Sulphide	< 5 ppm (max)	Product specifications
Mercury	< 5 mg (SM^3)	Corrosion
Aromatics	< 2 ppm (vol / max)	Solubility

- **Liquefaction:** At the production site, natural gas is converted into a liquid form through a cooling process called liquefaction, achieved through several refrigeration cycles. The set of units where natural gas is purified and liquefied is called a train. The largest liquefaction train in operation has a production capacity of 5 million tonnes per year and trains designed for 7.8 million tonnes per year are under construction in Qatar. This production capacity is expected to increase to 197 million tonnes in 2007 based on the facilities under construction.

There are 3 main liquefaction processes:

- The classical cascade, where refrigeration and liquefaction of the gas is achieved in a cascade process using three pure refrigerants: propane, ethylene and methane.
- The single flow mixed refrigerant process, where the mixed refrigerant made up of nitrogen, methane, ethane, propane and isopentane, is compressed and circulates using a single compression train.
- The propane precooled mixed refrigerant process where precooling is achieved by a multi-stage propane cycle and liquefaction and sub cooling are accomplished by a two-stage mixed refrigerant cycle.

Other processes have been developed but are not yet in operation:

- The cascade mixed refrigerant cycle, where three mixed refrigerant cycles are used for precooling, liquefaction, sub cooling.

- The dual mixed refrigerant process, where both the precooling and the liquefaction cycles use mixed refrigerants.

➢ The AP-X process, based on the propane precooled mixed refrigerant process with a separate nitrogen cycle for sub cooling.

- **LNG Storage:** After production of LNG in liquefaction plant, the same is stored in 9% Ni tanks specially fabricated with proper insulation between inner shell of 9% Ni and outer shell of Carbon steel or 9% Ni depending on the design. The storage tanks are cylindrical vessel constructed either above or underground with proper safety as per the available code in practice.

  LNG Storage facilities are also required at the receiving terminal where regasification of LNG takes place in a subsequent plant.

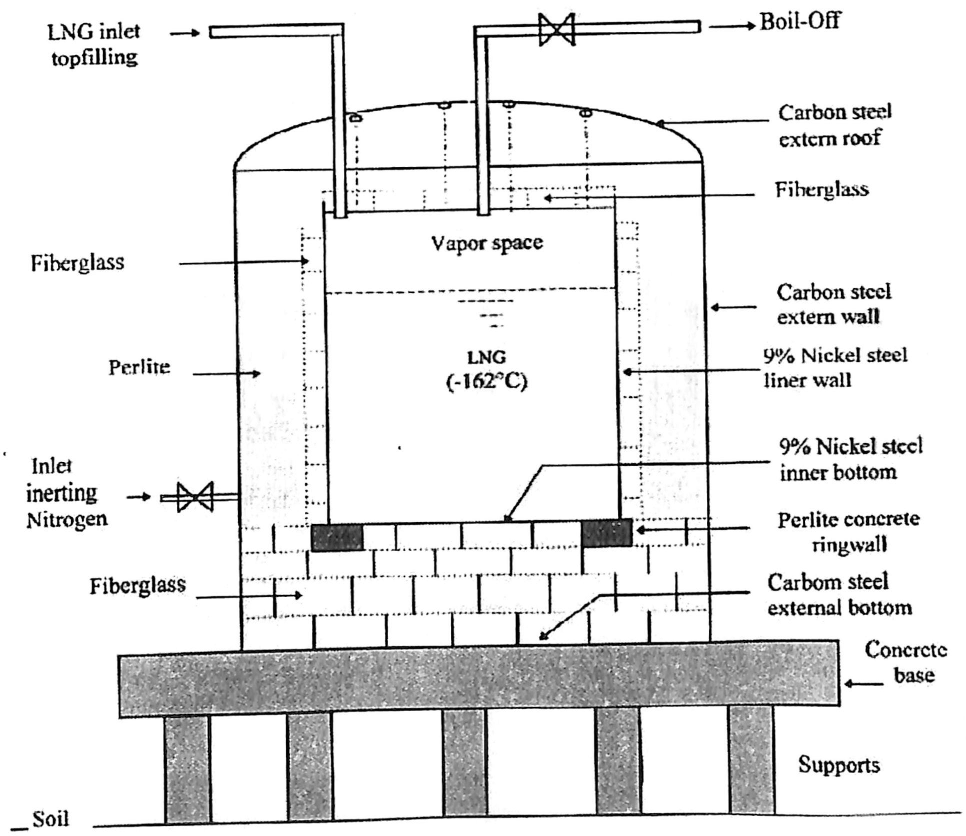

Figure 1.4 LNG Storage Tank

The main tank types are

1. Single containment.
2. Double containment.
3. Full containment.

- **LNG Shipping:** LNG is transported in double-hulled ships, specifically designed to handle the low temperature of LNG, to a receiving terminal where it is stored and regasified. These carriers are insulated to limit the amount of LNG that boils off or evaporates.
- **Re-gasification Terminal:** To convert the LNG stored in specially built storage tanks, from liquid phase to gaseous phase for further transportation through natural gas pipeline system or for captive use in associated & integrated facilities like power generation etc. the LNG is vaporized by taking away the cold of LNG either by sea water or by blowing air or by heating .

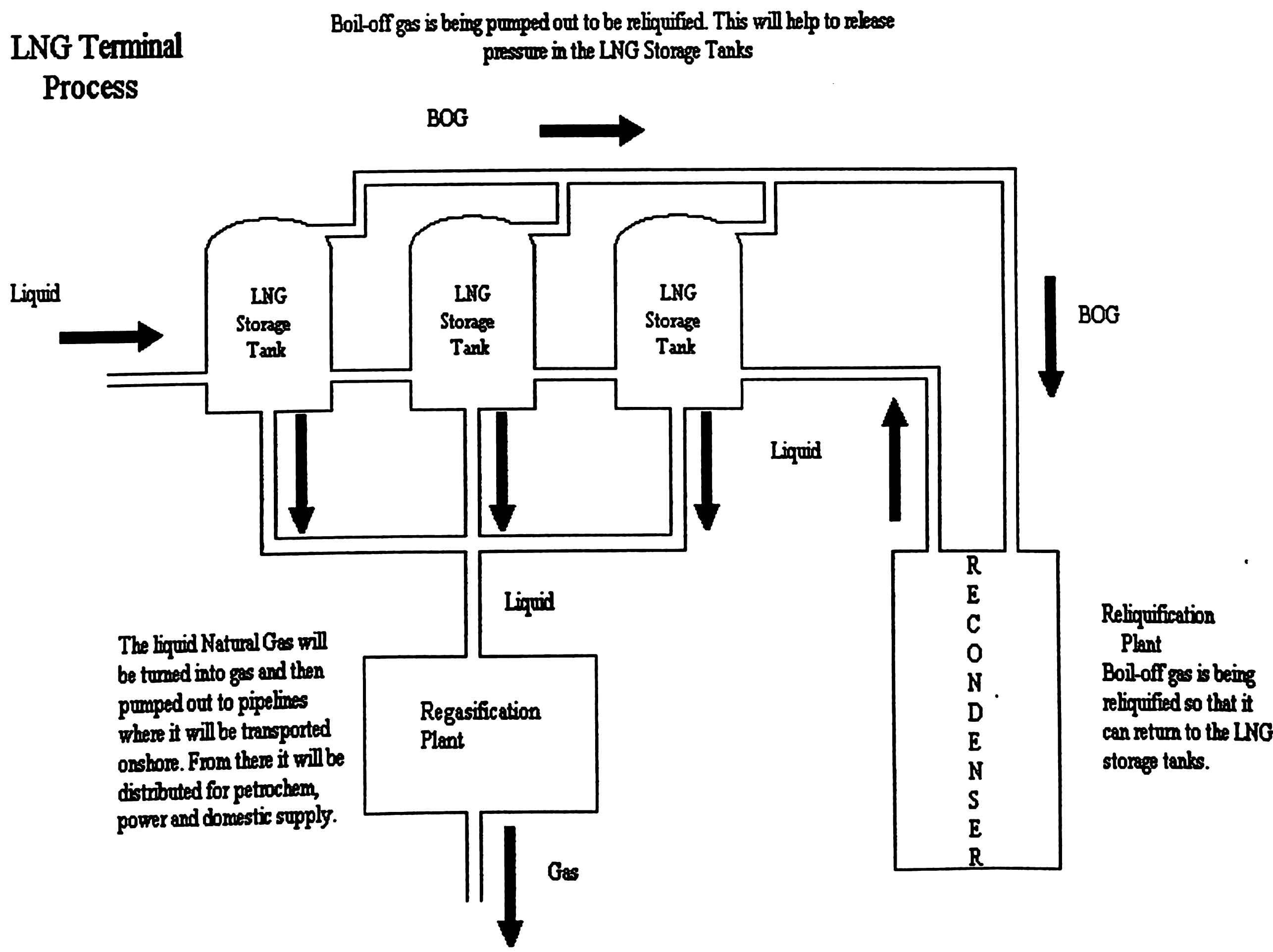

Figure 1.5: General view of LNG re-gasification terminal.

Certain marine facilities to be included at the terminal are:

Table 1.5 Various facilities at LNG re-gasification terminal.

S.NO.	Facility	Use and Dimension
1	Shipping Channel	A 15 meter depth of water in the shipping channel required to allow in the LNG tankers.
2	Turning Basin	A basin of 15 meter water depth required to provide a turning circle for an LNG tanker (up to 950 meters diameter)
3	Jetty	If the jetty is unprotected a breakwater is required to provide protection from waves. There may be tradeoff between the jetty length and the dredging, required to provide a channel. The requirements for a breakwater depend on wave conditions, wave heights and the 100 year wave size.
4	Harbor	The key issues relating to the harbor depend on whether it is dedicated to LNG use or is established remote. LNG ships require immediate access on arrival and isolation from other shipping and cargo handling.
5	Terminal Site	The size required depends on the quantity of storage required and the need for back-up liquid fuel storage. The site should be seismically and geodetically secure.
9	Break water	Breakwaters are structures constructed on coasts as part of coastal defense or to protect an anchorage from the effects of weather and long shore drift.
10	Tug Boats:	The deployment of the tugboats should be such that they can independently handle ship movement in the channel under all adverse sea conditions involving high tides, strong waves and severe wind conditions.
11	Navigational Aids	To identify the approach channel, adequate number of buoys to be provided to assist navigation of the LNG ship.

- **LNG send out facilities and pipeline to customers:** The vaporized LNG is measured before it is supplied to a system of pipeline connecting to various customers who have contracted LNG.
  The innovation for transporting LNG directly in road tankers has also generated good market for LNG.

CHAPTER 2

Liquefaction Processes and Refrigerant Properties

2.1) APCI Propane Pre-cooled Mixed Refrigerant Process:

This process accounts for a very significant proportion of the world's baseload LNG production capacity. Train capacities of up to 4.7 million tpy were built or are under construction. The Figure illustrates part of an overall LNG plant flow scheme. There are two main refrigerant cycles. The precooling cycle uses a pure component, propane. The liquefaction and sub-cooling cycle uses a mixed refrigerant (MR) made up of nitrogen, methane, ethane and propane. The precooling cycle uses propane at three or four pressure levels and can cool the process gas down to -40 °C. It is also used to cool and partially liquefy the MR. The cooling is achieved in kettle-type exchangers with propane refrigerant boiling and evaporating in a pool on the shell side, and with the process streams flowing in immersed tube passes. A centrifugal compressor with side streams recovers the evaporated C_3 streams and compresses the vapor to 15 - 25 bar to be condensed against water or air and recycled to the propane kettles. In the MR cycle, the partially liquefied refrigerant is separated into vapor and liquid streams that are used to liquefy and sub-cool the process stream from typically -35 °C to between -150 °C to -160 °C. This is carried out in a proprietary spiral wound exchanger, the main cryogenic heat exchanger (MCHE). The MCHE consists of two or three tube bundles arranged in a vertical shell, with the process gas and refrigerants entering the tubes at the bottom, which then flow upward under pressure. The process gas passes through all the bundles to emerge liquefied at the top. The liquid MR stream is extracted after the warm or middle bundle and is flashed across a Joule Thomson valve or hydraulic expander onto the shell side. It flows downwards and evaporates, providing the bulk of cooling for the lower bundles. The vapor MR stream passes to the top (cold bundle) and is liquefied and sub-cooled, and is flashed across a JT valve into the shell side over the top of the cold bundle. It flows downwards to provide the cooling duty for the top bundle and, after mixing with liquid MR, part of the duty for the lower bundles.

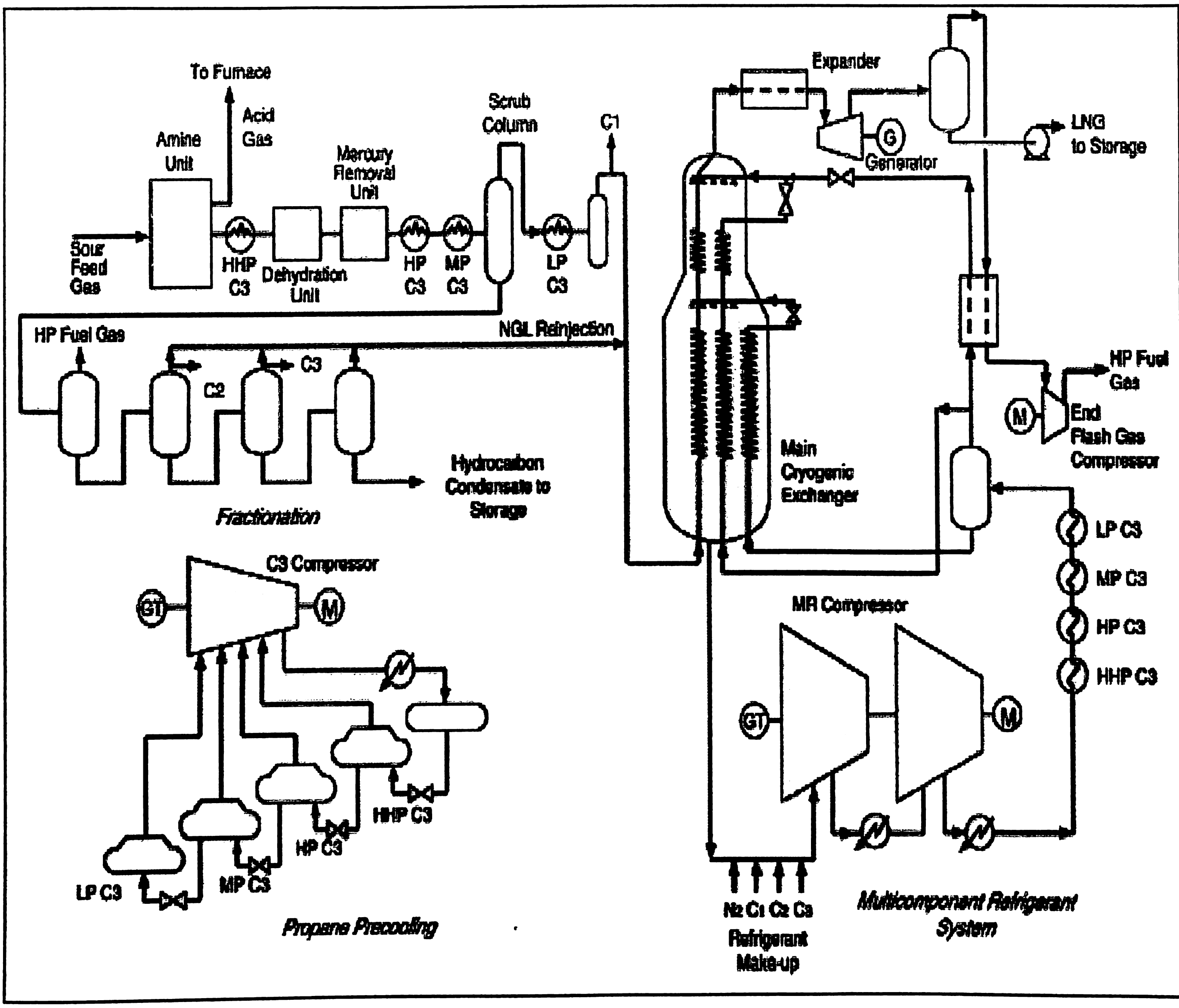

Figure 2.1 APCI propane pre-cooled mixed refrigerant process

Source LNG Technology Selection: Dr.Tariq Shukri

The overall vaporized MR stream from the bottom of the MCHE is recovered and compressed by the MR compressor to 45 - 48 bara. It is cooled and partially liquefied first by water or air and then by the propane refrigerant, and recycled to the MCHE. In earlier plants all stages of the MR compression were normally centrifugal, however, in some recent plants axial compressors have been used for the LP stage and centrifugal for the HP stage. Recent plants use Frame 6 and/or Frame 7 gas turbine drivers. Earlier plants used steam turbine drivers. A recent modification of the process, which is being considered for large LNG capacity plants (> 6 million tpy), is the APX-process, which adds a third refrigerant cycle (nitrogen expander) to conduct LNG sub cooling duties outside the MCHE.

2.2) Phillips Cascade LNG Process:

In this process the treated gas is fed to the liquefaction unit where it is cooled and sub-cooled prior to entering the LNG tanks. The liquefaction system utilizes the Phillips Optimized Cascade LNG Process, a modification of the original Phillips LNG plant design at Kenai, Alaska. This process uses two pure refrigerants – propane and ethylene circuits and methane flash circuit cascaded to provide maximum LNG production by utilizing the horsepower available from 6 Frame 5D gas turbines. Each circuit uses two 50% compressors with common process equipment. Brazed Aluminum Heat Exchangers and Core-in-Kettle Exchangers are used for the feed gas, propane, ethylene and methane circuits. All of these heat exchangers with the exception of the propane chillers are housed in two "Cold Boxes". All compressor inter-cooling, after-cooling and propane refrigerant condensing is provided. The LNG from the last stage flash drum is sent to the LNG tanks by the LNG transfer pumps where it is stored at approximately –161°C.

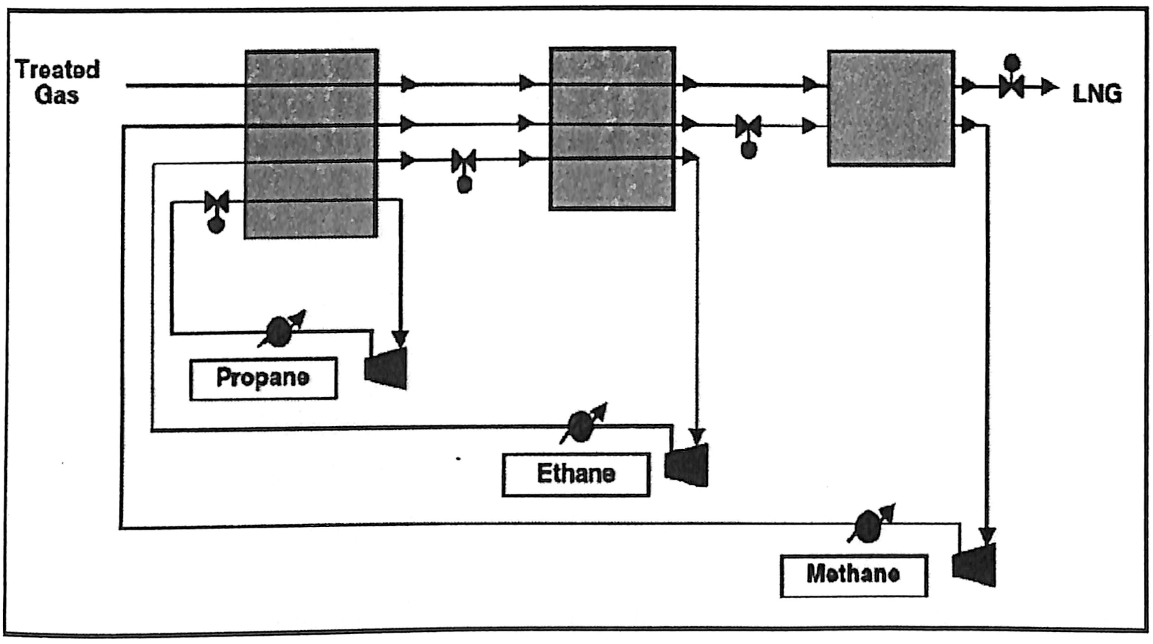

Figure 2.2: Phillips Cascade LNG Process

Source:LNG Technology Selection: Dr.Tariq Shukri

2.3) Statoil/Linde mixed fluid cascade process (MFCP):

In this process three mixed refrigerants are used to provide the cooling and liquefaction duty. The process is illustrated in Figure. Pre-cooling is carried out in PFHE by the first mixed refrigerant, and the liquefaction and subcooling are carried out in a spiral wound heat exchanger (SWHE) by the other two refrigerants. The SWHE is a proprietary exchanger made by Linde. It may also be used for the pre-cooling stage. The refrigerants are made up of components selected from methane, ethane, propane and nitrogen. The three refrigerant compression systems can have separate drivers or integrated to have two strings of compression. Frame 6 and Frame 7 gas turbine drivers have been proposed for large LNG trains (> 4 million tpy).

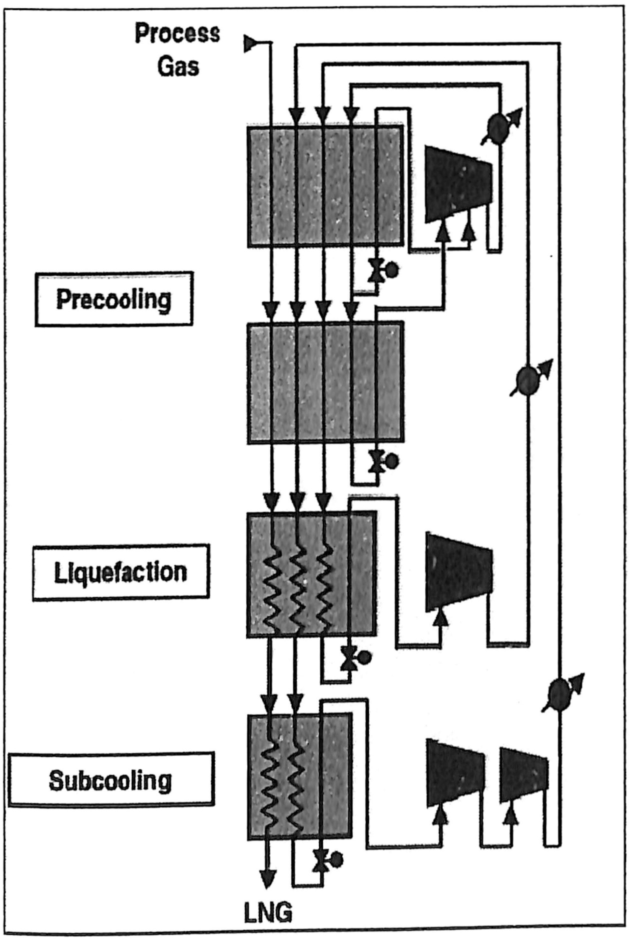

Figure 2.3: Statoil/Linde mixed fluid cascade process (MFCP)

Source LNG Technology Selection: Dr.Tariq Shukri

2.4) Axens Liquefin process:

This is a two-mixed refrigerant process, which is being proposed for some new LNG base load projects of train sizes up to 6 million tpy. It is illustrated in Figure. All cooling and liquefaction is conducted in PFHE arranged in cold boxes. The refrigerants are made up of components from methane, ethane, propane, butane and nitrogen. The first mixed refrigerant is used at three different pressure levels to precool the process gas and precool and liquefy the second mixed refrigerant. The second mixed refrigerant is used to liquefy and subcool the process gas. Using a mixed refrigerant for the precooling stage allows a lower temperature to be achieved (for example, -60 °C) depending on refrigerant composition. Two large drivers can drive the refrigerant compression systems. Frame 7 gas turbines are being proposed for the large LNG trains.

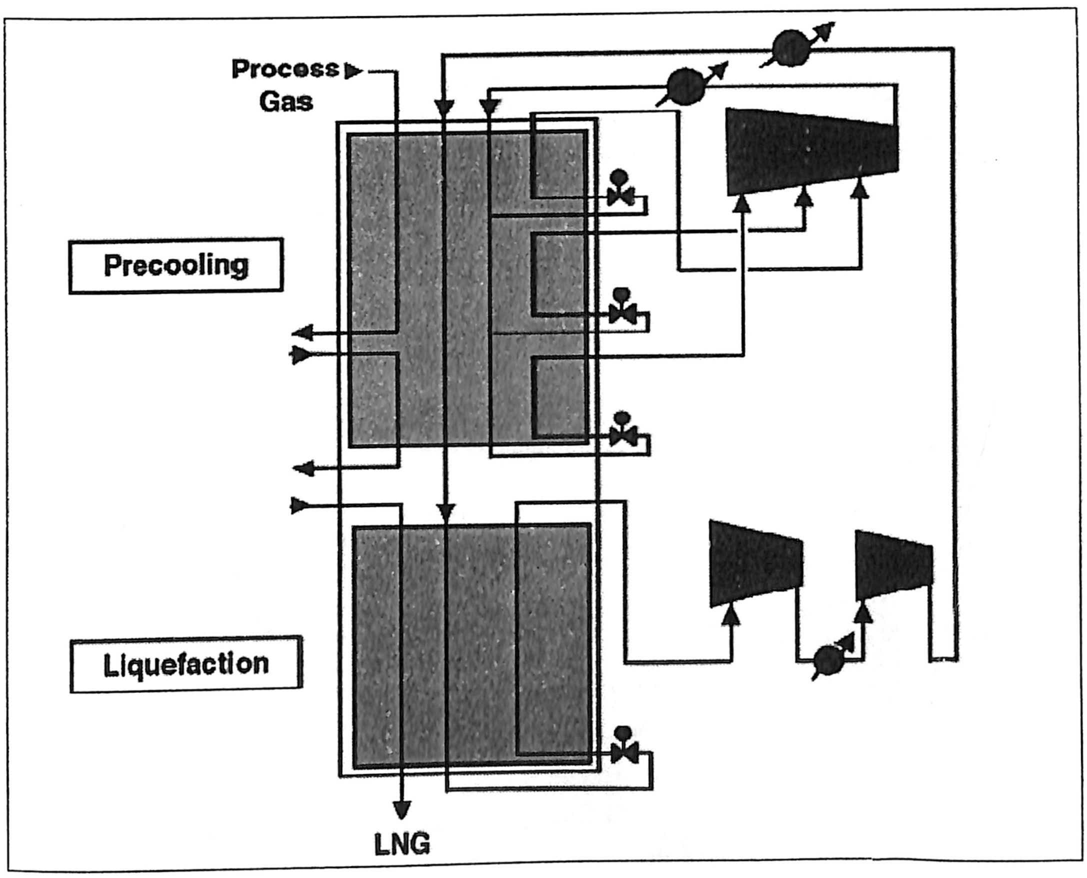

Figure 2.4 Axens Liquefin process

Source LNG Technology Selection: Dr.Tariq Shukri

2.5) Shell double mixed refrigerant process (DMR):

This is a dual mixed refrigerant process, which is being applied in the Sakhalin Island project with a capacity of 4.8 million tpy per train. Process configuration is similar to the propane pre-cooled mixed refrigerant process, with the precooling conducted by a mixed refrigerant (made up mainly of ethane and propane) rather than pure propane. Another main difference is that the precooling is carried out in SWHEs rather than kettles. Linde will supply the precooling and liquefaction SWHEs. The refrigerant compressors are driven by two Frame 7 gas turbines. An axial compressor is also used as part of the cold refrigerant compression stages.

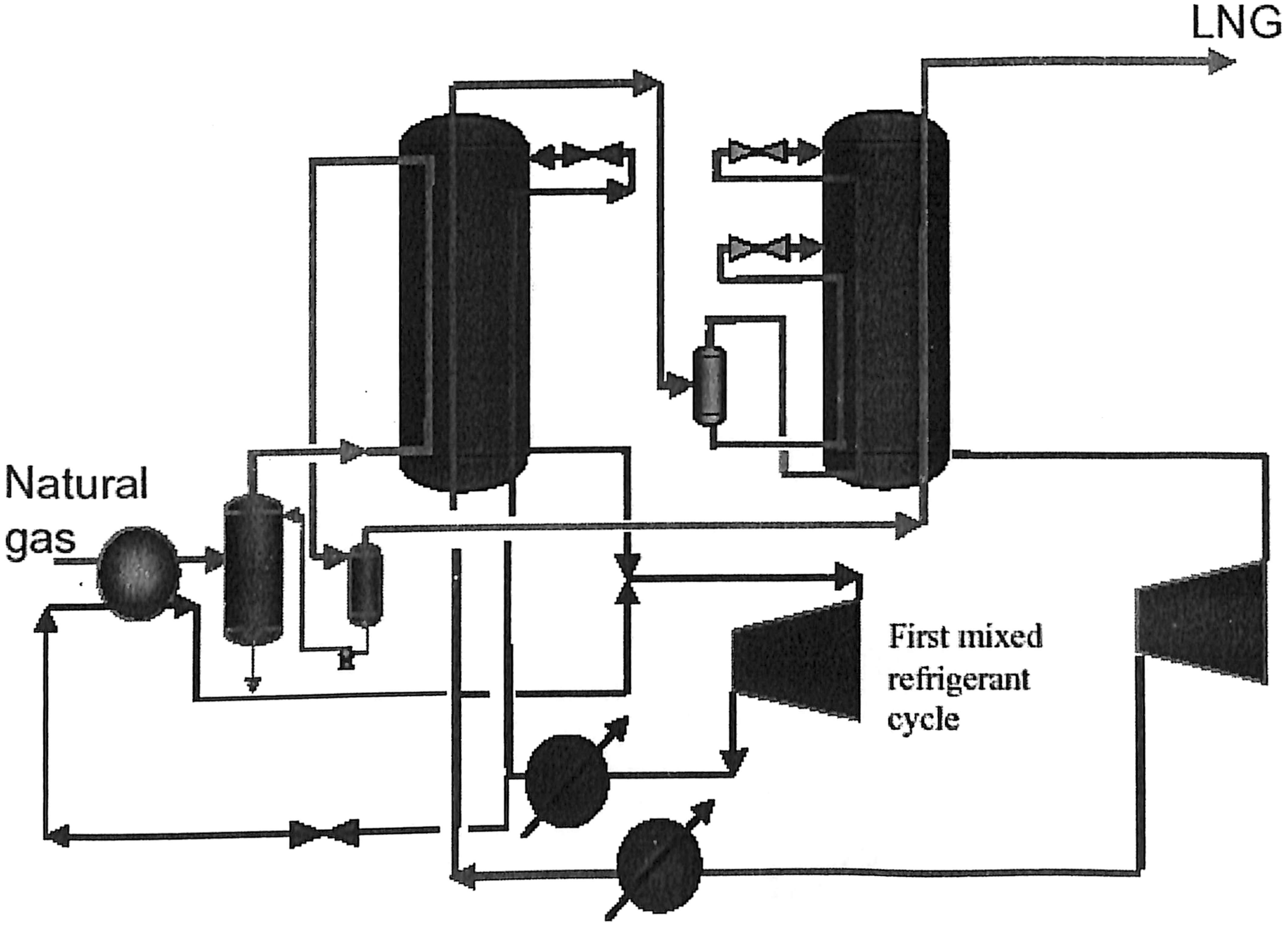

Figure 2.5 Shell double mixed refrigerant process

Source Natural gas liquefaction processes comparison, Poster PO-39

2.6) Modified processes to reduce costs and to increase capacity through effective techniques:

2.6.1) Phillips Optimized Cascade LNG Process: One change that has been incorporated in the LNG process over the years is the modification of the methane refrigerant circuit such that it is now an open circuit or a feed-flash system rather than the original closed circuit as used at Kenai. The major advantage of this modification is that it eliminates the separate fuel gas compressor. Also storage vapors and vapors from tanker loading are recovered and fed back to the liquefaction train and reliquefied rather than being routed directly to fuel or flare, thereby increasing LNG production. Other modifications to the basic Kenai plant configuration that could be considered are the replacement of the dual gas turbines/compressors used on each refrigerant, as at Kenai, with a single gas turbine/compressor on each refrigerant, particularly if two or more liquefaction trains are to be installed at the site. Also, if the feed gas has significant amounts of C_2+ hydrocarbons a hydrocarbon liquid draw system would be included.

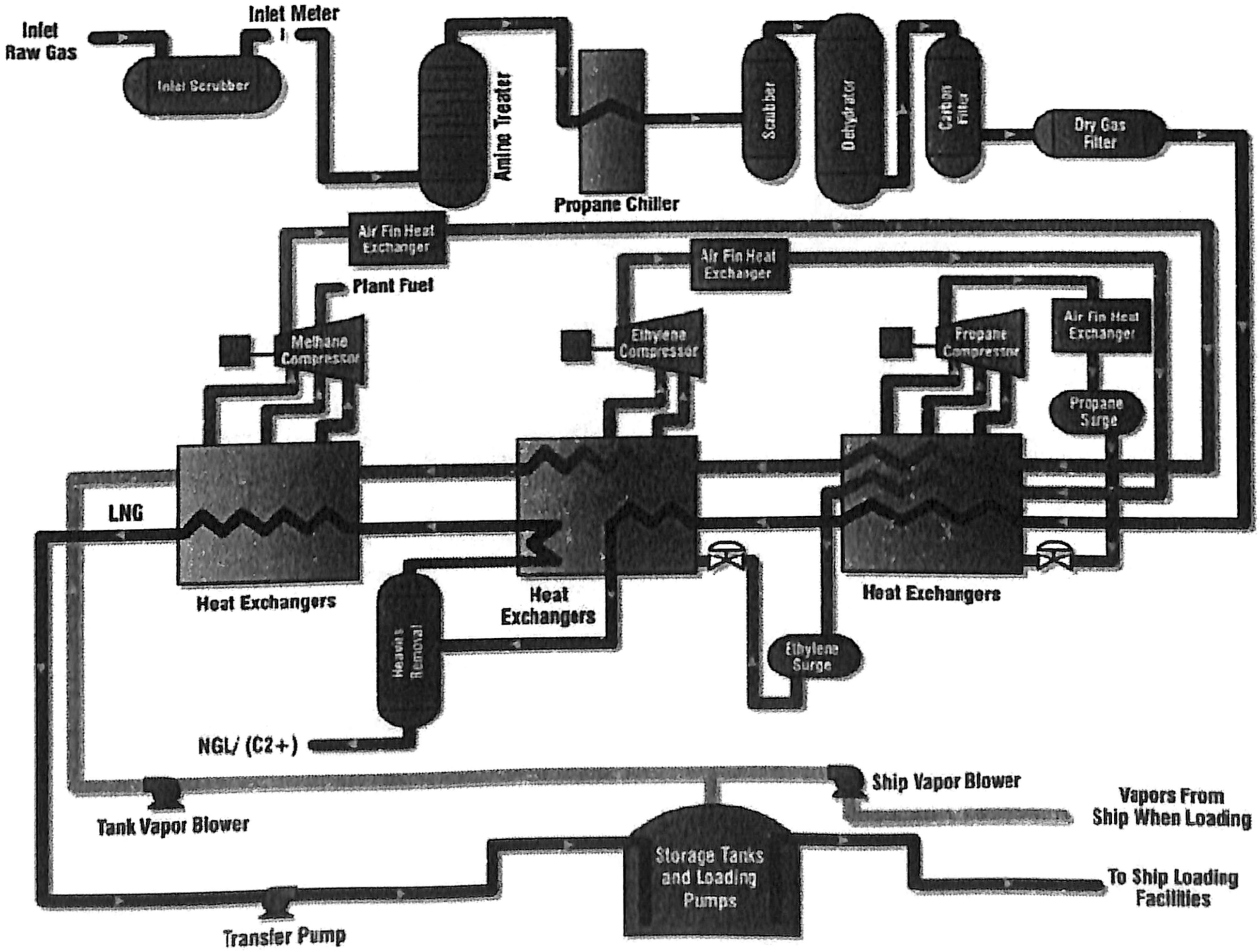

Figure 2.6: Phillips Optimized Cascade LNG Process

There are still other modifications that can be incorporated in the Phillips Optimized Cascade LNG Process that would increase its operational effectiveness in LNG production.

So, two compressors in a cycle increase the available capacity, if any one of the compressors trips down the operation does not stop.

Benefits:

1) Vapors from the ship loading also liquefied, and the production of the plant can be increase.
2) Decrease in vapor loss.
3) Capacity of the compressors available will be more.

2.6.2) APCI APX (expandable) process:

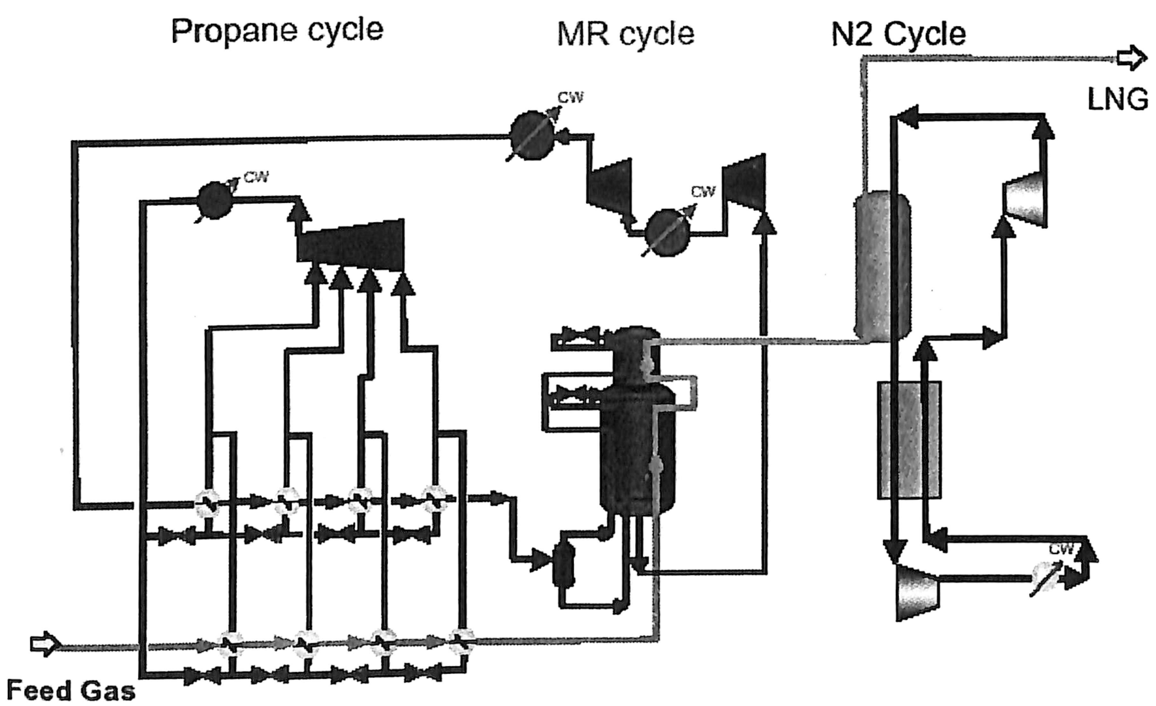

Figure 2.7: APCI APX (expandable) processes

Source Natural gas liquefaction processes comparison, Poster PO-39

Modification to this process is that, it adds a nitrogen cycle to the C_3MR process, which liquefy and sub cools the LNG. The exchangers used are kettles for the propane, spiral wound for the mixed refrigerant, another spiral wound and plate-fin exchanger for the nitrogen cycle. The process is going to have electric motors, Frame 9 or Frame 7 or combination of both. The efficiency of this process is high, hence the cost of the liquefier per unit of LNG

Benefits:

1) Nitrogen cycle will increase the capacity of plant, which will liquefy and subcool the extra amount of the gas.
2) The plate fin heat exchanger gives more efficient processing, having large surface area per unit volume.

2.6.3) Shell (PMR) Parallel Mixed Process:

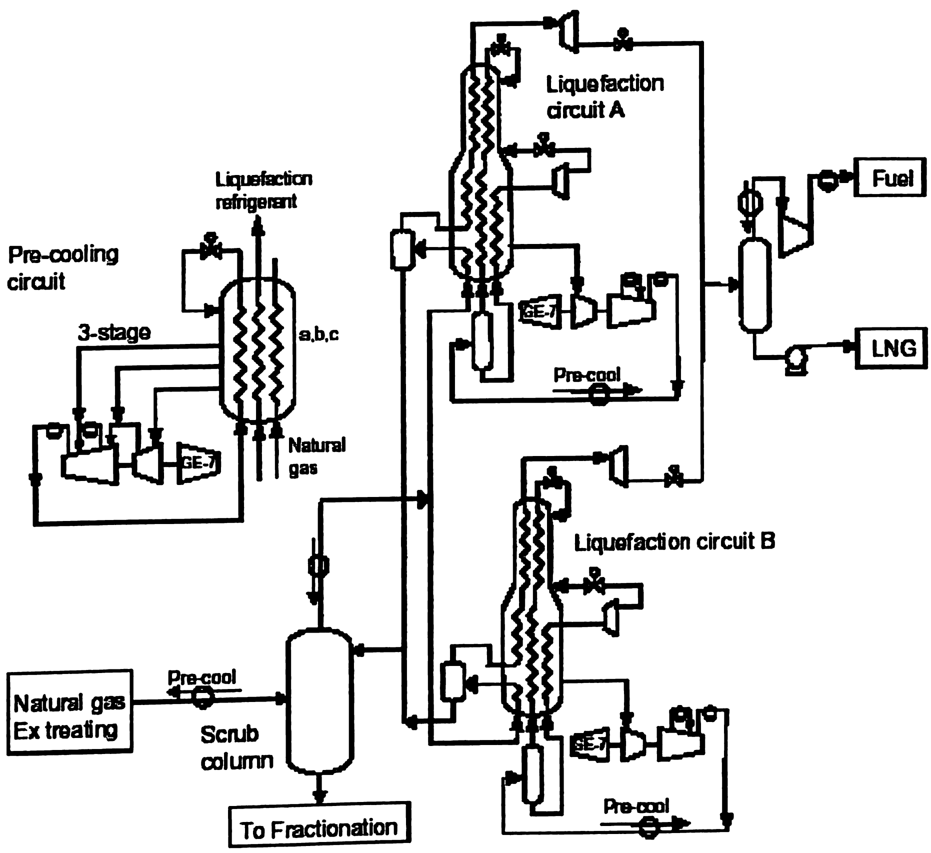

Figure 2.8 Shell Parallel Mixed Process (PMR)
Source LARGE CAPACITY LNG PLANT DEVELOPMENT- Paper PS5-3

The PMR process is basically derived in the DMR process lineup. This process will use three gas turbines. In one circuit precooling will be done, and then the string will split in two parallel circuits, each having a scrub column for NGL extraction and an MR cycle for liquefaction and sub-cooling of the natural gas. After this both the streams meet at end flash system (not shown) where the fuel gas is taken out and LNG is drawn. The GE-frame 7 turbines can be used which can drive large compressors. Production can be achieved up to 8 MMTPA.

Benefits:

1) Due to parallel flow efficiency will increase.
2) Even if anyone of the process circuit trips in parallel, other stream will be continues to produce LNG.

2.7) Properties of Refrigerants:

2.7.1) Propane Gas Properties

Fomula: C_3H_8

Molecular Weight

- Molecular weight : 44.096 g/mol

Solid phase

- Melting point : -187.7 °C
- Latent heat of fusion (1,013 bar, at triple point) : 94.98 kJ/kg

Liquid phase

- Liquid density (1.013 bar at boiling point) : 582 kg/m3
- Liquid/gas equivalent (1.013 bar and 15 °C (59 °F)) : 311 vol/vol
- Boiling point (1.013 bar) : -42.1 °C
- Latent heat of vaporization (1.013 bar at boiling point) : 425.31 kJ/kg
- Vapor pressure (at 21 °C or 70 °F) : 8.7 bar

Critical point

- Critical temperature : 96.6 °C
- Critical pressure : 42.5 bar

Gaseous phase

- Gas density (1.013 bar at boiling point) : 2.423 kg/m3
- Gas density (1.013 bar and 15 °C (59 °F)) : 1.91 kg/m3

- Compressibility Factor (Z) (1.013 bar and 15 °C (59 °F)) : 1.0193
- Specific gravity (air = 1) (1.013 bar and 21 °C (70 °F)) : 1.55
- Specific volume (1.013 bar and 21 °C (70 °F)) : 0.543 m3/kg
- Heat capacity at constant pressure (Cp) (1 bar and 25 °C (77 °F)) : 0.075 kJ/(mol.K)
- Heat capacity at constant volume (Cv) (1 bar and 25 °C (77 °F)) : 0.066 kJ/(mol.K)
- Ratio of specific heats (Gamma:Cp/Cv) (1 bar and 25 °C (77 °F)) : 1.134441
- Thermal conductivity (1.013 bar and 0°C (32 °F)) : 15.198 mW/(m.K)

Miscellaneous

- Solubility in water (1.013 bar and 20 °C (68 °F)) : 0.039 vol/vol
- Auto ignition temperature : 470 °C

2.7.2) Nitrogen Gas Properties

Formula: N_2

Molecular Weight

- Molecular weight : 28.0134 g/mol

Solid phase

- Melting point : -210 °C
- Latent heat of fusion (1,013 bar, at triple point) : 25.73 kJ/kg

Liquid phase

- Liquid density (1.013 bar at boiling point) : 808.607 kg/m3
- Liquid/gas equivalent (1.013 bar and 15 °C (59 °F)) : 691 vol/vol
- Boiling point (1.013 bar) : -195.9 °C
- Latent heat of vaporization (1.013 bar at boiling point) : 198.38 kJ/kg

Critical point

- Critical temperature : -147 °C
- Critical pressure : 33.999 bar
- Critical density : 314.03 kg/m3

Triple point

- Triple point temperature : -210.1 °C
- Triple point pressure : 0.1253 bar

Gaseous phase

- Gas density (1.013 bar at boiling point) : 4.614 kg/m3
- Gas density (1.013 bar and 15 °C (59 °F)) : 1.185 kg/m3
- Compressibility Factor (Z) (1.013 bar and 15 °C (59 °F)) : 0.9997
- Specific gravity (air = 1) (1.013 bar and 21 °C (70 °F)) : 0.967
- Specific volume (1.013 bar and 21 °C (70 °F)) : 0.862 m3/kg
- Heat capacity at constant pressure (Cp) (1.013 bar and 25 °C (77 °F)) : 0.029 kJ/(mol.K)
- Heat capacity at constant volume (Cv) (1.013 bar and 25 °C (77 °F)) : 0.02 kJ/(mol.K)
- Ratio of specific heats (Gamma:Cp/Cv) (1.013 bar and 25 °C (77 °F)) : 1.403846
- Viscosity (1.013 bar and 0 °C (32 °F)) : 0.0001657 Poise
- Thermal conductivity (1.013 bar and 0 °C (32 °F)) : 24 mW/(m.K)

Miscellaneous

- Solubility in water (1.013 bar and 0 °C (32 °F)) : 0.0234 vol/vol
- Concentration in air : 78.08 vol %

Chapter 3

Liquefaction Plant & Designing of Heat Exchanger

3.1) Liquefaction Plant:

Any liquefaction plant operates on refrigeration principle i.e. compression, condensation, expansion and evaporation.

The main components of the Liquefaction plant are mainly heat exchangers and compressors.

As shown in figure, two pre-cooling and liquefaction cycles have been taken to bring gas to -162^0 C. Propane is used for pre-cooling of the gas and nitrogen is used for liquefaction.

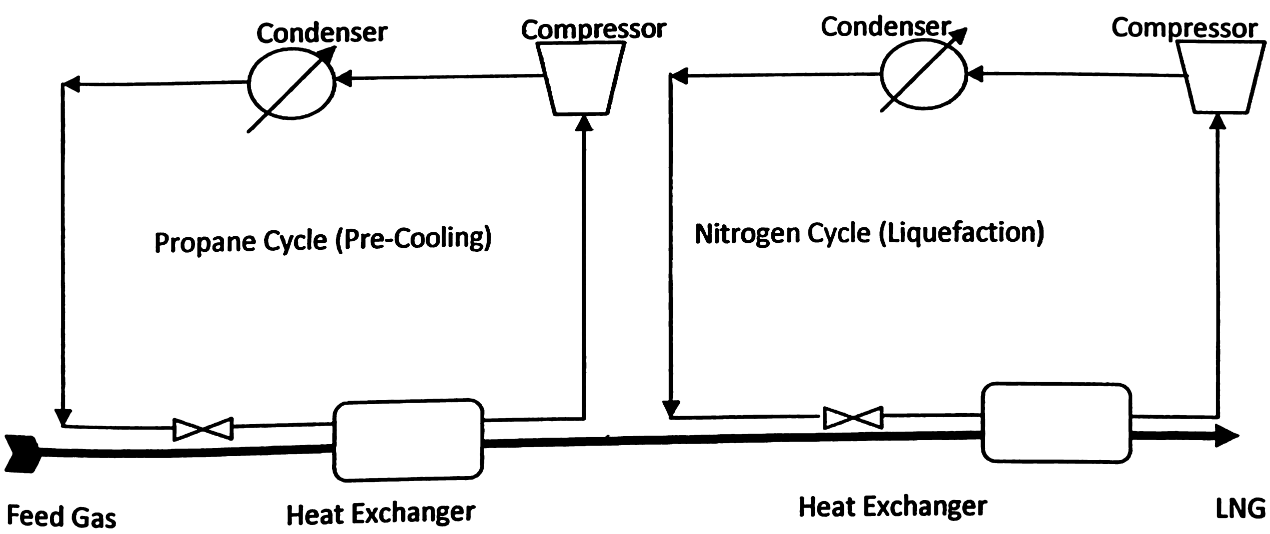

Figure 3.1: LNG liquefaction process

3.2) Components

3.2.1) Shell-and-Tube Heat Exchangers

Shell-and-tube exchangers are custom designed for virtually any capacity and operating condition, from high vacuums to ultrahigh pressures, from cryogenics to high temperatures, and for any temperature and pressure differences between the fluids, limited only by the materials of construction. They can be designed for special operating conditions: vibration, heavy fouling, highly viscous fluids, erosion, corrosion, toxicity, radioactivity, multicomponent mixtures, etc. They are made from a variety of metal and nonmetal materials, and in surface areas from less than 0.1 to 100,000 m^2 (1 to over 1,000,000 ft^2).They have generally an order of magnitude less surface area per unit volume than the compact exchangers, and require considerable space, weight, support structure, and footprint.

A shell-and-tube heat exchanger is essentially a bundle of tubes enclosed in a shell and so arranged that one fluid flows through the tubes and another fluid flows across the outside of the tubes, heat being transferred from one fluid to the other through the tube wall. A number of other mechanical components are required to guide the fluids into, through, and out of the exchanger, to prevent the fluids from mixing, and to ensure the mechanical integrity of the heat exchanger. A typical shell-and-tube heat exchanger is shown in Figure 3.2, but the basic design allows many modifications and special features, some of which are described below.

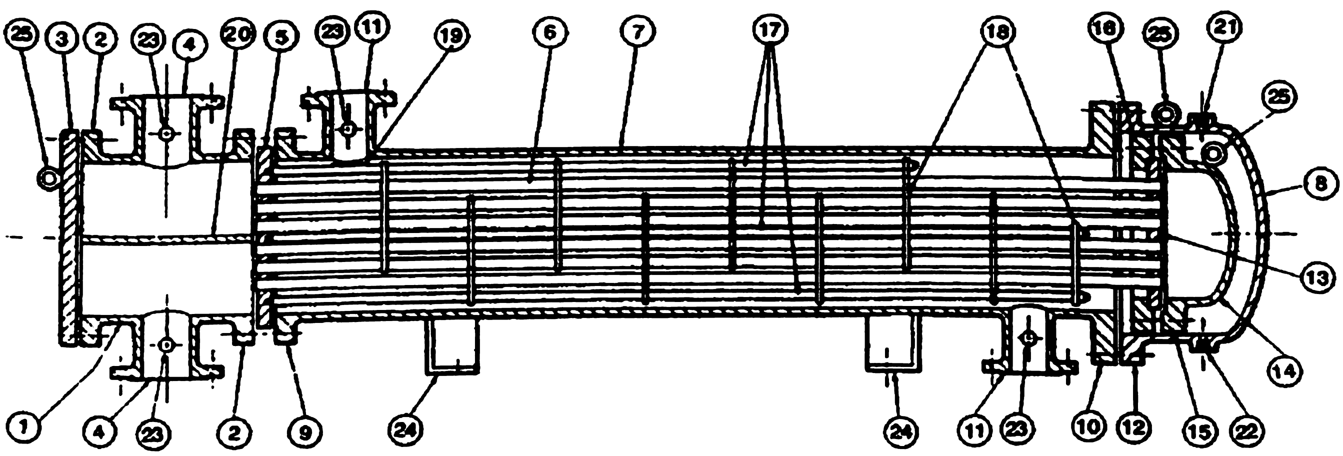

Figure 3.2: Longitudinal section of a typical shell-and-tube heat exchanger

1. Stationary Head-Channel
2. Stationary Head Flange-Channel or Bonnet
3. Channel Cover
4. Stationary Head Nozzle
5. Stationary Tube sheet

6. Tubes
7. Shell
8. Shell Cover
9. Shell Flange-Stationary Head End
10. Shell Flange-Rear Head End
11. Shell Nozzle
12. Shell Cover Flange
13. Floating Tube sheet
14. Floating Head Cover
15. Floating Head Cover Flange
16. Floating Head Backing Device
17. Tie rods and Spacers
18. Transverse Baffles or Support Plates
19. Impingement Plates
20. Pass Partition
21. Vent Connection
22. Drain Connection
23. Instrument Connection
24. Support Saddle
25. Lifting Lug

Shell-and-tube heat exchangers have been constructed with heat transfer areas from less than 0.1 m^2 (1 ft^2) to over 100,000 m^2 (1,000,000 ft^2), for pressures from deep vacuum to over 1000 bar (15,000psi), for temperatures from near 0 to over 1400 K (2000°F), and for all fluid services including single-phase heating and cooling and multiphase vaporization and condensation. The key to such flexibility is the wide range of materials of construction, forming and joining methods, and design features that can be built into these exchangers.

3.2.2) Compressors:

The selection of the compressor will depend upon the characteristics of the process, such as flowrate, composition of the refrigerant. Requiring proven compressors and taking and taking advantage of higher efficiencies of axial compressors in large refrigeration system are in demand. Axial compressors achieve typical polytropic efficincies of 86% compared with only 82% for centrifugal compressors. The axial compressor will be used on the coolest stage of the refrigeration. An axial compressor also benefits operations offering the very useful and cost effective possibility of varying inlet vane angles for control.

Table 3.1: Some technology selection parameters

Technology selection items	Pros	Cons
Spiral wound exchanger	Flexible operation	Proprietary/more expensive
PFHE	Competitive vendors available. Lower pressure drop and temperature differences	Require careful design to ensure good 2-phase flow distribution in multiple exchanger configurations
Axial compressors	High efficiency	Suitable only at high flow rates.
Large gas turbines	Proven, efficient and cost effective	Less reliable/strict maintenance cycle more complicated control/fixed speed
Large motor drivers	Efficient, flexible & more available	Untried in LNG at speeds needed/require large power plant.
Mixed refrigerant process	Simpler compression system. Adjusting composition allows process matching	More complex operation.
Pure component cascade process	Potential higher availability with parallel compression	More equipment and complicated compression system
Air cooling (compared to sea water cooling)	Lower cooling system CAPEX	Less efficient process/higher operating costs
Fluid medium heating (Compared to steam)	Eliminates the need for steam generation & water treatment	Higher reboiler costs

Larger train capacity	Lower specific costs (CAPEX per tonne LNG)	Some equipment/ processes may require further development

Source Hydrocarbon engineering -Feb 2004- Dr. Tariq Shukri

3.3) Factors influencing the price of LNG liquefaction capacity

- **Infrastructure**

  In addition to the liquefaction plant, well and pumping equipment at the gas field, pipelines from the gas field to the liquefaction plant and harbor facilities for storage and loading LNG on to the LNG ships, need to be designed and built. Some of these costs, particularly onshore facilities directly related to the LNG plant, are sometimes included in the official cost figures, and make it necessary to control for these other cost factors.

- **Number of trains/capacity on trains**

  A liquefaction plant has a certain capacity, commonly measured in million tons natural gas processed per year. Taking into account that a liquefaction plant has a certain optimal lifetime, the lifetime capacity of the plant should of course not exceed the total amount of gas resources available at the site. Facilities made up of more trains are likely to have less supply problems, that is, if one train is down, the plant can still process LNG from the other trains. Gas turbine efficiency combined with larger turbine units have reduced the number of gas turbines needed in LNG plants by half since the industry start-up. This may have made it easier to build larger trains.

- **Organizational learning**

  There are 20 export plants currently in operation or under construction. This hypothesis is that organizational learning with respect to constructing and building a liquefaction train takes place each time a train is built, independent of whether the train is added to an existing plant or the train is set up on a Greenfield site. Experience with production of

  LNG may however have decreased the need for design redundancy, and hence, reduced the capital costs of later trains.

- **Experience induced process R&D**

  Many of the inventions seem to be of a process nature, that is, they have gradually reduced running costs. A steady decrease in power requirements may be the most apparent development of this kind.

- **Autonomous technological change**

  Some of the technologies used for LNG liquefaction, have improved independent of the development of the LNG business. One example is gas turbine technology, and we would expect materials technology to have improved in general.

- **Competition between suppliers of liquefaction technology**

  The number of competing technologies has varied a lot since the start of the LNG business. It is therefore also likely that the intensity of competition and the mark-up have varied considerably.

- **Output prices**

  The price on LNG has tended to follow the price on oil, mostly as a formal link in bilateral trade contracts. LNG may be regarded as a substitute for oil; both can be used as input to the chemical industry, to heat buildings, to produce electricity etc.

- **Other factors**

  Environmental/safety regulations may be important. In many cases, innovations that reduce running costs also reduce the environmental impact. Since energy for liquefaction is provided from burning natural gas at the site, an LNG facility emits large amounts of greenhouse gases like CO2, and local pollutants like nitric oxides etc. By improving energy efficiency, the emission intensity of the liquefaction process is also reduced. In fact, we may have that the quest for energy efficiency and reduced emissions drives capital costs up.

3.4) Design of Heat Exchanger:

3.4.1) Heat exchanger for pre-cooling (using propane):

Considering the gas flow rate as 400 MMscfd.

Density = 1.3686 Kg/m^3

Inlet of gas at 10^0 and leaving at -15^0 C

Inlet of propane -40^0 and leaving at -20^0 C

V=400 MMscfd = 11.328 MMscmd

=131.11 m^3 /sec.

= 131.11 * 1.3686 = 179.43 Kg/sec

Enthalpy of methane at -10^0 C = -35.277 KJ/Kg

Enthalpy of methane at -15^0 C = -89.891 KJ/Kg

Heat load of the heat exchanger Q = (H_1-H_2) V = (89.891-35.277) *179.43

= 9799.39 KJ/sec

Propane flowrate required W = Q/(Cp* ΔT)

= $9799.39*10^3/(2257.1*20)$

= 217.079 Kg/sec

Overall heat transfer coefficient U = 450 W/m^2 °C

Log mean temperature difference ΔT_{lm} = (ΔT1-ΔT2)/ ln{ΔT1/ΔT2)

= (50-5)/ln{50/5}

$$= 19.54^0 C$$

Area of heat transfer can be calculated as

$$Q = U*A*\Delta T_{lm}$$

$$A = Q/(U*\Delta T_{lm})$$

$$= 9799.39*10^3/(450*19.54)$$

$$= 1114.45\ m^2$$

Taking tube inner dia. 0.024 m and thickness 3mm

Number of tubes: $A = \Pi*do*L*n$ where, do= outside diameter of pipe

L = length of pipe

n = no. of tubes

$$n=A/(\Pi*do*L)$$

$$= 1114.45/(\Pi*0.03*6.9)$$

$$= 1714$$

Mass flowrate of gas Gh= m/(n*Ai)

$$= 179.43/(1714*\{\Pi/4*(0.024)^2\})$$

$$= 231.4\ Kg/m^2sec$$

Reynolds number $Re = di*G/\mu$ where, di= inner diameter of pipe

$$= 0.024*231.4/1.027*10^{-5}$$

$$= 540759.5$$

Sieder-Tate equation

$$hi*(D/k)= 0.023*(Re)^{0.8}*(Pr)^{0.33}$$

$$hi=(0.023*0.03281/0.024)*(540759.5)^{0.8}*(2209*1.027*10^{-5}/0.03281)^{0.33}$$

$= 1072.56\ W/m^2\,^{\circ}C$

Tube bundle diameter $Db = do\ (N/k)^{1/n}$

$= 0.03(1714/0.03281)^{1/2.207}$

$= 4.12\ m$

Tubes are arranged in square pitch and single pass flow arrangement.

Shell diameter $Ds = 4.12*1.1 = 4.53\ m$

Baffle spacing, $B = Ds/26$

$= 4.53/26$

$= 0.1742\ m$

Shell side flow area $= Ds*B*C/Pt$

$= 4.53*0.1742*0.006/0.036$

$= 0.1315\ m^2$

Mass flowrate of propane $Gc = 217.079/0.1315$

$= 1650.8\ Kg/m^2sec$

Equivalent diameter of square pitch arrangement

$De = 4\{Pt^2 - \Pi/4*do^2\}/\Pi\ do$

$= 4\{0.036^2 - \Pi/4*0.03^2\}/\ \Pi*0.03$

$= 0.025\ m$

Reynolds number $Re = De*Gc/\mu$

$= 0.025*\ 1650.8\ /1.3*10^{-5}$

$= 3174597.84$

$ho*(De/k) = 0.36(Re)^{0.55}(Pr)^{0.4}$

$ho = 0.36*(0.015198/0.025)*(3174597.84)^{0.55}*(2257.1*1.3*10^{-5}/0.015198)^{0.4}$

$= 1072.41 W/m^2 °C$

Overall heat transfer coefficient

$$Uo=1/\{(do/di)/hi +(1/ho)\}$$

$$= 1/\{ (1/1072.56)*(0.03/0.024) + (1/1072.41) \}$$

$$= 476.66 W/m^2 °C$$

Heat transfer area required = $\{9799.39*10^3/(476.66*19.54)\}$

$$= 1052.11\ m^2$$

Percentage excess area = (1114.45-1052.11)/ 1052.11

= 5.93%

So, the design can be accepted.

3.4.2) Heat Exchanger for Liquefaction (using Nitrogen):

V=179.43 Kg/sec

Inlet of gas at -15^0 and leaving at -162^0 C

Inlet of nitrogen -185^0 and leaving at -172^0 C

Enthalpy of methane at -15^0 C = -89.891 KJ/Kg

Enthalpy of methane at -162^0 C = -912.6 KJ/Kg

Heat load of the heat exchanger $Q = (H_1-H_2)\ V = (912.6-89.891)\ 179.43$

$= 147618.67$ KJ/sec

Nitrogen flowrate required $W = Q/(Cp^*\ \Delta T)$

$= 147618.67 * 10^3/(2201.9*13)$

$= 5157.03$ Kg/sec

Overall heat transfer coefficient U = 600 W/m^2 °C

Log mean temperature difference $\Delta T_{lm} = (\Delta T1-\Delta T2)/\ \ln\{\Delta T1/\Delta T2)$

$= (170-10)/\ln\{170/10\}$

$= 56.47^0$ C

Area of heat transfer can be calculated as $Q = U^*A^*\Delta T_{lm}$

$A = Q/(U^*\Delta T_{lm})$

$= 147618.67 * 10^3/(600*56.47)$

$= 4356.84$ m^2

Taking tube inner dia. 0.024 m and thickness 3mm

Number of tubes $A = \Pi*do*L*n$ where, do= outside diameter of pipe

L = length of pipe

N = no. of tubes

$n=A/(\Pi*do*L)$

$= 4356.84 /(\Pi*0.03*20.5)$

$= 2256$

Mass flowrate of gas $Gh= m/(n*Ai)$

$= 179.43/(2256*\{\Pi/4*(0.024)^2\})$

$= 175.9\ Kg/m^2sec$

Reynolds number $Re = di*Gh/\mu$ where, di= inner diameter of pipe

$= 0.024*175.9 /1.027*10^{-5}$

$= 411.06*10^3$

Sieder-Tate equation

$hi*(D/k)= 0.023*(Re)^{0.8}*(Pr)^{0.33}$

$hi=(0.023*0.03281/0.024)*(411.06*10^3)^{0.8}*(2209*1.027*10^{-5}/0.03281)^{0.33}$

$= 862.52 W/m^2\,^{o}C$

Tube bundle diameter $Db=do\ (N/k)^{1/n}$

$= 0.03(2256/0.03281)^{1/2.207}$

$= 4.67\ m$

Tubes are arranged in square pitch and single pass flow arrangement.

Shell diameter Ds= 4.67*1.1= 5.14 m

Baffle spacing, B = Ds/26

= 5.14/5

= 0.19769 m

Shell side flow area = Ds*B*C/Pt

$= 5.14*0.19769*0.006/0.036$

$= 0.1694\ m^2$

Mass flow rate of nitrogen Gc = 5157.03/0.1694

$= 30450.75\ Kg/m^2sec$

Equivalent diameter of square pitch arrangement

$De=4\{Pt^2-\Pi/4do^2\}/\Pi\ do$

$= 4\{0.036^2-\Pi/4*0.03^2\}/\ \Pi*0.03$

$= 0.025\ m$

Reynolds number Re = De*Gc/μ

$= 0.025*\ 30450.75\ /1.657*10^{-5}$

$= 45942594.62$

$ho*(De/k)=0.36(Re)^{0.55}(Pr)^{0.4}$

$ho=0.36*(0.015198/0.025)*(45942594.62)^{0.55}(2201.9*1.657*10^{-5}/0.015198)^{0.4}$

$= 5087.46\ W/m^2\ {}^{o}C$

Overall heat transfer coefficient

$Uo=1/\{(do/di)\ /hi\ +(1/ho)\}$

$= (1/862.52)*(0.03/0.024) + (1/5087.46)$

$= 607.61\ W/m^2\ {}^{o}C$

Heat transfer area required = $(147618.67\ *10^3/\{607.61\ *56.47\})$

$= 4302.31\ m^2$

Percentage excess area = (4356.84 - 4302.31)/ 4302.31

= 1.27%

So, the design can be accepted.

Chapter 4

Conclusion

The rate of expansion in the LNG industry is spectacular. Over the past five years, trade flows have increased by 29%, the liquefaction capacity by 48 bcm per year, the LNG fleet has grown by 75%. Major new LNG flows are connecting previously distinct regional markets & a global LNG market seems to be emerging. So in coming years there will be need for more and more liquefaction plant as the demand of LNG is going to grow many folds.

The need of liquefaction plant with efficient and cheaper liquefaction cost will be the demand of the hour so selection of technology is to be of prime consideration so that the components of the plant are of optimum size and should be cost wise viable for the plant.

Heat exchanger is the heart of the liquefaction plant its designing should be done with care so that its size is upto the requirement. In this report we have presented how the designing of the exchanger for both the pre-cooling and liquefaction can be done for a certain flow rate of feed gas. Following is the size of heat exchanger for a gas flow rate of 400 MMscfd:

Heat Exchanger for pre-cooling (using propane):

Overall heat transfer coefficient, U = 450 W/m^2 °C

Propane flow rate, W = 217.079 Kg/sec

Area of heat transfer, Q= 1114.45 m^2

Tube inner dia. 0.024 m and thickness 3mm

Number of tubes, n= 1714

Tube bundle diameter, Db= 4.119 m

Tubes are arranged in square pitch and single pass flow arrangement

Shell side flow area = 0.1315 m^2

Heat transfer area required = 1052.11 m^2

Percentage excess area = (1114.45-1052.11)/ 1052.11

= 5.93%

Heat Exchanger for Liquefaction (using Nitrogen):

Overall heat transfer coefficient, U = 600 $W/m^2\ ^oC$

Nitrogen flow rate, W = 5157.03 Kg/sec

Area of heat transfer = 4356.84 m^2

Tube inner dia. 0.024 m and thickness 3mm

Number of tubes, n= 2256

Tube bundle diameter, Db= 4.67 m

Tubes are arranged in square pitch and single pass flow arrangement

Shell side flow area= 0.1694 m^2

Heat transfer area required = 4302.3 m^2

Percentage excess area = (4356.84 - 4302.3)/ 4302.3

= 1.27%

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