

**A Project Report**  
**Production of LNG from Natural Gas**

*A Report submitted in partial fulfillment for the award of the requirements  
for the Degree of*

**Masters of Technology**

**IN**

**Gas Engineering**

**By**

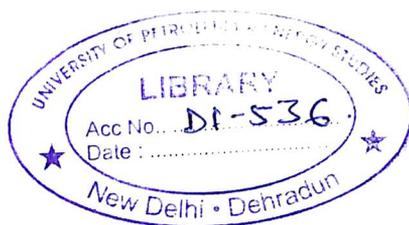
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### CERTIFICATE

This is to certify that the project work entitled on "**Production of LNG form natural gas**" submitted by **Akashy Govindrao Gajbhiye** in partial fulfillment of the requirement for the award of degree of Masters of Technology (Gas Engineering), at College of Engineering, University of Petroleum & Energy Studies, Dehradun is a record of the work carried out by him under my supervision and guidance.

  
17th May 2007

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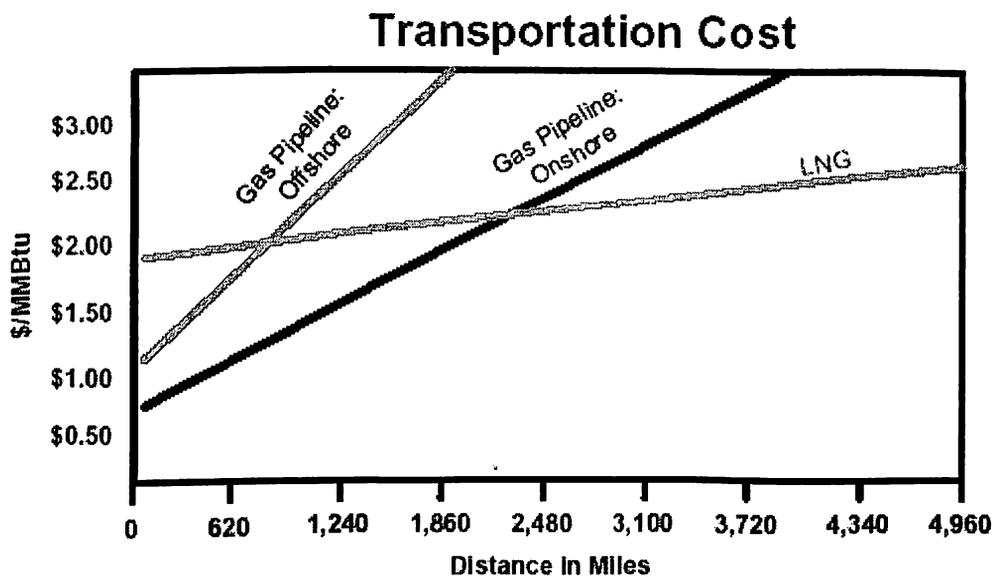
## **Abstract**

LNG production capacity continues to grow and the value of natural gas remains high. As of October 2005, there were a total of 19 liquefaction facilities in operation worldwide with a total of 69 trains. To cope up with rising energy demand and for the conveyance ease of LNG new LNG plants are setting up worldwide. Keeping this view in focus this project is a try to find out new technological trends, liquefaction processes, large trains through expandable processes, large drive equipments, etc. that have been briefly featured.

The parameters for selecting the process and factors influencing the expansion or installation of the large process plant have also been briefed. The case studies have been taken to tackle some problems arise in the plant during operation or during planning itself.

## 1) Introduction

Natural gas is set to become one of the most important primary energy sources for the 21<sup>st</sup> century. Compared with other fossil fuels, gas is relatively clean with regards to air pollution and greenhouse gas emissions and has larger proven reserves. Similarly, its ratio for heating purposes and as automotive fuel is growing rapidly. However, 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: litre of liquefied natural gas (LNG) corresponds roughly to 600 litres of natural gas. 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.



Source: Institute of Gas Technology.

Figure 1: Transportation cost of natural gas by various modes

## 1.2) Natural gas

Natural gas is primarily methane, with low concentrations of other hydrocarbons, water, carbon dioxide, nitrogen, oxygen and some sulphur 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 180000 billion m<sup>3</sup>, 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 billion m<sup>3</sup>. The annual increase was 1.5% in 1995 and 2.5% in 2005.

### 1.2.1) Composition of Natural Gas and LNG

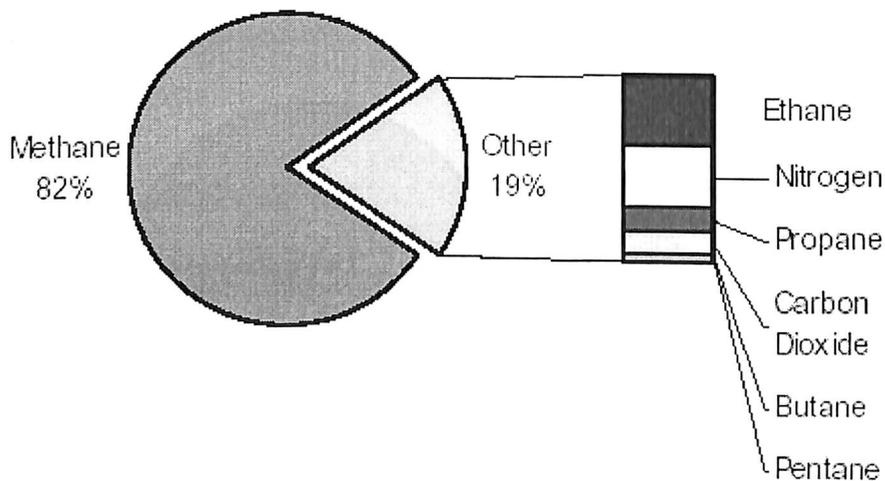


Figure 2: composition of natural gas

Source: center for energy economics, "Introduction to LNG" Jan '03

### 1.2.2) LNG:

Liquefied natural gas, or LNG, is natural gas in its liquid form. When natural gas is cooled to minus 259 degrees Fahrenheit (-161 degrees Celsius), it becomes a clear, colorless, odorless liquid. LNG is neither corrosive nor toxic. Natural gas is primarily methane, with low concentrations of other hydrocarbons, water, carbon dioxide, nitrogen, oxygen and some sulfur compounds. During the process known as liquefaction, natural

gas is cooled below its boiling point, removing most of these compounds. The remaining natural gas is primarily methane with only small amounts of other hydrocarbons. LNG weighs less than half the weight of water so it will float if spilled on water.

### 1.2.3) Properties of LNG

- Extremely low temperature - minus 260°F (minus 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 minus 160°F (minus 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.

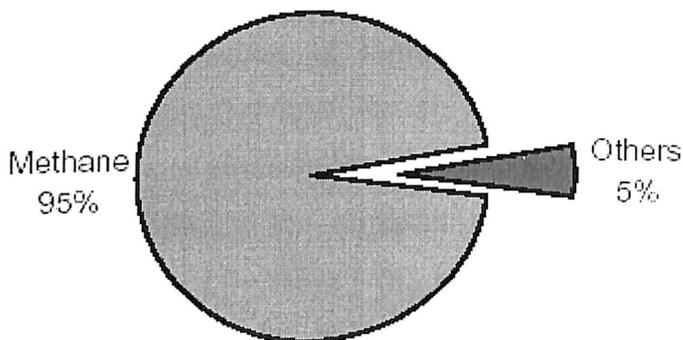


Figure 3: composition of LNG.  
Source: center for energy economics, "Introduction to LNG" Jan '03

### 1.3) The LNG chain:

In a typical scheme 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<sub>2</sub>S 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<sub>2</sub> removal to a few tenths of a percent may be required to prevent formation of solid CO<sub>2</sub>.

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.

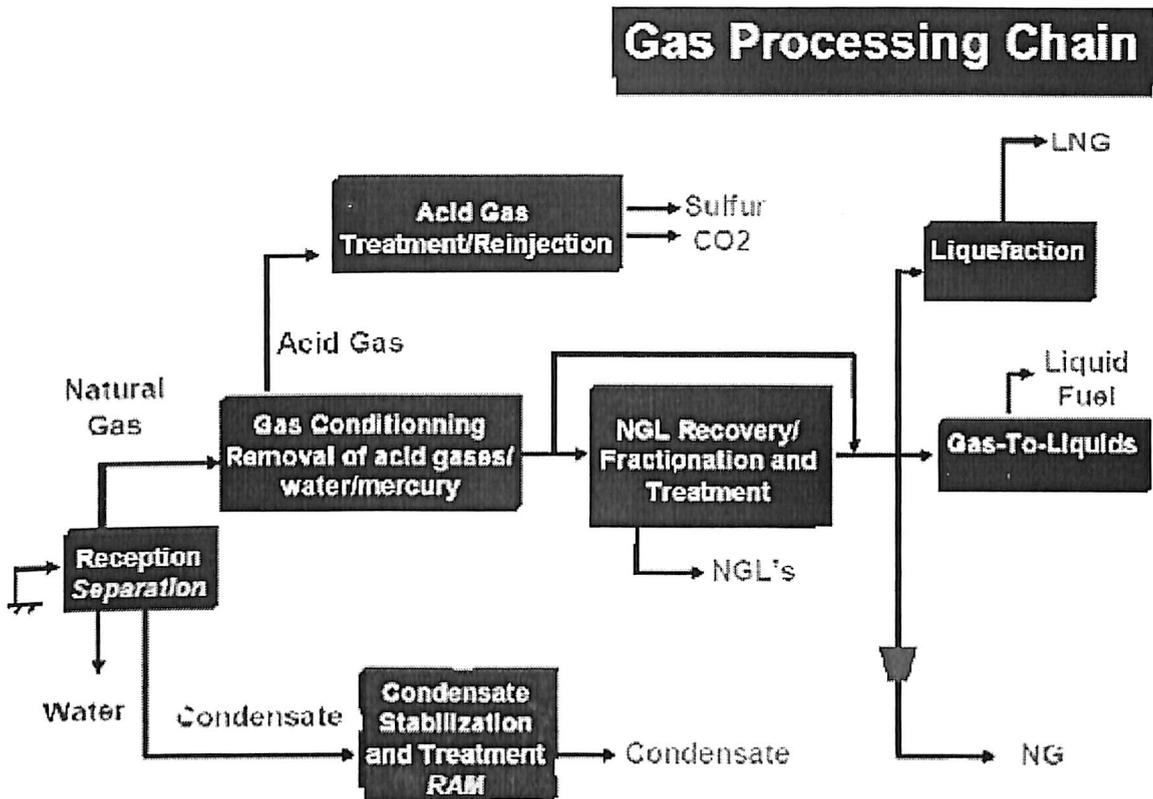


Figure 4: Gas processing chain

An example of a LNG plants overall flow scheme, and the main process units and supporting utilities, is shown in 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. It is further cooled in the cryogenic section to approximately  $-160\text{ }^{\circ}\text{C}$  and is completely liquefied. The resulting LNG is stored in atmospheric tanks ready for export by ship. 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. LNG is stored in specifically built insulated flat bottom storage tanks, operating at atmospheric pressure. These tanks can be above or below ground and keep the liquid at very low temperatures to minimize evaporation. The largest tanks currently in operation are 160 000 m<sup>3</sup> above ground and 200 000 m<sup>3</sup> under ground. LNG is warmed to a point where it converts back to its gaseous state. This is accomplished using special exchangers fed with high-pressure pumps for achieving the final gas pressure. Two types of exchangers are

used mainly: open-rack vaporizers (ORV) using sea water and submerged combustion vaporizers (SCV) that are water baths heated by combustion of fuel gas, nearly all the gas is distributed to end-users via a conventional gas pipeline network with recompressing stations at distances 150 to 250 km. LNG may be transported in special tanker trucks to small facilities, called satellite plants, where it is stored and regasified into a local pipeline for heating or process technology.

#### **1.4) 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 subcooling 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, subcooling.
- 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 subcooling.

### 1.5) Storage of LNG

There are three common types of LNG storage tanks, known as single containment, double containment, and full containment. The single containment storage has a 9 percent nickel self-supporting inner tank and a carbon steel outer wall. There is perlite insulation between the two tanks. In the event of an inner tank leak, the outer wall may fail because carbon steel is not suitable to cryogenic temperatures. In this case secondary containment is provided by a dike surrounding the tank.

The double containment tank has a post-tensioned concrete outer wall capable of holding cryogenic materials, and no dike is needed because the outer wall provides the secondary containment. However the cold vapors contacting the roof may cause the roof to fail, thus the containment is not "full containment" because vapors may be released in the event of an inner tank leak. The full containment tank is similar to double containment except that the roof is made of materials that can handle cryogenic temperatures; if the inner tank leaks, all liquids and vapors are still contained within the outer wall and roof.

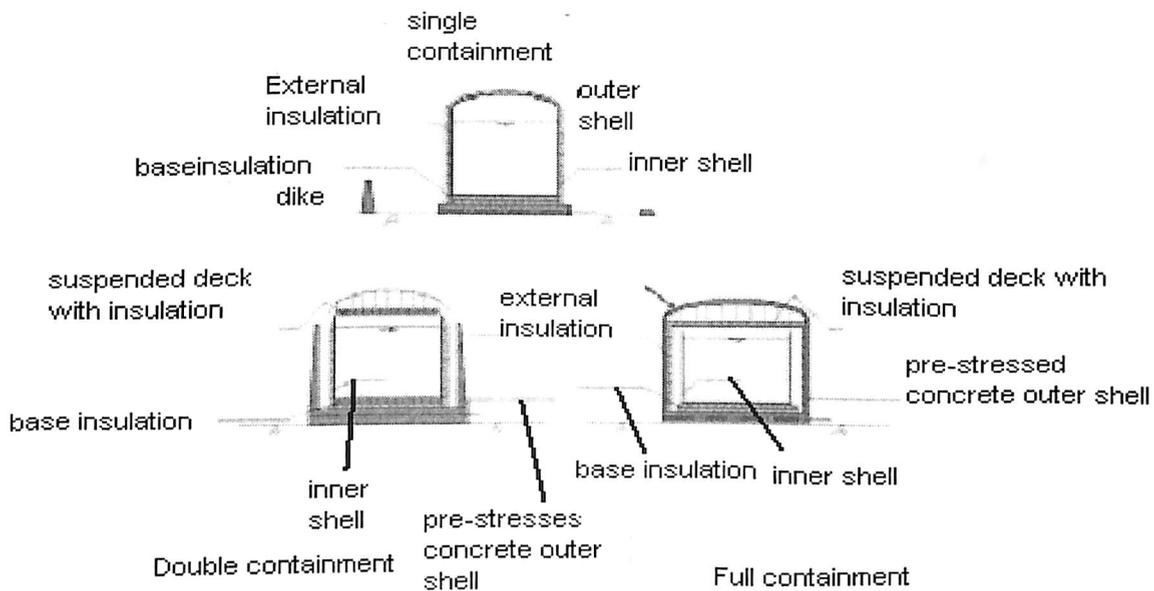


Figure 5: LNG storage tanks

Source: Proceedings of the twenty-second international pump users symposium-2005

## 2) Gas Processing

### 2.1) Gas sweetening:

#### Amine process for gas sweetening:

Natural gas is considered "sour" if hydrogen sulfide ( $H_2S$ ) is present in amounts greater than 0.25 grains per 100 standard cubic feet [gr/100 scf]. If  $H_2S$  is present, the gas is usually sweetened by absorption of the  $H_2S$  in an amine solution. Many chemical processes are available for sweetening natural gas. At present, the amine process is the most widely used method for  $H_2S$  removal. Other methods, such as carbonate processes, solid bed absorbents, and physical absorption, are also employed in sweetening plants.

The process is illustrated in Figure.

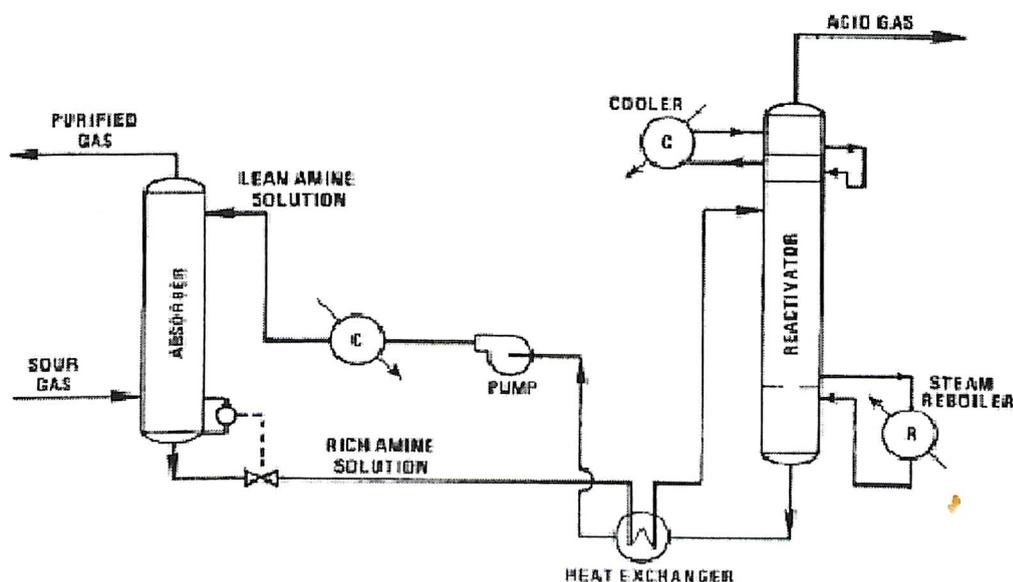


Figure 6: Amine process for gas sweetening  
Source: Natural gas processing

In this process the sour gas is sent upward through the contactor tower (absorber), countercurrent to the flow of amine. Acid gases are readily absorbed by the amine solution. The rich solution from the bottom of the contactor is sent through a heat exchanger where it gets preheated due to incoming lean solution, then the rich solution enters the top of the stripper (reactivator/regenerator) where it is stripped by steam that is generated by the reboiler. Outcoming steam from the top of the stripper is sent to a condensing unit to recover amine liquid.

The recovered hydrogen sulfide gas stream may be:

- (1) Vented,
- (2) Flared in waste gas flares or modern smokeless flares,
- (3) Incinerated, or
- (4) Utilized for the production of elemental sulfur or sulfuric acid.

If the recovered  $H_2S$  gas stream is not to be utilized as a feedstock for commercial applications, the gas is usually passed to a tail gas incinerator in which the  $H_2S$  is oxidized to  $SO_2$  and is then passed to the atmosphere out a stack.

## 2.2) Gas dehydration plant:

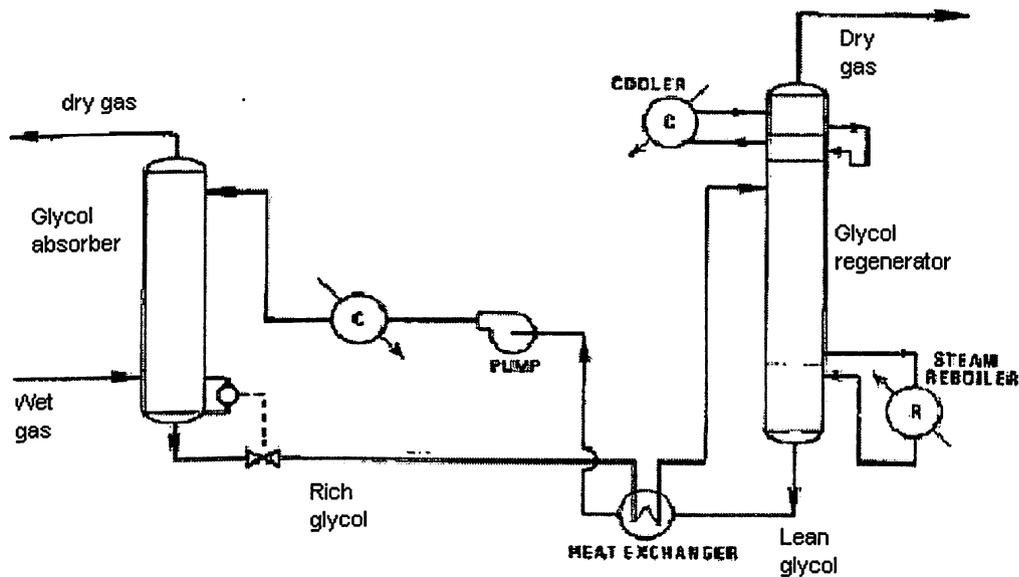


Figure 7: Gas dehydration  
Source: Natural gas processing

Natural gas can be dehydrated by several processes, including liquid desiccants (glycols), solid desiccants (alumina, silica gel, molecular sieves), expansion refrigeration, and calcium chloride. Usually the choice of dehydration method is in between glycol and solid desiccants. As glycol possesses some advantages over solid desiccants, hence glycol method is discussed here.

As shown in figure, wet natural gas flows through separator or scrubber to remove all liquid and solid impurities. Then the gas flows into and upward through the contactor where it is contacted countercurrently and dried by the glycol.

Reconcentrated on lean glycol enters the top of the contactor where it flows downward from tray to tray and absorbs water from the rising natural gas. The wet or rich glycol leaves the absorber and flows through the coil in the accumulator where it is preheated by hot lean glycol. After the glycol/glycol heat exchanger, the glycol enters the stripping column and flows down the packed bed section in to the reboiler. Steam generator in the reboiler strips water from the liquid glycol as it rises up the packed bed. The water vapor and desorbed natural gas are vented from the top of the stripper. The hot reconcentrated glycol flows out of the reboiler into the accumulator where it is cooled by heat exchange with rich glycol. Finally, the lean glycol flows through the glycol/gas exchanger and is pump back into the top of the absorber.

**Table-1: Typical Feed LNG Plant Specifications after processing:**

Nitrogen	$\leq 1\%$
Carbon Dioxide	$< 50$ ppmv
Hydrogen Sulphide	$< 4$ ppmv
Total S (H <sub>2</sub> S+CO <sub>2</sub> +organic Sulphur)	$< 20$ ppmv
Water	$< 0.5$ ppmv
Mercury	$< 0.01$ microgram/Nm <sup>3</sup>
Butanes	2% max
Pentanes+	0.1% max

### 3) Properties of Refrigerants:

#### 3.1) Methane Properties

**Formula:** CH<sub>4</sub>

**Molecular weight:** 16.043-g/mol

##### Solid phases

- Melting point : -182.5 °C
- Latent heat of fusion (1,013 bar, at triple point) : 58.68 kJ/kg

##### Liquid phase

- Liquid density (1.013 bar at boiling point) : 422.62 kg/m<sup>3</sup>
- Liquid/gas equivalent (1.013 bar and 15 °C (59 °F)) : 630 vol/vol
- Boiling point (1.013 bar) : -161.6 °C
- Latent heat of vaporization (1.013 bar at boiling point) : 510 kJ/kg

##### Critical point

- Critical temperature : -82.7 °C
- Critical pressure : 45.96 bar

##### Gaseous phase

- Gas density (1.013 bar at boiling point) : 1.819 kg/m<sup>3</sup>
- Gas density (1.013 bar and 15 °C (59 °F)) : 0.68 kg/m<sup>3</sup>
- Compressibility Factor (Z) (1.013 bar and 15 °C (59 °F)) : 0.998
- Specific gravity (air = 1) (1.013 bar and 21 °C (70 °F)) : 0.55
- Specific volume (1.013 bar and 21 °C (70 °F)) : 1.48 m<sup>3</sup>/kg
- Heat capacity at constant pressure (Cp) (1 bar and 25 °C (77 °F)) : 0.035 kJ/(mol.K)
- Heat capacity at constant volume (Cv) (1 bar and 25 °C (77 °F)) : 0.027 kJ/(mol.K)
- Ratio of specific heats (Gamma:Cp/Cv) (1 bar and 25 °C (77 °F)) : 1.305454

- Viscosity (1.013 bar and 0 °C (32 °F)) : 0.0001027 Poise
- Thermal conductivity (1.013 bar and 0 °C (32 °F)) : 32.81 mW/(m.K)

### **Miscellaneous**

- Solubility in water (1.013 bar and 2 °C (35.6 °F)) : 0.054 vol/vol
- Autoignition temperature : 595 °C

### **3.2) Propane Gas Properties**

**Formula: C<sub>3</sub>H<sub>8</sub>**

#### **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/m<sup>3</sup>
- 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/m<sup>3</sup>

- Gas density (1.013 bar and 15 °C (59 °F)) : 1.91 kg/m<sup>3</sup>
- 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 m<sup>3</sup>/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
- Autoignition temperature : 470 °C

### **3.3) Nitrogen Gas Properties**

**Formula : N<sub>2</sub>**

#### **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/m<sup>3</sup>

- 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/m<sup>3</sup>

### **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/m<sup>3</sup>
- Gas density (1.013 bar and 15 °C (59 °F)) : 1.185 kg/m<sup>3</sup>
- 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 m<sup>3</sup>/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 %

### **3.4) Ethane Gas Properties**

**Formula: C<sub>2</sub>H<sub>6</sub>**

#### **Molecular Weight**

- Molecular weight : 30.069 g/mol

#### **Solid phase**

- Melting point : -183.3 °C
- Latent heat of fusion (1,013 bar, at triple point) : 94.977 kJ/kg

#### **Liquid phase**

- Liquid density (1.013 bar at boiling point) : 546.49 kg/m<sup>3</sup>
- Liquid/gas equivalent (1.013 bar and 15 °C (59 °F)) : 432 vol/vol
- Boiling point (1.013 bar) : -88.7 °C
- Latent heat of vaporization (1.013 bar at boiling point) : 488.76 kJ/kg
- Vapor pressure (at 21 °C or 70 °F) : 38.3 bar

#### **Critical point**

- Critical temperature : 32.2 °C
- Critical pressure : 48.839 bar

#### **Gaseous phase**

- Gas density (1.013 bar at boiling point) : 2.054 kg/m<sup>3</sup>
- Gas density (1.013 bar and 15 °C (59 °F)) : 1.282 kg/m<sup>3</sup>
- Compressibility Factor (Z) (1.013 bar and 15 °C (59 °F)) : 0.9912
- Specific gravity (air = 1) (1.013 bar and 15 °C (59 °F)) : 1.047
- Specific volume (1.013 bar and 21 °C (70 °F)) : 0.799 m<sup>3</sup>/kg

- Heat capacity at constant pressure ( $C_p$ ) (1 bar and 25 °C (77 °F)) : 0.053 kJ/(mol.K)
- Heat capacity at constant volume ( $C_v$ ) (1 bar and 25 °C (77 °F)) : 0.044 kJ/(mol.K)
- Ratio of specific heats ( $\gamma:C_p/C_v$ ) (1 bar and 25 °C (77 °F)) : 1.193258
- Viscosity (1.013 bar and 0 °C (32 °F)) : 0.0000855 Poise
- Thermal conductivity (1.013 bar and 0 °C (32 °F)) : 18 mW/(m.K)

### **Miscellaneous**

- Solubility in water (1.013 bar and 20 °C (68 °F)) : 0.052 vol/vol
- Autoignition temperature : 515 °C

### **3.5) Ethylene Gas Properties**

**Formula :  $C_2H_4$**

### **Molecular Weight**

- Molecular weight : 28.054 g/mol

### **Solid phase**

- Melting point : -169.2 °C
- Latent heat of fusion (1,013 bar, at triple point) : 119.37 kJ/kg

### **Liquid phase**

- Liquid density (1.013 bar at boiling point) : 567.92 kg/m<sup>3</sup>
- Liquid/gas equivalent (1.013 bar and 15 °C (59 °F)) : 482 vol/vol
- Boiling point (1.013 bar) : -103.8 °C
- Latent heat of vaporization (1.013 bar at boiling point) : 482.86 kJ/kg
- Vapor pressure (at 5 °C or 41 °F) : 47.7 bar

### **Critical point**

- Critical temperature : 9.5 °C
- Critical pressure : 50.76 bar

### **Gaseous phase**

- Gas density (1.013 bar at boiling point) : 2.085 kg/m<sup>3</sup>
- Gas density (1.013 bar and 15 °C (59 °F)) : 1.178 kg/m<sup>3</sup>
- Compressibility Factor (Z) (1.013 bar and 15 °C (59 °F)) : 0.9935
- Specific gravity (air = 1) (1.013 bar and 0 °C (32 °F)) : 0.974
- Specific volume (1.013 bar and 21 °C (70 °F)) : 0.862 m<sup>3</sup>/kg
- Heat capacity at constant pressure (Cp) (1.013 bar and 15 °C (59 °F)) : 0.042 kJ/(mol.K)
- Heat capacity at constant volume (Cv) (1.013 bar and 15 °C (59 °F)) : 0.034 kJ/(mol.K)
- Ratio of specific heats (Gamma:Cp/Cv) (1.013 bar and 15 °C (59 °F)) : 1.242623
- Viscosity (1.013 bar and 0 °C (32 °F)) : 0.0000951 Poise
- Thermal conductivity (1.013 bar and 0 °C (32 °F)) : 16.83 mW/(m.K)

### **Miscellaneous**

- Solubility in water (1.013 bar and 0 °C (32 °F)) : 0.226 vol/vol
- Autoignition temperature : 425 °C

#### **4) Processes:**

##### **4.1) 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  $-40\text{ }^{\circ}\text{C}$ . 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  $\text{C}_3$  streams and compresses the vapor to 15 - 25 bara 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  $-35\text{ }^{\circ}\text{C}$  to between  $-150\text{ }^{\circ}\text{C}$  to  $-160\text{ }^{\circ}\text{C}$ . 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.

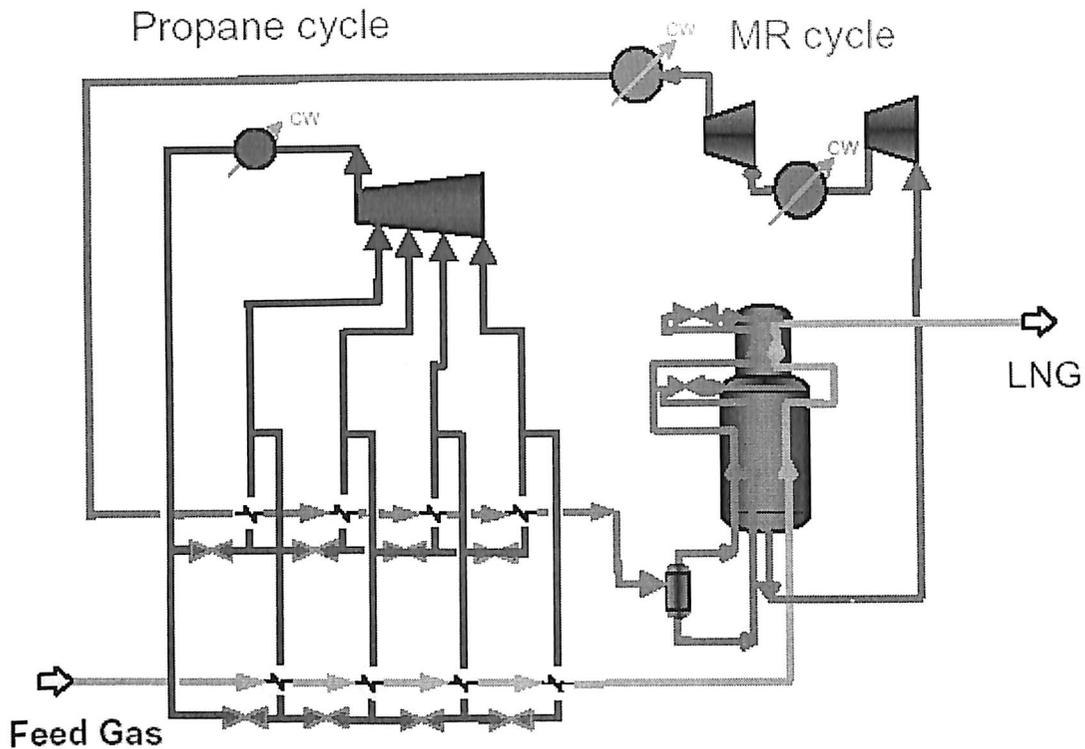


Figure 8: APCI propane pre-cooled mixed refrigerant process  
 Source: Natural gas liquefaction processes comparison, Poster PO-39

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 subcooling duties outside the MCHE.

#### 4.2) 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  $-161^{\circ}\text{C}$ .

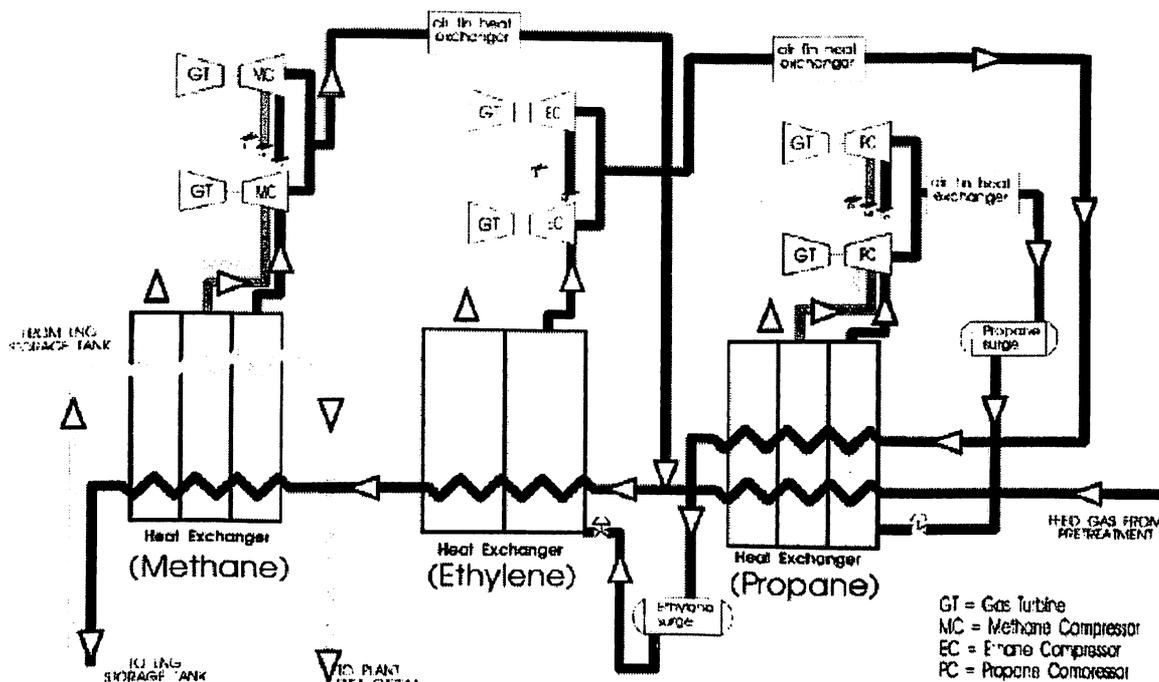


Figure 9: Phillips Cascade LNG Process

### 4.3) 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).

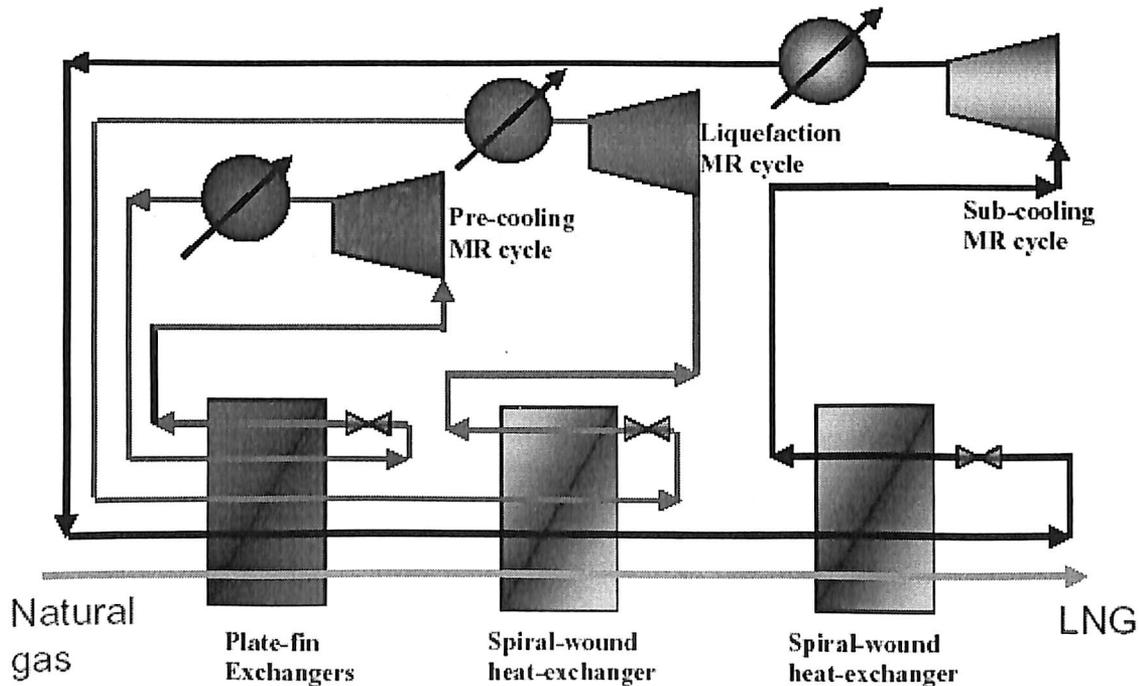
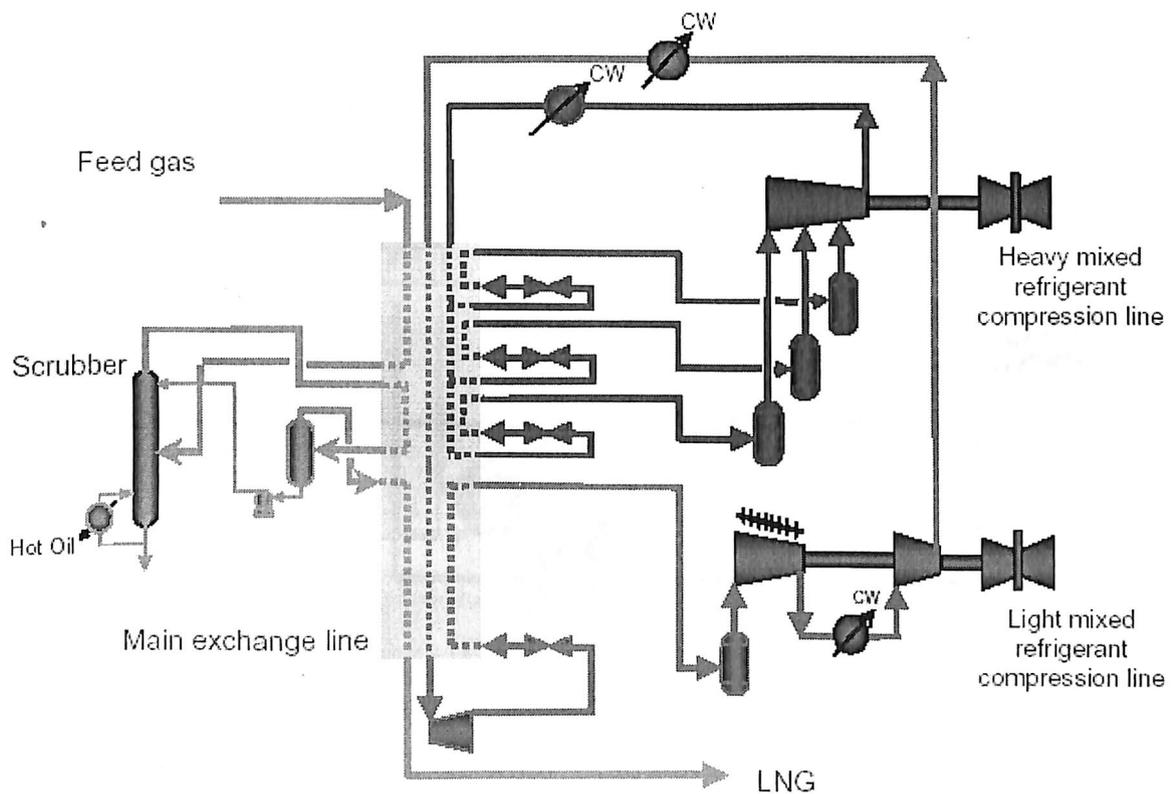


Figure 10: Statoil/Linde mixed fluid cascade process  
 Source: Natural gas liquefaction processes comparison, Poster PO-39

**4.4) 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.



Source: Natural gas liquefaction processes comparison, Poster PO-39  
 Figure 11: Axens Liquefin process

#### 4.5 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.

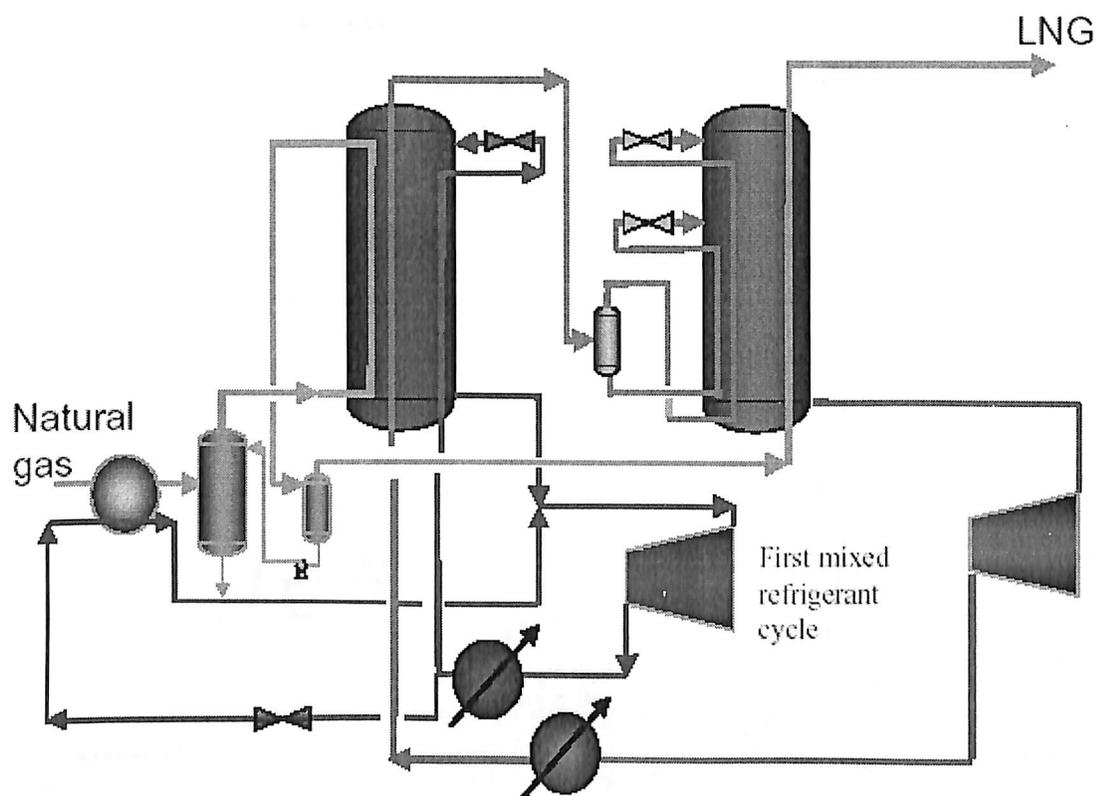


Figure 12: Shell double mixed refrigerant process  
Source: Natural gas liquefaction processes comparison, Poster PO-39

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

### 5.1) 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<sub>2</sub>+ hydrocarbons a hydrocarbon liquid draw system would be included.

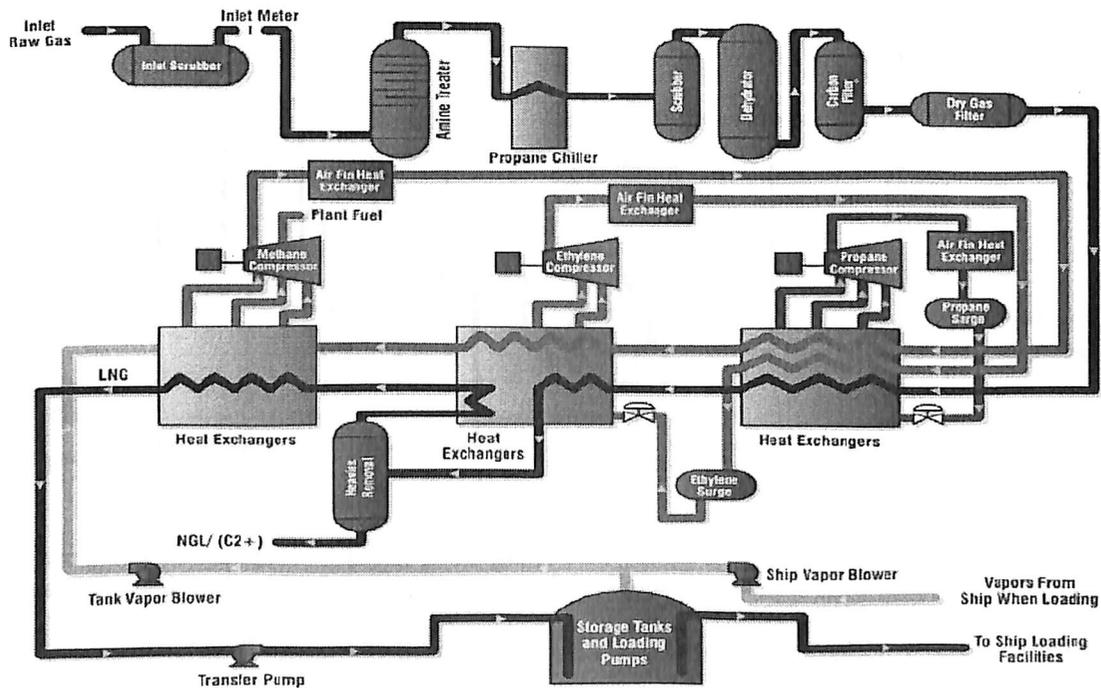


Figure 13: Phillips Optimized Cascade LNG Process

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.

**5.2) APCI APX (expandable) process:**

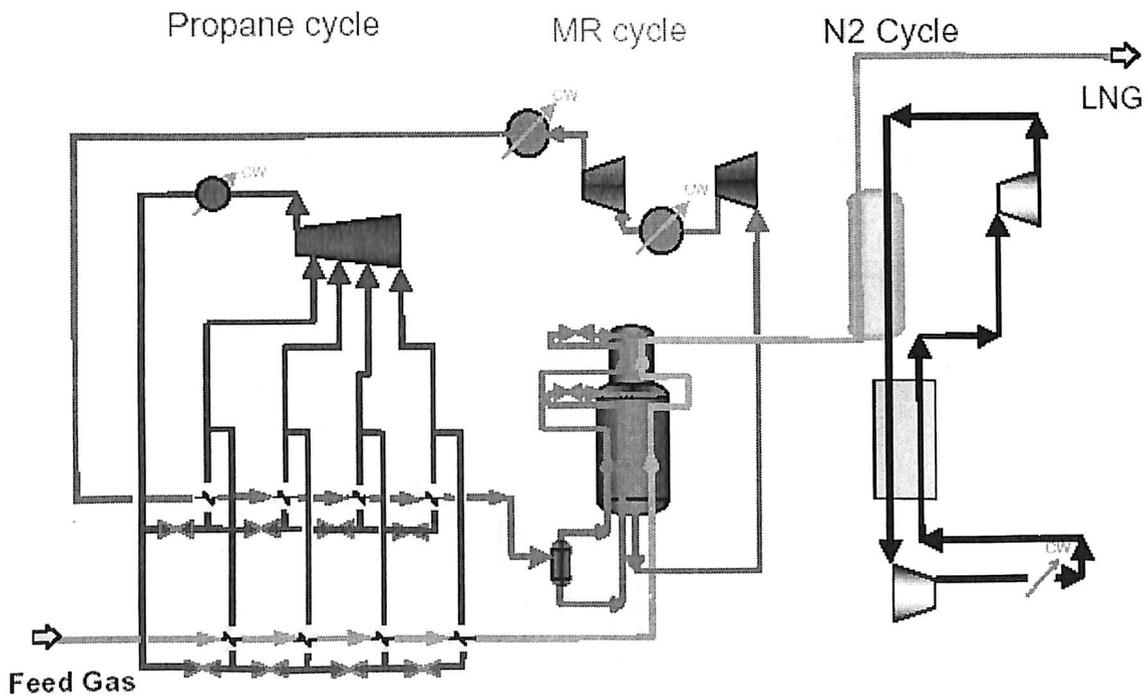


Figure 14: APCI APX (expandable) process  
Source: Natural gas liquefaction processes comparison, Poster PO-39

Modification to this process is that, it adds a nitrogen cycle to the C<sub>3</sub>MR process, which liquefy and subcools 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.

**5.3) Shell (PMR) Parallel Mixed Process:**

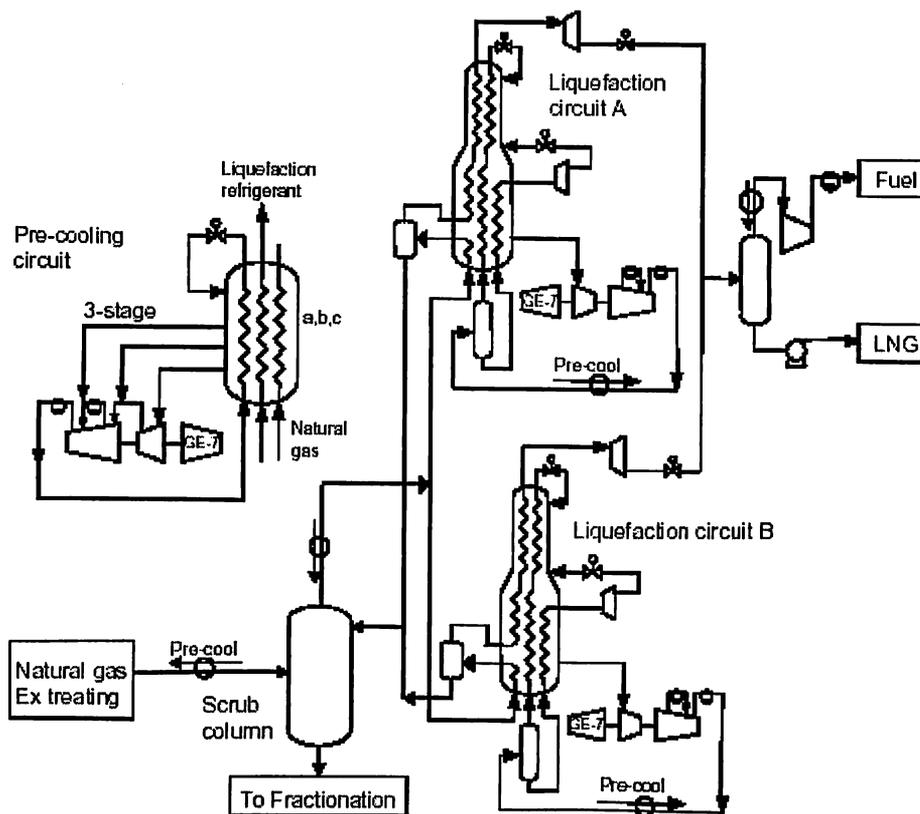


Figure 15: Shell (PMR) Parallel Mixed Process  
 Source: LARGE CAPACITY LNG PLANT DEVELOPMENT- Paper PS5-3

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, 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 upto 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.

**6) Equipments**

**6.1) plate fin heat exchanger**

Brazed aluminium plate-fin heat exchangers are capable of handling a wide variety of streams. The fins provide more surface area and improve the heat transfer coefficient, thereby reducing the size of the heat exchanger. The heat-transfer surface consists of a series of stainless-steel plates, each stamped with a corrugated pattern designed for a combination of high strength, efficiency, and fouling resistance. The number and design of the plates vary according to the desired heat-transfer capacity. The plates are stacked together with thin sheets of copper or nickel between each plate. The plate pack, end plates, and connections are brazed together in a vacuum furnace to secure the plates together at the edges and at all contact points. Inlet and outlet connections are available in a number of different styles. The brazed-plate heat exchanger is compact, rugged, and provides high heat transfer capability. Its heat-transfer surface area is concentrated in a very small volume. The corrugations in the plates induce turbulent flow to increase heat transfer and reduce fouling.

The plate-fin heat exchanger is characterized by its compact, lightweight, efficient design. It provides significantly more cooling per cubic inch. The plate-fin heat exchanger usually is fabricated from aluminum, but regardless of the materials of construction, the core is furnace brazed in a controlled atmosphere or high vacuum.

Because plate-fin heat exchangers require more precise manufacturing, they are usually more expensive than the other heat exchangers

### **6.2) Spiral wound heat exchanger:**

In this type of heat exchanger the tube bundles are wound spirally, just like weaving the knit. Spiral wound exchanger is expensive. For the higher capacities planned for the future, it would be possible to be increasing LMTD of this exchanger to stay within a feasible size.

### **6.3) Compressors and drivers:**

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 advantage of higher efficiencies of axial compressors in large refrigeration system are in demand. Axial compressors achieve typical polytropic efficiencies 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.

The choice of drivers, compressors and driver arrangements, is depend upon the process and power generation. Earlier steam and gas turbines were used mostly in the process, but nowadays electric motors have got attention for large capacity LNG plants to sustain the baseloads.

**Table 2: 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)	Some equipment/ processes may require further development

	LNG)	
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Source: Hydrocarbon engineering -Feb 2004- Dr. Tariq Shukri

### 6.4) Important parameters for selection of technology

The selection of the technology is mainly based on for the process, utilities and offsite units of the plant, which include proprietary and non-proprietary technologies. The selection could be between alternative processing technologies for the operating units, the type of major equipment, or utilities schemes.

**6.5.1) Compressor Efficiency.** Depending upon the compressor efficiency considered, the LNG production can vary a lot: the LNG production is increased by nearly 10% when the polytropic efficiency is changed from 79 to 85% (see figure). This has to be checked carefully when making comparisons.

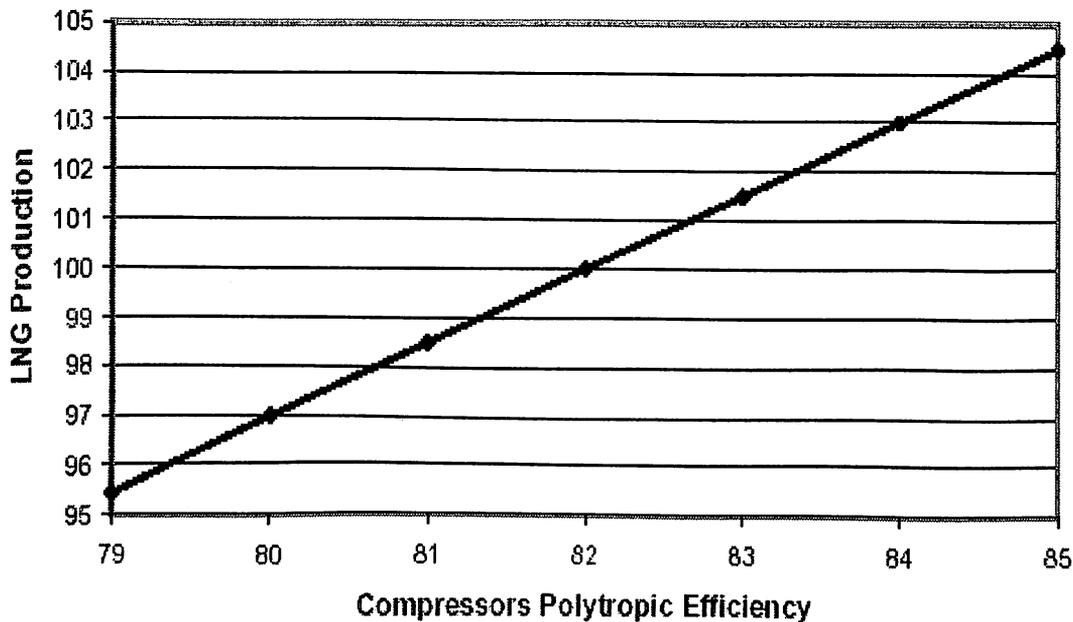


Figure 16: Effect on LNG production of compressor efficiency  
Source: Natural gas liquefaction processes comparison, Poster PO-39

### 6.5.2) Temperature Approach on The Main Condenser

The size of the condenser will depend upon whether the refrigerant of the first cycle is a pure component (propane as in the C3/MR and cascade), or a mixed refrigerant (Liquefin and Linde process). With a pure component, the condensation is done at a fixed temperature (the dew point temperature is the same as the bubble point temperature), whereas with a mixed refrigerant, the temperature varies linearly between the dew point temperature and the bubble point temperature. (See figure). Either the condenser will be much smaller with a mixed refrigerant in the first cycle, or inversely with the same condenser size, the LNG production will be increased.

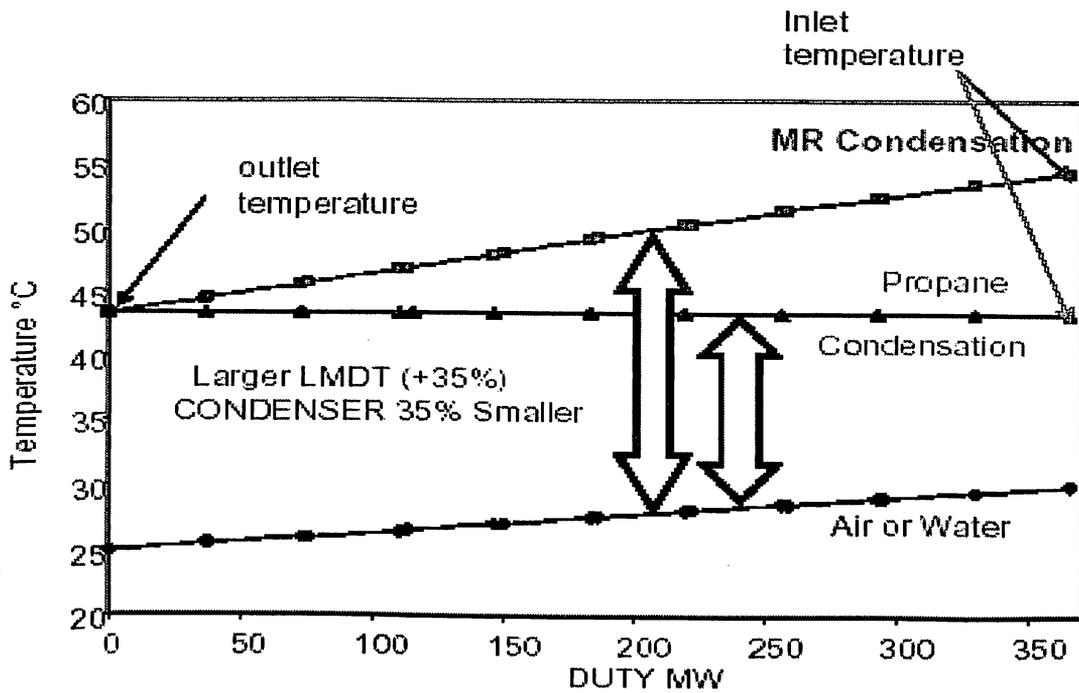


Figure 17: Effect of using propane or mixed refrigerant on the size of the condenser  
 Source: Natural gas liquefaction processes comparison, Poster PO-39

## **6.6) Factors influencing the price of LNG liquefaction capacity**

### **6.6.1) 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 harbour 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.

### **6.6.2) 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.

### **6.6.3) 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.

#### **6.6.4) 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.

#### **6.6.5) 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.

#### **6.6.6) 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.

#### **6.6.7) 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.

#### **6.6.8) 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 CO<sub>2</sub>, 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.

## **7) Case study:**

### **1) Trinidad LNG project on expansion (constructing 4<sup>th</sup> Train)**

**Case:** optimization of the process and equipments to overcome land constraint.

#### **Introduction**

The Atlantic LNG facility located in Point Fortin, Trinidad is comprised of three LNG trains using the Phillips Optimized Cascade LNG Process and each producing approximately 3.3 MTPA of LNG. Two 102,000 m<sup>3</sup> and one 160,000 m<sup>3</sup> LNG tanks along with one jetty the LNG export complete the facilities. The fourth LNG train will be designed for a nominal production capacity of approximately 5.2 MTPA. Included in the project is one additional 160,000 m<sup>3</sup> LNG tank and one additional jetty. A new 56" onshore pipeline will supply additional feed gas to the LNG Trains at Pt Fortin, Trinidad. 3.5 trillion cubic feet (tcf) of gas has been earmarked to support Train 1 production over a 20-year period, the two-train expansion proposes to utilize some 7.7tcf of gas over a similar 20-year period. Current proven and probable reserves of natural gas in Trinidad and Tobago are estimated at 30.7tcf. Experts estimate that another 40tcf of gas reserves may exist and this will keep Trains 3 and 4 busy for many years to come.

#### **The major shareholders**

British Gas, British Petroleum, Repsol, ENI and PetroCanada.

#### **Challenges**

Bechtel Overseas Inc has utilized the Phillips optimized cascade LNG process to develop the 4<sup>th</sup> train. The basic design premise used for all Atlantic LNG Company of Trinidad & Tobago facilities apply for the fourth LNG train and maintains the basic "Two-Trains-in-One" at the higher production level.

The challenge was to accommodate the drivers, compressors, heat exchangers, power generators, condensers, process equipment, etc. in the space constraint. So the whole team and management were looking for the layout of the possible train and the

optimization of the process, process equipments so that effective work can be extracted from available resources.

### **Solution to the problem**

- 1) Land reclamation was the alternative to the space constraint.
- 2) The fourth LNG train was designed to use to three propane and three ethylene compressors and turbines the two methane compressors and turbines sets. Intercoolers can be added to the methane compressors above the standard design to improve efficiency resulting in increased LNG production.
- 3) The propane and ethane single case compressor can be split into two.
- 4) Additional propane condenser can be installed to improve refrigeration.
- 5) Plate fin heat exchanger can be used, as it takes less space and gives large surface area per unit volume.
- 6) The foam glass insulation can be replaced by perlite.
- 7) The turbines of high capacity can be installed to couple two different compressors or units in a single driver.

### **Suggestions:**

The main consideration should be the replacement of the high capacity electric drivers. Another thing is the use of high capacity plate fin heat exchanger for more refrigeration. Other small effective techniques may help in to increase efficiency of the process.

**Description to the solution:**

1) In the first three trains, they were using Frame 5C or 5D turbines; instead of that Frame 7EA turbines can be used.

**MS5002D (“Frame-5D”)**

Two-shaft, heavy-duty gas turbine, specifically designed for high operating efficiency over a wide range of speed and load for mechanical drive applications, shows ease of operation and very high reliability and availability.

<b>MS5002D</b>	<b>Mech. Drive</b>
ISO shaft power	32.6MW
Speed (rpm)	4,670
Thermal efficiency	29.4%

**MS7001EA (“Frame-7”)**

Its simplicity and size make it an extremely flexible component in plant layout and an easy, low-cost addition of power when phase capacity expansion is needed. The unit is ideal for plants that require high efficiency and shaft speed for direct coupling to the generator.

<b>MS7001EA</b>	<b>Power Gen</b>	<b>Mech. Drive</b>
ISO shaft power	85.4MW	87.3MW
Speed (rpm)	3,600	3,600
Thermal efficiency	32.7%	33.1%

Due to flexibility of the Frame 7 and high efficiency, this version can be useful.

2) Use of perlite instead of foam glass.

Perlite is naturally occurring siliceous rock. When heated to a suitable point in its softening range, it expands from four to twenty times its original volume. When quickly heated to above 1600°F (871°C), the crude rock pops in a manner similar to popcorn as the combined water vaporizes and creates countless tiny bubbles which account for the amazing light weight and other exceptional physical properties of expanded perlite.

**Typical physical properties**

Specific Gravity	- 2.2 - 2.4
Fusion Point	- 1260-1343°C
Specific Heat	- 387 J/kg·K
Thermal Conductivity at 75°F (24°C)	- 0.04-0.06 W/m·K

**Foam glass:**

It is a lightweight, extremely fine-pored expanded glass with millions of hermetically sealed pores. Since no diffusion can take place, the material is watertight and achieves an efficient barrier against soil humidity. Foam glass can be manufactured fully out of waste glass, with only a minimum of virgin additives.

**Composition**

Glass (recycled or virgin) -98 % wt  
Additives (inorganic salts) -2 % wt

**Thermal Conductivity:** (W/m X K)- 0.092

So, comparing the thermal conductivities, perlite insulation can be used.



### 3) Plate-fin heat exchangers

Most plate-fin heat exchangers are made of aluminium, with a vacuum-brazed core. Corrosion-resistant and heat-resistant brazing alloys can be used. The maximum operating temperature of a plate-fin heat exchanger is a function of its construction materials. Aluminium brazed plate-fin heat exchangers can be used from cryogenic temperatures (-270°C) up to 200°C, depending on the pipe and header alloys. Stainless steel plate-fin heat exchangers are able to operate at up to 650°C, while titanium units can tolerate temperatures approaching 550°C. Aluminium brazed units can operate at up to 120 bar, depending on the physical size and the maximum operating temperature.

A plate-fin heat exchanger with 6 fins/cm provides approximately 1,300 m<sup>2</sup> of surface per m<sup>3</sup> of volume. This heat exchanger would be approximately 10% of the volume of an equivalent shell and tube heat exchanger with 19 mm tubes.

So, compared to shell and tube heat exchanger plate fin heat exchanger will give more heat transfer per unit volume.

## **2. The proposed project of expansion at Nigeria LNG**

**Case:** Reducing flaring by the gas producers in the Niger Delta, which contributes to global warming.

### **Introduction**

The original NLNG plant site was designed to provide scope for future expansion and the current 4th and 5th train (NLNGPlus) project is the next stage in the development of the NLNG project. The Base plant (two trains) has a production capacity of 5.8 million tonnes per annum (Mtpa) of LNG and the Expansion project (3rd train) will increase this by 2.9 Mtpa. The NLNGPlus project will further increase the production capacity by 8.1 Mtpa. The Base Project trains were originally designed to process non oil-associated gas (NAG), but additional facilities were installed and modifications made during construction of the 3rd train to enable all the three trains process oil associated gas (AG). Each new train comprises an Acid Gas Removal Unit, a Mol sieve unit to remove water, a Mercury Removal Unit made of guard bed to trap any trace levels of mercury in the feed gas, a propane pre-cooled mixed refrigerant unit to liquefy the gas and a fractionation unit to produce export grade propane and butane products

### **Challenges:**

The NLNGPlus project will increase the capacity of the LNG complex to process oil-associated gas, much of which is currently wasted through flaring. The flaring of the gas will increase the wastage of the gas, and hence the Greenhouse effect.

### **Solution to the problem:**

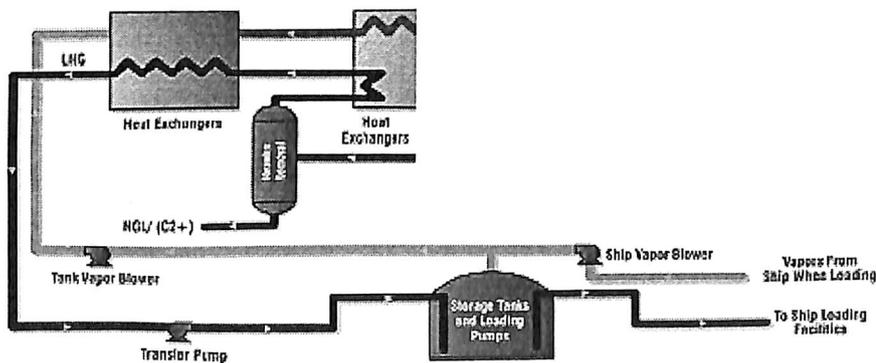
- 1) The associated gas stream can be sent into the separators to separate the condensate. This condensate can be fractionated to recover valuable products like propane, butane, etc.
- 2) The gas which is being flared, the stream can be diverted to the liquefaction process by passing it through the main cryogenic heat exchanger so that the



additional gas stream also get liquefied and can be added to the production of the plant.

1) Re-liquefaction of the vaporize gas

- During the ship loading operations an important amount of vaporized LNG is sent to the flare system where it burns through the duration of the loading operation. In a world-class base load plant processing 4 MMtpy of LNG, it is estimated that between 10 to 14 MMUS \$ are burned per year. The amount of gas vented is equivalent to a rate of 100 MMscfd, during the 12 hours of loading.
- For the installed five process trains with an overall capacity of 2,810 MMscfd gas intake. If we consider the BOG rate at normal that from both storage tanks and loading operation at 0.4% per day then it will count to 11.24 MMscfd.
- If the this vaporize gas is send for re-liquefaction and if considered that nearly 40% get added to the LNG production, than additional nearly 5 MMscfd of gas get utilized, which may ultimately reduce the greenhouse gases.



2) From the process stream the heavier fractions can be taken out after condensing it. Then each circuit may have separate fractionation unit to recover propane and butane products and other heavy products.

**Suggestions:**

- The C<sub>3</sub>MR process with another nitrogen cycle can be the best alternative for the expansion plan, as they're already C<sub>3</sub>MR process is operating.
- While loading of the tanker boil off gas can also be recycled to the liquefaction process to add to the LNG, where the gas being flared off.

### 8) Development of LNG liquefaction plant equipments:

Natural gas is liquefied at  $-162^{\circ}\text{C}$ . after passing through the process of phase separation in separators, sweetening for removal of acid gases, dehydration for moisture removal and then for liquefaction at required specifications, the gas is liquefied. The temperature is bringing down by the help of refrigerants. Here, the pure refrigerants used are propane for pre-cooling; to bring down the temperature of process gas to  $-15^{\circ}\text{C}$  in shell and tube heat exchanger and nitrogen for liquefaction, to liquefy the process gas at  $-162^{\circ}\text{C}$  in plate heat exchanger.

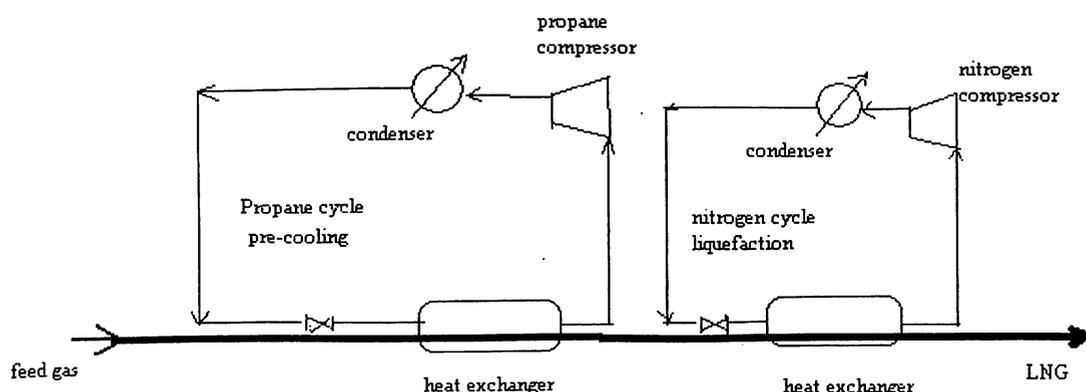
Any liquefaction plant operates on refrigeration principle.

➤ Compression, condensation, expansion and evaporation.

In refrigeration cycle the refrigerant is first compressed, then it is cooled in condenser, the condensed refrigerant is then expands in valve and finally vaporizes with the process gas, as it transfers heat to gas to cool it.

The main components of the refrigeration plant are mainly heat exchangers and compressors. As shown in figure, two pre-cooling and liquefaction cycles have been taken to bring gas to  $-162^{\circ}\text{C}$ . Propane is used for pre-cooling of the gas and nitrogen is used for liquefaction.

For propane and nitrogen heat exchangers and compressor are developed. In this figure the process (feed) gas is send to pre-cooling cycle and then to liquefaction section to plate heat exchanger.



**Figure 18:** LNG liquefaction process

The objective behind developing the heat exchangers is to find out the surface area required for heat transfer and the dimensions of the heat exchanger. The shell and tube heat exchanger is developed for pre-cooling purpose and the plate heat exchanger for liquefaction. The fundamentals remain same here.

- First assume the flowrate of the gas to be process.
- For finding out the heat load, heat to be taken out can be calculated using the enthalpies at desired temperature and the flowrate.

$$Q = (H_1 - H_2) V$$

- So, the required flowrate of refrigerant can be found out by

$$W = Q / (C_p \Delta T)$$

- Calculating the log mean temperature difference of the fluids, heat transfer area can be calculated as

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

The power required for circulating the refrigerant in the refrigeration cycle is calculated for two compressors, one for propane cycle and other for nitrogen.

### 8.1) Heat exchanger for pre-cooling (using propane): Shell and tube heat exchanger

Considering the process gas flowrate as 400 MMscfd being supplied at plant.

$$400 \text{ MMscfd} = 11.328 \text{ MMscmd}$$

$$\text{Volumetric flowrate} = 131.11 \text{ m}^3/\text{sec.}$$

$$131.11 * 1.3686 = 179.43 \text{ Kg/sec} \quad = V \dots\dots\dots$$

$$\text{As density taken as} \quad = 1.3686 \text{ Kg/m}^3$$

$$\text{Enthalpy of methane at } 10^0 \text{ C} \quad = -35.277 \text{ KJ/Kg}$$

$$\text{Enthalpy of methane at } -15^0 \text{ C} \quad = -89.891 \text{ KJ/Kg}$$

$$\begin{aligned} \text{Heat load of the heat exchanger } Q &= (H_1 - H_2) V = (89.891 - 35.277) 179.43 \\ &= 9799.39 \text{ KJ/sec} \end{aligned}$$

$$\begin{aligned} \text{Propane flowrate required} \quad W &= Q / (C_p * \Delta T) \\ &= 9799.39 * 10^3 / (2257.1 * 20) \\ &= 217.079 \text{ Kg/sec} \end{aligned}$$

$$\text{Assuming overall heat transfer coefficient} \quad U = 450 \text{ W/m}^2\text{K}$$

Temperatures assumed = inlet of gas at  $10^0$  and leaving at  $-15^0$  C  
Inlet of propane  $-40^0$  and leaving at  $-20^0$  C

$$\begin{aligned} \text{Log mean temperature difference} \quad \Delta T_{lm} &= (\Delta T_1 - \Delta T_2) / \ln \{ \Delta T_1 / \Delta T_2 \} \\ &= (50 - 5) / \ln \{ 50 / 5 \} \\ &= 19.54^0 \text{ C} \end{aligned}$$

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

*Production of LNG from natural gas*

$$\begin{aligned} A &= Q/(U*\Delta T_{lm}) \\ &= 9799.39*10^3/(450*19.54) \\ &= 1114.45 \text{ m}^2 \end{aligned}$$

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

$L$  = length of pipe

$N$  = no. of tubes

$$\begin{aligned} N &= A/(\Pi*d_o*L) \\ &= 1114.45/(\Pi*0.03*6.9) \\ &= 1714 \end{aligned}$$

Taking tube inner dia. 0.024 m and thickness 3mm

Mass flowrate of gas  $G_h = m/(n*A_i)$

$$\begin{aligned} &= 179.43/(1714*\{\Pi/4*(0.024)^2\}) \\ &= 231.4 \text{ Kg/m}^2\text{sec} \end{aligned}$$

Reynolds number  $Re = d*G/\mu$

$$\begin{aligned} &= 0.024*231.4/1.027*10^{-5} \\ &= 540759.5 \end{aligned}$$

Sieder-Tate equation

$$\begin{aligned} h_i D/k &= 0.023*(Re)^{0.8}*(Pr)^{0.33} \\ h_i &= (0.023*0.03281/0.024)*(540759.5)^{0.8}*(2209*1.027*10^{-5}/0.03281)^{0.33} \\ &= 1072.56 \text{ W/m}^2\text{k} \end{aligned}$$

$h_i$  = heat transfer coefficient inside tube,  $\text{W/m}^2\text{k}$

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

$$\begin{aligned} &= 0.03(1714/0.03281)^{1/2.207} \\ &= 4.119 \text{ m} \end{aligned}$$

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

Shell diameter  $Ds = 4.119*1.1 = 4.53 \text{ m}$

Assume baffle spacing

$$\begin{aligned} &= Ds/26 \\ &= 4.53/26 \\ &= 0.1712 \text{ m} \end{aligned}$$

*Production of LNG from natural gas*

$$\begin{aligned}\text{Shell side flow area} &= D_s * B * C / Pt \\ &= 4.11 * 0.1712 * 0.006 / 0.036 \\ &= 0.1173 \text{ m}^2\end{aligned}$$

$$\begin{aligned}\text{Mass flowrate of propane} &G_c = 217.079 / 0.1173 \\ &= 1850.5 \text{ Kg/m}^2\text{sec}\end{aligned}$$

$$\begin{aligned}\text{Equivalent diameter of square pitch arrangement} &De = 4 \{ Pt^2 - \Pi / 4 d_o^2 \} / \Pi d_o \\ &= 4 \{ 0.036^2 - \Pi / 4 * 0.03^2 \} / \Pi * 0.03 \\ &= 0.025 \text{ m}\end{aligned}$$

$$\begin{aligned}\text{Reynolds number} &Re = De * G_c / \mu \\ &= 0.025 * 1850.5 / 1.3 * 10^{-5} \\ &= 3558657.74 \\ h_o De / k &= 0.36 (Re)^{0.55} (Pr)^{0.4} \\ h_o &= 0.36 (3558657.74)^{0.55} (2257.1 * 1.3 * 10^{-5} / 0.015198)^{0.4} \\ &= 1212.7 \text{ W/m}^2\text{k}\end{aligned}$$

$h_o$  = heat transfer coefficient outside tube,  $\text{W/m}^2\text{k}$

Overall heat transfer coefficient

$$\begin{aligned}1/U_o &= (1/h_i) * (d_o/d_i) + (1/h_o) \\ &= (1/1072.56) * (0.03/0.024) + (1/1212.7) \\ &= 489.65 \text{ W/m}^2\text{k}\end{aligned}$$

$$\begin{aligned}\text{Heat transfer area required} &= (9799.39 * 10^3) / \{ 489.65 * 19.54 \} \\ &= 1024.2 \text{ m}^2\end{aligned}$$

$$\begin{aligned}\text{Percentage excess area} &= (1114.45 - 1024.2) / 1024.2 \\ &= 8.8\%\end{aligned}$$

So, the design can be accepted.

**Propane compressor:**

The propane is to be compressed from 147 psia to 588 psia at 50<sup>0</sup> F.

Assuming

$$Z_1, Z_2 = 1.0$$

$$K = 0.075/0.066 = 1.136$$

$$R = 588/147 = 4$$

Molecular weight

$$M = (28.97)y_g = 28.97 * 1.55$$

$$= 44.9 \text{ lbm/lb-mole} \quad \text{where } y_g = \text{specific gravity}$$

$$\text{Flow capacity} = 1264.91 \text{ MMscfd} = 1264.91 * 10^6 / (24 * 60)$$

$$= 878.4 * 10^3 \text{ cfm}$$

$q = 878.4 * 10^3$  cfm and polytropic efficiency  $\eta = 0.75$

$$k-1/(k\eta) = 1.136-1/1.136*0.75 = 0.159$$

The applicable actual head

$$h_p = (1545)(460+T) / \{(M)*(k-1/(k\eta))\} (3.333^{0.159} - 1)$$

$$= 23248.18 \text{ ft-lbf/lbm}$$

Gas density at suction conditions

$$\rho = Pm/(Z_1RT)$$

$$= 147 * 44.9 / (1 * 1.732 * 510)$$

$$= 1.2 \text{ lbm/ft}^3$$

Mass flowrate

$$M = q\rho = 12649 * 10^6 / 1440 \text{ (ft}^3/\text{day)} / (\text{min/day}) * 1.2 \text{ lbm/ft}^3$$

$$= 1.05 * 10^6 \text{ lbm/min}$$

Now,

$$\text{(Gas horsepower) GHP} = m * h_p / 33000 * \eta$$

$$= 1.05 * 10^6 * 23248.18 / 33000 * 0.75$$

$$= 986278.78 \text{ hp}$$

Considering HPLm=30 hp and HPLh =1% mechanical and hydraulic horsepower losses.

$$\text{BHP} = 1.01 * \text{GHP} + \text{HPLm}$$

$$= 1.01 * 986278.78 + 30$$

$$= 996201 \text{ hp}$$

**8.2) Heat exchanger for Liquefaction (using Nitrogen): Plate heat exchanger  
(Dimensions by Chart industries)**

$$V=179.43 \text{ Kg/sec}$$

$$\text{Enthalpy of methane at } -15^{\circ} \text{ C} = -89.891 \text{ KJ/Kg}$$

$$\text{Enthalpy of methane at } -162^{\circ} \text{ C} = -916.2 \text{ KJ/Kg}$$

$$\begin{aligned} \text{Heat load of the heat exchanger } Q &= (H_1 - H_2) V = (916.2 - 89.891) 179.43 \\ &= 147618.67 \text{ KJ/sec} \end{aligned}$$

$$\begin{aligned} \text{Nitrogen flowrate required } W &= Q / (C_p \Delta T) \\ &= 147618.67 \cdot 10^3 / (2201.9 \cdot 20) \\ &= 3352.07 \text{ Kg/sec} \end{aligned}$$

Assuming overall heat transfer coefficient  $U = 1000 \text{ W/m}^2\text{K}$

Temperatures assumed = inlet of gas at  $-15^{\circ}$  and leaving at  $-162^{\circ} \text{ C}$   
Inlet of nitrogen  $-190^{\circ}$  and leaving at  $-170^{\circ} \text{ C}$

$$\begin{aligned} \text{Log mean temperature difference } \Delta T_{lm} &= (\Delta T_1 - \Delta T_2) / \ln \{ \Delta T_1 / \Delta T_2 \} \\ &= (175 - 8) / \ln \{ 175 / 8 \} \\ &= 54.12^{\circ} \text{ C} \end{aligned}$$

Heat transfer area can be calculated as:

$$Q = U \cdot A \cdot F_q \cdot \Delta T_{lm} \quad A_T = Q / (U \cdot F_q \cdot \Delta T_{lm})$$

Assume  $\Delta T_{lm}$  correction factor = 0.95

$$\begin{aligned} A_T &= 1476118.67 \cdot 10^3 / (1000 \cdot 0.95 \cdot 54.12) \\ &= 2871.17 \text{ m}^2 \end{aligned}$$

Taking length of plate = 7620 mm

Width of plate = 1520 mm

$$\text{Projected heat transfer area of plate } A_p = 7.62 \cdot 1.52 = 11.43 \text{ m}^2$$

*Production of LNG from natural gas*

Number of plates  $N = 2871.17 / 11.43 = 252$

For looped flow arrangement, number of streams  $n_h = (N+1)/2$   
 $= (252+1)/2 = 126$

There will be 126 streams of gas fluid and 127 streams of nitrogen.

Gas flowrate per stream  $m_h = 179.43/126 = 1.42 \text{ Kg/sec}$

Cross sectional area of flow between the two plates  $= 1.52 * 0.006$   
 $= 9.12 * 10^3 \text{ m}^2$

Mass flowrate per unit area of gas stream  $Gh = m_h/A$   
 $= 1.42/9.12 * 10^3$   
 $= 155.7 \text{ Kg/m}^2\text{sec}$

Width between channels  $= 0.006 \text{ m}$

Equivalent diameter of flow area between two plates  $d_e = 2 * 0.006 = 0.012 \text{ m}$

Reynolds number for gas  $Re = Gh * d_e / \mu$   
 $= 155.7 * 0.012 / 1.027 * 10^{-5}$   
 $= 181927.9$

Prandlt number  $Pr = Cp * \mu / k$   
 $= 2209 * 1.027 * 10^{-5} / 0.03281$   
 $= 0.691$

Reynolds number is greater than 1000, so for  $Jh = 0.178(Re)^{-0.24}$   
 $= 0.0097$   
 $Jh = (St)(Pr)^{0.667} (\mu/\mu_w)$   $(\mu/\mu_w) = 1$   
 $St = Jh / (Pr)^{0.667}$   
 $= 0.0097 / (0.691)^{0.667}$   
 $= 0.012$   
 $St = h_h / (Cp * Gh)$   
 $h_h = St * Cp * Gh$   
 $= 0.012 * 2209 * 155.7$   
 $= 4103.96 \text{ W/m}^2\text{sec}$

Mass flowrate per stream of nitrogen  $Wc = W/n$   
 $= 3352.07 / 127$

*Production of LNG from natural gas*

$$\begin{aligned}
 &= 26.39 \text{ Kg/sec} \\
 \text{Mass flowrate per unit area of nitrogen} \quad G_c &= 26.39/9.12*10^{-3} \\
 &= 2893.6 \text{ Kg/m}^2\text{sec} \\
 \text{Reynolds number for nitrogen} \quad Re &= G_c*d_c/\mu \\
 &= 2893.6*0.012/1.657*10^{-5} \\
 &= 2095546.16 \\
 \text{Prandlt number for nitrogen} \quad Pr &= (2201.9*1.65*10^{-5})/0.024 \\
 &= 1.52
 \end{aligned}$$

Reynolds number is greater than 1000, so for  $J_h = 0.178(Re)^{-0.24}$

$$\begin{aligned}
 &= 0.0054 \\
 J_h &= (St)(Pr)^{0.667}(\mu/\mu_w) \quad (\mu/\mu_w)=1 \\
 St &= J_h/(Pr)^{0.667} \\
 &= 0.0054/(1.52)^{0.667} \\
 &= 0.004 \\
 St &= h_c/(C_p*G_c) \\
 h_c &= St*C_p*G_c \\
 &= 0.004*2201.9*2893.6 \\
 &= 25485.67 \text{ W/m}^2\text{sec}
 \end{aligned}$$

Overall heat transfer coefficient can be calculated as

$$1/U = 1/h_h + 1/h_c + x/k$$

where, x = thickness of plate, mm

k = conductivity of plate, W/Km

$$1/U = 1/4103.96 + 1/25485.67 + 0.01/17$$

$$U = 1148.01 \text{ W/m}^2\text{sec}$$

Heat transfer area can be calculated as

$$\begin{aligned}
 A_T &= Q/(U*F_q*\Delta T_{lm}) \\
 &= 147618.67*10^3/(54.12*0.95*1148.01) \\
 &= 2501.003\text{m}^2
 \end{aligned}$$

$$\text{Percentage excess area} = (2871.17 - 2501.003)/2501.003$$

$$= 14.8 \%$$

**Heat exchanger for Liquefaction (using Nitrogen).**

**(Dimensions assumed)**

$$V = 179.43 \text{ Kg/sec}$$

$$\text{Enthalpy of methane at } -15^{\circ} \text{ C} = -89.891 \text{ KJ/Kg}$$

$$\text{Enthalpy of methane at } -162^{\circ} \text{ C} = -916.2 \text{ KJ/Kg}$$

$$\begin{aligned} \text{Heat load of the heat exchanger } Q &= (H_1 - H_2) V = (916.2 - 89.891) 179.43 \\ &= 147618.67 \text{ KJ/sec} \end{aligned}$$

$$\begin{aligned} \text{Nitrogen flowrate required} \quad W &= Q / (C_p \cdot \Delta T) \\ &= 147618.67 \cdot 10^3 / (2201.9 \cdot 20) \\ &= 3352.07 \text{ Kg/sec} \end{aligned}$$

$$\text{Assuming overall heat transfer coefficient} \quad U = 1000 \text{ W/m}^2\text{K}$$

Temperatures assumed = inlet of gas at  $-15^{\circ}$  and leaving at  $-162^{\circ}$  C  
Inlet of nitrogen  $-190^{\circ}$  and leaving at  $-170^{\circ}$  C

$$\begin{aligned} \text{Log mean temperature difference} \quad \Delta T_{lm} &= (\Delta T_1 - \Delta T_2) / \ln \{ \Delta T_1 / \Delta T_2 \} \\ &= (175 - 8) / \ln \{ 175 / 8 \} \\ &= 54.12^{\circ} \text{ C} \end{aligned}$$

Heat transfer area can be calculated as:

$$Q = U \cdot A \cdot F_q \cdot \Delta T_{lm} \quad A_T = Q / (U \cdot F_q \cdot \Delta T_{lm})$$

Assume  $\Delta T_{lm}$  correction factor = 0.95

$$\begin{aligned} A_T &= 1476118.67 \cdot 10^3 / (1000 \cdot 0.95 \cdot 54.12) \\ &= 2871.17 \text{ m}^2 \end{aligned}$$

Taking length of plate = 8000 mm

Width of plate = 1520 mm

$$\text{Projected heat transfer area of plate} \quad A_p = 8 \cdot 1.52 = 12.16 \text{ m}^2$$

*Production of LNG from natural gas*

Number of plates  $N = 2871.17 / 12.16 = 237$

For looped flow arrangement, number of streams  $n_h = (N+1)/2$   
 $= (238+1)/2 = 119$

There will be 119 streams of gas fluid and 119 streams of nitrogen.

Gas flowrate per stream  $m_h = 179.43 / 119 = 1.5 \text{ Kg/sec}$

Cross sectional area of flow between the two plates  $= 1.52 * 0.006$   
 $= 9.12 * 10^3 \text{ m}^2$

Mass flowrate per unit area of gas stream  $Gh = m_h / A$   
 $= 1.5 / 9.12 * 10^3$   
 $= 164.47 \text{ Kg/m}^2\text{sec}$

Width between channels  $= 0.006 \text{ m}$

Equivalent diameter of flow area between two plates  $d_e = 2 * 0.006 = 0.012 \text{ m}$

Reynolds number for gas  $Re = Gh * d_e / \mu$   
 $= 164.47 * 0.012 / 1.027 * 10^{-5}$   
 $= 192175.26$

Prandlt number  $Pr = Cp * \mu / k$   
 $= 2209 * 1.027 * 10^{-5} / 0.03281$   
 $= 0.691$

Reynolds number is greater than 1000, so for  $Jh = 0.178(Re)^{-0.24}$   
 $= 0.0096$   
 $Jh = (St)(Pr)^{0.667} (\mu/\mu_w)$   $(\mu/\mu_w) = 1$   
 $St = Jh / (Pr)^{0.667}$   
 $= 0.0096 / (0.691)^{0.667}$   
 $= 0.012$

$St = h_h / (Cp * Gh)$   
 $h_h = St * Cp * Gh$   
 $= 0.012 * 2209 * 164.47$   
 $= 4359.77 \text{ W/m}^2\text{sec}$

Mass flowrate per stream of nitrogen  $Wc = W/n$   
 $= 3352.07 / 119$

*Production of LNG from natural gas*

$$\begin{aligned}
 &= 28.16 \text{ Kg/sec} \\
 \text{Mass flowrate per unit area of nitrogen} \quad G_c &= 28.16/9.12 \cdot 10^{-3} \\
 &= 3087.71 \text{ Kg/m}^2\text{sec} \\
 \text{Reynolds number for nitrogen} \quad Re &= G_c \cdot d_c / \mu \\
 &= 3087.71 \cdot 0.012 / 1.657 \cdot 10^{-5} \\
 &= 2236120.7 \\
 \text{Prandlt number for nitrogen} \quad Pr &= (2201.9 \cdot 1.65 \cdot 10^{-5}) / 0.024 \\
 &= 1.52 \\
 \text{Reynolds number is greater than 1000, so for } J_h &= 0.178(Re)^{-0.24} \\
 &= 0.0053 \\
 J_h &= (St)(Pr)^{0.667} (\mu/\mu_w) \quad (\mu/\mu_w)=1 \\
 St &= J_h / (Pr)^{0.667} \\
 &= 0.0053 / (1.52)^{0.667} \\
 &= 0.004 \\
 St &= h_c / (C_p \cdot G_c) \\
 h_c &= St \cdot C_p \cdot G_c \\
 &= 0.004 \cdot 2201.9 \cdot 3087.71 \\
 &= 27195.3 \text{ W/m}^2\text{sec}
 \end{aligned}$$

Overall heat transfer coefficient can be calculated as

$$\begin{aligned}
 1/U &= 1/h_h + 1/h_c + x/k \\
 \text{where, } x &= \text{thickness of plate, mm} \\
 k &= \text{conductivity of plate, W/Km} \\
 1/U &= 1/4359.77 + 1/27195.3 + 0.01/17 \\
 U &= 1170.86 \text{ W/m}^2\text{sec}
 \end{aligned}$$

Heat transfer area can be calculated as

$$\begin{aligned}
 A_T &= Q / (U \cdot F_q \cdot \Delta T_{lm}) \\
 &= 147618.67 \cdot 10^3 / (54.12 \cdot 0.95 \cdot 1170.86) \\
 &= 2452.19 \text{ m}^2
 \end{aligned}$$

$$\begin{aligned}
 \text{Percentage excess area} &= (2871.17 - 2452.19) / 2452.19 \\
 &= 17.08 \%
 \end{aligned}$$

**Comparison between plate type heat exchanger developed by taking Chart industry and Assumed dimensions:**

The table below shows that the plate heat exchanger developed by Chart industry and by the assumed dimensions shows the following difference:

Parameters	By Chart Dimension	By assumed dimension
Length	7620 mm	8000 mm
Width	1520 mm	1520 mm
Number of plates	253	238
Height	4.048 m	3.808
Reynolds number	Less	More
Excess area provided	14.8 %	17.08 %

By extending the length of the plate by 380 mm, the excess area provided would increase by 15.4% as compared to chart industry, which will take care of the fouling in the plates.

Advantages of using plate heat exchanger for liquefaction:

- Comparatively the surface area provided by plate heat exchanger per unit volume is more, than shell and tube heat exchanger.
- The Reynolds number in plate heat exchanger is more, so the heat transfer would be more.
- The temperature of operation is from  $-269^{\circ}\text{C}$  to  $200^{\circ}\text{C}$ .
- Cleaning of the plates is easier.

**Nitrogen compressor:**

The nitrogen is to be compressed from 147 psia to 588 psia at 50<sup>0</sup> F.

Assuming  $Z_1, Z_2 = 1.0$

$$K = 0.029/0.02 = 1.45$$

$$R = 588/147 = 4$$

$$\text{Molecular weight } M = (28.97)y_g = 28.97 * 10 * 967$$

$$= 28.013 \text{ lbm/lb-mole} \quad \text{where } y_g = \text{specific gravity}$$

$$\text{Flow capacity} = 18643.58 \text{ MMscfd} = 18643.58 * 10^6 / (24 * 60)$$

$$= 12.94 * 10^6 \text{ cfm}$$

$$q = 878.4 * 10^3 \text{ cfm and polytropic efficiency } \eta = 0.75$$

$$k - 1 / (k\eta) = 1.445 - 1 / 1.45 * 0.75 = 0.413$$

$$\text{The applicable actual head } h_p = (1545)(460 + T) / \{(M) * (k - 1 / (k\eta))\} (3.333^{0.413} - 1)$$

$$= 19077.41 \text{ ft-lbf/lbm}$$

Gas density at suction conditions

$$\rho = P_m / (Z_1 RT)$$

$$= 147 * 28.013 / (1 * 1.732 * 222)$$

$$= 1.72 \text{ lbm/ft}^3$$

Mass flowrate

$$m = q\rho = 18643.58 * 10^6 / 1440 \text{ (ft}^3/\text{day)} / (\text{min/day}) * 1.72 \text{ lbm/ft}^3$$

$$= 22.268 * 10^6 \text{ lbm/min}$$

Now,

$$\text{GHP} = m * h_p / 33000 * \eta$$

$$= 22.268 * 10^6 * 19077.41 / 33000 * 0.75$$

$$= 17.16 * 10^6 \text{ hp}$$

Considering HPL<sub>m</sub> = 30 hp and HPL<sub>h</sub> = 1% mechanical and hydraulic horsepower losses.

$$\text{BHP} = 1.01 * \text{GHP} + \text{HPL}_m$$

$$= 1.01 * 17.16 * 10^6 + 30$$

$$= 17.33 * 10^6 \text{ hp}$$

**Compressor driver:**

**Frame-7 or Frame-9**

Its simplicity and size make it an extremely flexible component in plant layout and an easy, low-cost addition of power when phase capacity expansion is needed. The units are ideal for plants that require high efficiency and shaft speed for direct coupling to the generator.

The above driver can give the better operation and efficiency, as most of the LNG plants are looking for the high capacity drivers so as to increase the production capacity and to ultimately reduce the cost.

The above plant configuration can produce LNG:

With 400 MMscfd of gas input and 10 percent of this to other facilities can give the production upto 2.9MMTPA of LNG.

$$1 \text{ MMTPA} = 3.5 \text{ MMscmd of gas}$$

**Salient features of the plant:**

1. The plant can produce more LNG with additional circuits.
2. Assuming certain parameters develops the above heat exchanger and compressors.
3. 400 MMscfd of gas can be produced from above refrigeration process.
4. Condenser and expansion valves are not developed due to complicity.
5. As nitrogen-boiling temperature is very low it is suitable for liquefaction cycle.
6. With use of proper insulation like perlite, the thermal losses can be prevented and ultimately processing cost for condensers can be reduced.
7. With the additional facilities like NGL extraction plant and BOG re-compressor plant cost can be reduced.

## **Conclusion**

In the growing market of LNG, the liquefaction processes are coming in new trend for reduction in cost and large capacity trains. Through effective improvement and technical reviews some conventional processes are modified to give rise to large LNG trains and high production with lower cost.

- The APCI's C<sub>3</sub>MR expandable process with additional nitrogen cycle for liquefaction and subcooling, Shell's PMR process with two parallel process stream, Philips optimized cascade process with additional facility of liquefying boil off gas from the tank are some of the processes which can be recommended for upcoming LNG plants.
- Turboexpanders, high capacity drivers or high capacity electric motors can be suggested for the large equipment, like compressors.
- Vaporized gas from the loading operation of tanker or boil off gas can be re-liquefied to add to the production of LNG and to reduce vapor loss through flare.

### **Development of the LNG liquefaction plant equipments:**

Development of heat exchangers for the refrigeration cycle is the key for the LNG production. The use of shell & tube and plate heat exchanger for pre-cooling and liquefaction cycles respectively. The plate heat exchanger is found effective as it brings down the temperature significantly (-160 °C), to liquefy the gas to LNG. Nitrogen as its boiling point is much lesser, is used as refrigerant for liquefaction cycle. Increasing the length of the plates in plate heat exchanger the effectiveness can be increased.

## **Recommendation**

The plate heat exchanger developed using Chart Industry dimensions and the assumed present standard value gives the following results:

<b>Parameters</b>	<b>By Chart Dimension</b>	<b>By assumed dimension</b>
Length	7620 mm	8000 mm
Width	1520 mm	1520 mm
Number of plates	253	238
Height	4.048 m	3.808
Excess area provided	14.8 %	17.08 %

By extending the length of the plate by 380 mm, the excess area provided would increase by 15.4% as compared to chart industry, which will take care of the fouling in the plates. Also, Comparatively the surface area provided by plate heat exchanger per unit volume is more, than shell and tube heat exchanger, so the higher heat load can be taken care of by plate heat exchanger.

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