Study of Operation of Cost Items in LNG Plant

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CERTIFICATION OF APPROVAL

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A project dissertation submitted to the
Chemical Engineering Programme
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Approved by,

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TRONOH, PERAK

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CERTIFICATION OF ORIGINALITY

This is to certify that I am responsible for the work submitted in this project, that the original work is my own except as specified in the references and acknowledgements, and that the original work contained herein have not been undertaken or done by unspecified sources or persons.

NANCY HII ZHUN WEI

ABSTRACT

Due to the increasing demand for natural gas in the world today, transportation of natural gas from different parts of the world has become a necessity. Liquefying the natural gas provides a safer and cheaper alternative for transportation and also increases its storage capabilities. However, it has been accounting for the highest operating cost if compared to the other chain of the industry. Hence liquefaction process has been a key area that constantly in need for development so as to save cost and increase LNG plant capacity through production, which these are, also the objectives to this research.

The scopes of studies for this project cover the development of LNG technologies, the cost items that affects the overall cost consumption of the plant, the operating parameters in liquefaction process, equipments' efficiency and also the process line-up.

The methodology of this project covers two areas, which are research and stimulation. Research covers the study and collections of information regarding the LNG industry and its development, while stimulation that reflects the process flow of liquefaction will be done in HYSYS. Through HYSYS, comparison can be done from the varies schemes stimulated to obtain the most optimum process output.

The flowsheet selected for simulation will be utilising the technology of APCI C3MR as basis. For the study in the power consumption of the plant, the focus lies in the propane refrigerant system where its composition will be manipulated to obtain the temperature output and power required by the compressor. Based on the power obtained, the cost can be calculated to obtain the cheapest operating cost. Nevertheless, further studies and in depth understanding on the fundamentals of liquefaction process is still required in order to develop innovative methods to further increase the capacity, efficiency and consequently the production of LNG in a LNG plant. The enhancement method can be in terms of process and machinery integration advancements.

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CHAPTER 1
INTRODUCTION

1.1 Background of study

In January 2009, it will be 50 years since the world's first LNG tanker Methane Pioneer carried LNG from United States to the United Kingdom. This started an LNG export industry that has been growing ever since.

Natural gas is used across the residential, commercial and industry sectors. Industry is the biggest consumer of natural gas; mainly use as a heat source to manufacture goods. It is also used as an ingredient in fertilizer, photographic film, ink, glue, paint, plastics, laundry detergent, and insect repellents, besides providing crucial chemicals for the making of synthetic rubber and man-made fibers like nylon. The residential/commercial sector (homes and business) used natural gas for heating as it is clean burning. Natural gas is also used to make electricity. As the demand of electricity increase, the energy industry people believe that natural gas will play a bigger role in the production of electricity in the future.
Natural gas power plant is cleaner than coal plant and can be brought on line very quickly. It also produces electricity more efficient with fewer emissions. It is also sometimes used as transportation fuel in any vehicle with regular internal combustion engine, but must be fitted with a special carburettor and fuel tank. It burns cleaner than gasoline, costs less and has higher octane rating. The higher the octane, the greater the power.

Natural gas is much more preferable among all the fossil fuels such as coal and petroleum as it is the most environmentally friendly fuel. It produces less sulphur, carbon and nitrogen and also emits little ash particulate into the air when it is burned.

The popularity and attractiveness of natural gas causes its demand to have grown appreciably, especially for use as fuel for power generation in modern combined-cycle gas turbine plants. Furthermore, it is also mainly due to the concern about the deleterious affects that emissions, from the burning of hydrocarbons, have on the environment. Natural gas produces 50% fewer emissions than that of oil and 85% less compared to coal as it is rich with methane which is a light hydrocarbon with less emission when burned with comparison to heavy hydrocarbons. With an increasing concern for environmental issues and global warming, countries and companies are turning to natural gas as an energy source. Besides, the Clean Air Act Amendment (CAAA) and the Energy Policy Act of 1992 are forcing companies with large vehicle fleets operating in areas with ozone problems, railroads, and some stationery unit operators to convert to cleaner burning fuel. Thus from figure 1, we can foresee a steady climb of demand as the climate legislation gets stricter.

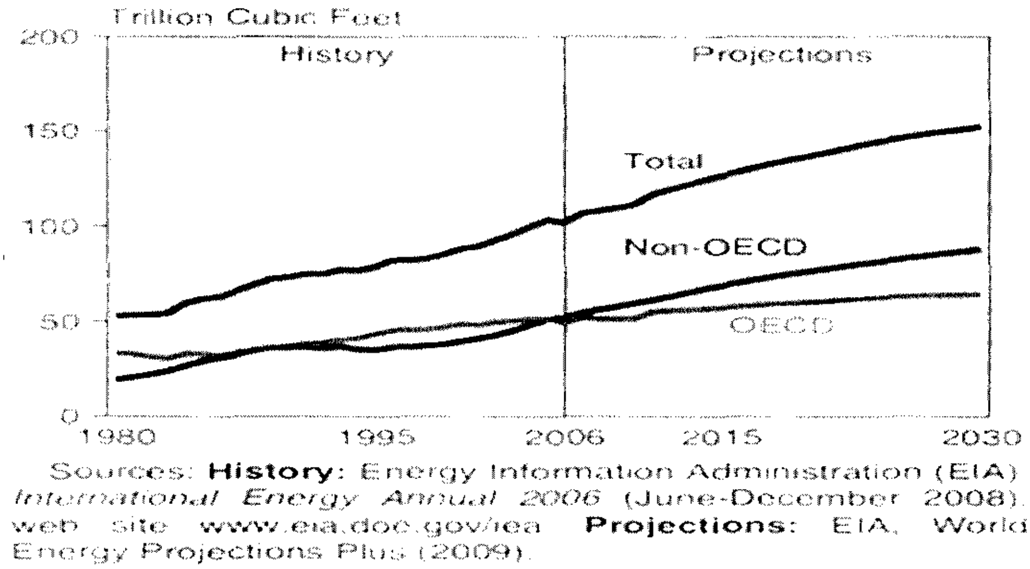


Figure 1: World Natural Gas Consumption, 1980-2030 [1]

Note: *OECD (Organisation for Economic Co-operation and Development) brings together the governments of countries committed to democracy and the market economy from around the world.*

1.1.1 LNG business chain and its importance

An LNG projects represents a 'chain' of capital-intensive investments, consisting five links – field development, pipeline to coast, the liquefaction facility, tanker transportation and the receipt/regasification terminal. Nevertheless, the liquefaction process has been accounting for up to 50% of total project cost of a liquefaction plant as shown in Figure 2. Thus, the liquefier has been the key area where a process engineer can make the largest cost savings and influence the project viability.

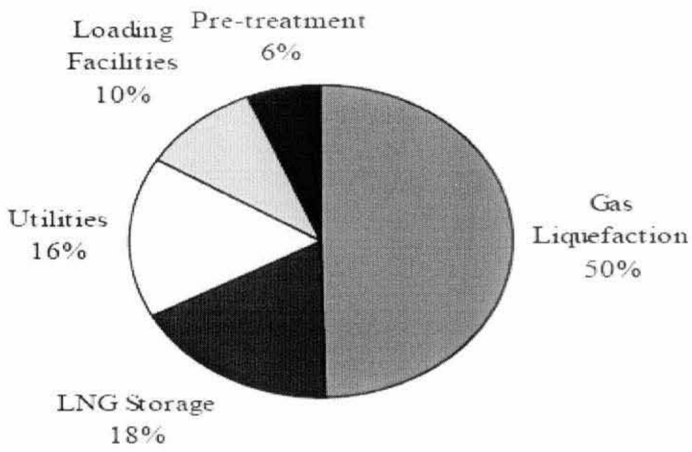


Figure 2: Typical breakdown of liquefaction plant capital costs [2]

The low density of natural gas makes it more costly to contain and transport then any other energy sources. Hence, when the transportation by pipeline is overly expensive, natural gas must be either liquefied or converted to high value liquid products. Liquefaction has advantages over chemical conversion in that LNG has a heating value about 40% greater than liquid fuels derived from chemical conversion of natural gas. Besides, by temporarily converting natural gas to liquid form at atmospheric pressure, it takes up about $1/600^{th}$ volume of natural gas in gaseous state, easing storage and transportation. However, special processing and containment requirements to transport gas as LNG come at a significant cost.

In order to meet the rapid growth in LNG demand, this drives the urge for continual improvement in liquefaction technology development to push the limits of capacity and functionality of LNG production in plants. These developments have resulted in a wide portfolio of liquefaction technologies and train sizes as we can see in Figure 3 and 4.

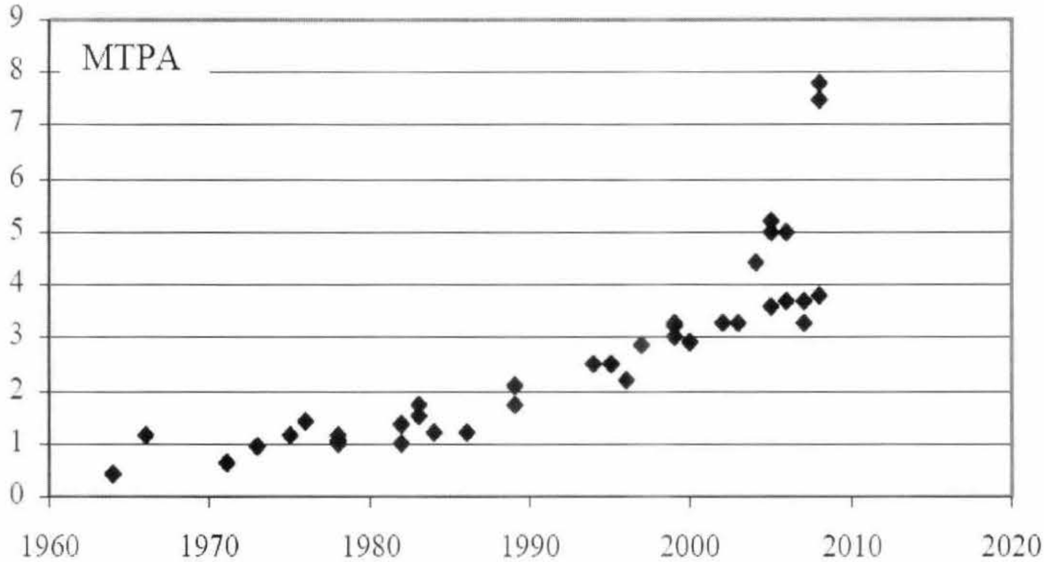


Figure 3: LNG train size growth [3]

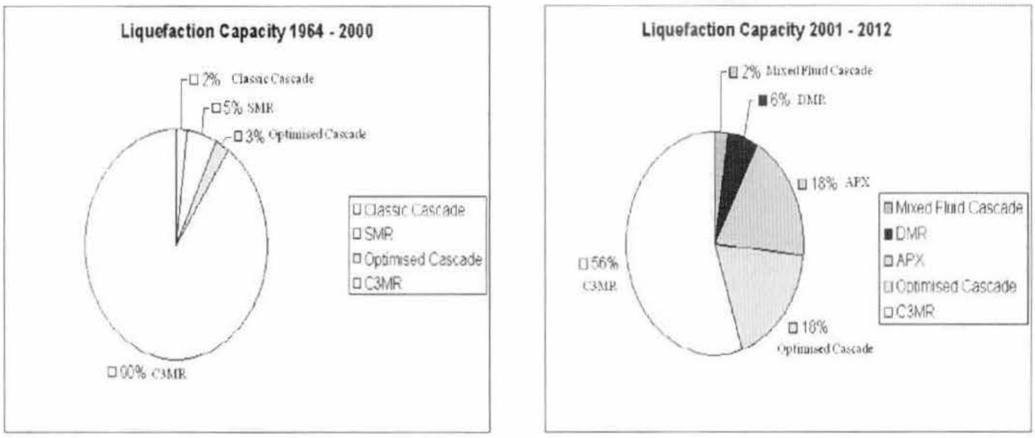


Figure 4: Evolution of LNG technologies; worldwide applied natural gas liquefaction technologies [4]

One of the criterions for the selection of liquefaction process is the capacity requirements. Capacity is the maximum amount of LNG that can be produced in a year. Designing a large plant and running it far below the capacity rates is a waste of investments and potentially could result in greater maintenance issues. Economic of scale means maximising profit based on fixed capital investment. Hence in order to fully take advantage of their economics of scale, production must be maintained near capacity. Since production level is based on what the market will support, if the demand goes down, so would production. Likewise, if production of LNG is greater than demand, the sale price will weaken and production rates must be decreased which leads to further waste of capital investments. Therefore, the selection of technical method is concerned with capacity and stability but it is more concerned with maximising profit based on market demand.

For many years, the propane pre-cooled mixed refrigerant (C3MR) process has remained the dominant liquefaction cycle in the LNG industry. This is due to the versatility of this cycle that makes it well-suited to accommodate this ever changing industry. Among the technology advancements made in meeting the industry's needs are:

- Enhancement in Air Products' main cryogenic heat exchangers (MCHEs) have been coupled with advancements in refrigerant compressors and drivers to significantly increase C3MR train capacity beyond 5 million tons per annum (MTA) using a single MCHE.
- Use of alternative pre-cooling fluids with the same cycle and equipment configuration as for C3MR allow for productive utilisation of this reliable and efficient process in colder climates.
- Efficient integration of NGL/LPG recovery with the liquefaction process plays a key role in achieving lower heating value LNG requirements for a variety feed conditions.

Though the mixed refrigerant processes rank the highest in equipment cost, it has the lowest operating cost. It consists of one big heat exchanger tower, massive compressors and propane chillers which raise the equipment cost and reduces the operating cost. The operating cost in mixed refrigerant process is lower due to two reasons:

- Having large heat exchanger tower which leads to reduction in ambient heat loss since all the heat exchanging process takes place in the tower rather than in separate heat exchangers.
- Requires less work due to having one large compressor with a heat exchanging tower which is more efficient than having many compressors running for each loop. Having many compressors in each loop results in more frictional and head loss which requires more work by compressors.

1.2 Problem Statement

Based on the pioneering of liquefaction technologies available, the industry has been constantly looking for further improvements resulting on modern designs, which are larger, more efficient and cost effective. However, the challenge still lies in finding the best use of power from cost effective and reliable equipment to meet the requirements of the market served, collaborating with two issues that must be taken into consideration: wasted capital outlay and maintenance issues. Wasted capital should be avoided at all cost, regardless of the project, besides application of newer and yet to be proven technologies or equipments will lead to more maintenance which means loss of money and lowered production.

Nevertheless, the trend within the LNG industry is to push the limits of LNG production capacity of a single train through process improvements. In energy-intensive liquefaction process, machinery constitutes a major portion of the total capital cost. Hence, in order to have the strongest economic and environmental merits, it is crucial for the plant to run in maximum thermal efficiency. Higher thermal efficiency is a trade off between capital and lifecycle costs. Improving the thermal efficiency will reduce the power requirement, thus the overall lifecycle costs. Therefore, the liquefaction area which is the heart of an LNG plant lies the decision in the selection of technology and it is also where the operation of major equipment are. This provides the most flexibility in improving the thermal efficiency of the overall plant.

Thus, major technical developments have mainly been done on the three key elements in liquefaction:

- Compression needed in the refrigeration cycles
- The power to drive those cycles
- The heat exchanger technology and process line up as medium to remove heat from the natural gas feed

1.3 Objectives and scope of study

The objectives of this project include:

- Identify and study the operation of cost items in liquefaction process
- Provide innovative suggestions on how to save cost while increasing LNG production

In order to achieve the above objectives, research and analysis need to be carried out to collect all the regarding technical details on the LNG industry from available source of information. The scope of studies involves:

- The development of LNG liquefaction technologies
- The cost items involved in LNG plant
- The operating parameters in liquefaction process
- The enhancement in equipment efficiency
- Process line up to achieve highest thermal efficiency

CHAPTER 2
LITERATURE REVIEW

2.1 Background for LNG

Liquefied natural gas (LNG) was proven viable in 1917, when the first LNG plant went into operation in West Virginia. The first commercial liquefaction plant was built in Cleveland, Ohio in 1941. In January 1959, the world's first LNG tanker carried LNG cargo from Lake Charles, Louisiana to Canvey Island, United Kingdom. This event demonstrated that large quantities of LNG could be transported safely across the ocean.

In 1961, Britain signed a 15-year contract to take less than 1 million tonnes per annum (mtpa) from Algeria, commencing in 1965. The first liquefaction plant in the world was commissioned at Arzew in Algeria to supply this contract with gas production coming from huge gas reserves found in the Sahara. The following year the French signed a similar deal to buy LNG from Algeria.

Alaska's Kenai plant (which currently has a capacity of 1.3 mtpa) began LNG deliveries to Japan's Tokyo Gas and Tokyo Electric Power Company (Tepco) in 1969. In 1972, Brunei became Asia's first producer, bringing on stream an LNG plant at Lumut that now has a capacity of 6.5 mtpa and supplies Korea as well as Japan. Libya's plant at Marsa el Brega began deliveries to Spain in 1970. Italy was also supplied by Libya, marking the entry of a new producer and two new buyers into the ranks of LNG trade.

U.S. imports from Algeria were approved in 1972 with Boston's Distrigas committing to buy 50 million standard cubic feet per day (mmscfd) from the Skikda plant over a 20-year period.

1979 witnessed the first LNG contract expiration: the 15-year contract between Algeria and the UK came to an end. Deliveries from Algeria continued into the 1980s but were eventually terminated. During 1979, the market was shaken by disputes over pricing between the U.S. buyers and Sonatrach which eventually resulted in the termination of the contracts, retiring of six LNG carriers (three of which were subsequently scrapped) and the mothballing of two of the U.S.'s four LNG terminals.

However, demand for LNG in Asia continued to rise and Malaysia entered the LNG market in 1983 (contract volume originally 6 mtpa but subsequently increased to 7.5 mtpa), followed by Australia in 1989 (similarly with an initial contract volume of 6 mtpa which has now been increased to 7.5 mtpa).

Qatar became the second Middle Eastern LNG producer with the delivery of its first cargo of LNG from the Qatargas LNG plant in January 1997. More recently several plants have come on line: Trinidad (3 mtpa) started up in April 1999; Ras Laffan (6.6 mtpa) in May 1999; Nigeria (5.6 mtpa) in October 1999. In April 2000, Oman commenced production with a plant of design capacity of 6.6 mtpa delivering its first cargo to Korea.

2.2 Overview on LNG process flow

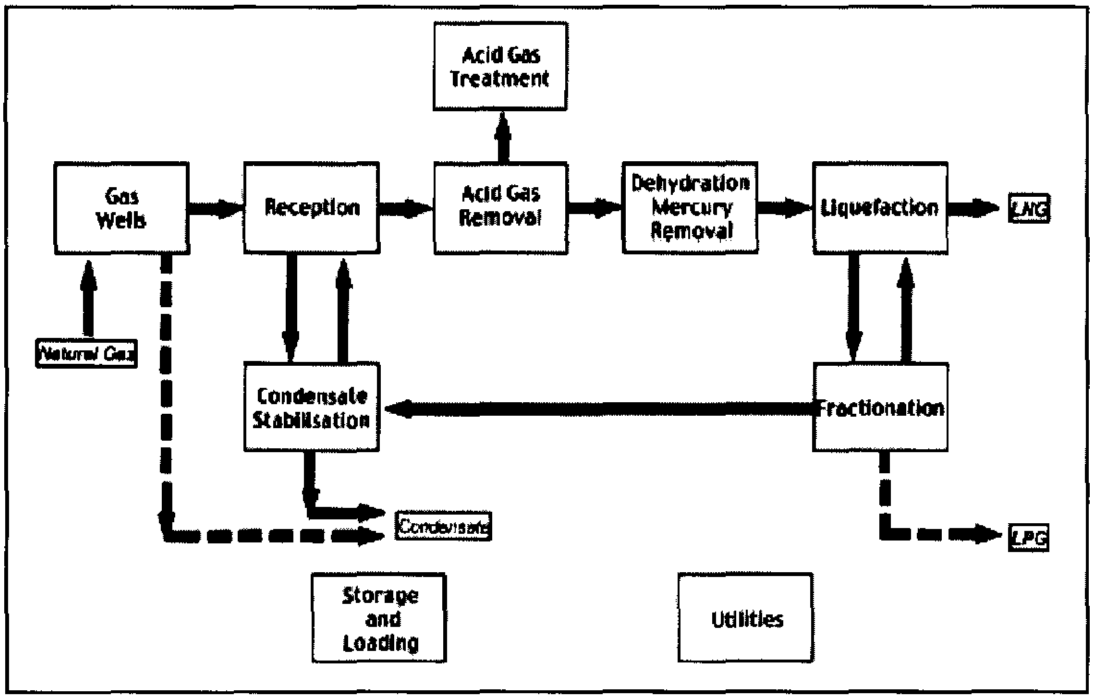


Figure 5: LNG block flow diagram [5]

Table 1: Natural Gas composition [6]

Component	Composition
Methane	70-90%
Ethane	0-20%
Propane	
Butane	
Carbon Dioxide	0-8%
Oxygen	0-0.2%
Nitrogen	0-5%
Hydrogen Sulfide	0-5%
Rare gases	trace

Figure 5 projected the block flow diagram for the liquefaction process of natural gas, while the table above shows the composition of each component available in natural gas. Generally, the liquefaction process can be divided into 3 main sections which will be explained below.

2.2.1 Pretreatment

Before natural gas can be liquefied, it must be treated. Treatment involves the removal of CO2, condensate, organic sulphur compounds, and Hg. This is done to avoid blockage in the liquefaction process, prevent damage done to the equipments, and to meet the heat content standard of natural gas which differs from one country to the other. Heat content of the natural gas increases with the increase of C3 to C4 in the composition of the gas. The pretreatment equipment specification is dependent upon the inlet contaminant concentrations.

The first pretreatment step process consists of four main stages. First, CO2 and H2S removal stage which is constructed to assure that CO2 would not exceed 50 ppm in the natural gas feed. If the composition of CO2 exceeded that limit it would freeze in the liquefaction process pipelines. There are two available methods to remove CO2 from the natural gas.

- The first method is using sulfinol, which used to be one of the famous methods for gas CO2 removal in most industrial applications but it started to vanish and few pretreatment plants use sulfinol any more. The reason is that sulfinol does not work well with rich natural gas (rich with heavy hydrocarbons) because it tends to attract or attach to heavy hydrocarbons and then it drags the heavies to the sulfinol pump or circulation which leads to allowing the heavy hydrocarbons to vent into the air. That usually leads to a decrease in the quality and heat content of the natural gas in the feed and according to the new EPA (Environmental Protection Agency) rules heavy hydrocarbons can not be vented to the atmosphere because it would lead to an increase in the air pollution level in the surroundings significantly.

- Another method to remove CO2 is using DEA (Diethyl Amine) but since DEA alone can not remove CO2 to a ppm (particle per million) level, it is activated by injecting a chemical called Piperzine which activates the DEA to MDEA (Methyl Diethyl Amine). MDEA is also used for this pretreatment stage versus sulfinol because MDEA is cheaper to install and it has less utility since it requires less rate of solvent circulation compared to sulfinol. Therefore, this pretreatment stage is named the amine wash section.

The second stage is dehydration. Water is removed from natural gas also to avoid freezing in the pipeline of the liquefaction process. The natural gas feed should be completely dry, even from the smallest traces of water molecules (95°F) ambient temperatures.

The third stage involves the removal of mercury usually by adsorption/reaction to form mercury sulphide in a high porosity catalyst, activated carbon, impregnated with sulphur. The catalyst is not regenerable, and once its capacity has been used, it will be disposed of and replaced by a new catalyst charge. Mercury, present even in trace quantities in natural gas feed will corrode aluminium, a material used in the main cryogenic heat exchanger rapidly under certain conditions of temperature and moisture.

After above stages, NGLs such as ethane, propane, butane and pentanes (heavy hydrocarbons) are removed and collected. In many cases, this step is done at the upstream of liquefaction unit using traditional gas processing technology. However, in other cases, NGLs recovery may be done as an integral step in liquefaction process. The NGLs collected are valuable products in their own right. It may be used as refrigerants for liquefaction process or may be reinjected into LNG streams at a later point to adjust the BTU content and flammability characteristics of LNG. Pentanes and other heavy hydrocarbons are generally exported as gasoline product. Butane and propane are often exported as separate products, used as refrigerant and/or as LPG. Ethane is often reinjected into LNG stream and may also be used as a refrigerant.

2.2.2 Refrigeration of gas until it liquefies

The refrigeration and liquefaction section is the key element of a LNG plant where it typically accounts for 30-40% of the capital cost of the overall plant. Liquefaction of natural gas involves the transfer of energy from hot stream of natural gas to cold stream of the refrigerant via LNG heat exchangers in order to change the phase of natural gas from vapour to liquid. The basic principle of cooling and liquefying the gas using refrigerant to cryogenic temperature approximately minus 160°C is to match the cooling/heating curve of the process gas and refrigerant as closely as possible, resulting in a more efficient thermodynamic process requiring less power per unit of LNG produced.

2.2.3 Movement of LNG to storage and ultimately into the tanker

After liquefaction process, LNG is pumped into a cryogenic storage tank. These tanks are typically double-walled, with an outer wall of reinforced concrete line with carbon steel and an inner wall of nickel steel. Between the two walls is insulation to prevent ambient air from warming the LNG. LNG is stores in these tanks until a tanker is available to take LNG to market. After an empty tanker docks at the berth, which is located as close to the storage tanks as possible, LNG is loaded into the tanker through insulated pipes that are attached to the tanker by rigid loading arms. Once the tanker is filled, the pipes are disconnected, the loading arm will swing away from the ship and the tanker is ready to sail.

2.3 Fundamentals of liquefaction process

2.3.1 Refrigeration Cycles

Based on Figure 6, a basic refrigeration cycle consists of two heat exchangers, a valve and a compressor. The refrigerant flows through evaporator where it is heated. The evaporator represents the cooling that a gas or liquid would receive from the refrigerant. From the evaporator, the refrigerant flows through a compressor to get the stream back to its design pressure. It also converts the stream from two phases to one phase. After the evaporator the refrigerant might be at or past its boiling point. After the compressor, the refrigerant flows through condenser to get to its bubble point. The refrigerant then flows through an expansion valve, after which it is cool enough to absorb the heat that is transferred in the evaporator.

The liquefaction plants use variations of this method to cool any refrigerants need to cool natural gas to the required temperature usually using spiral wound heat exchangers.

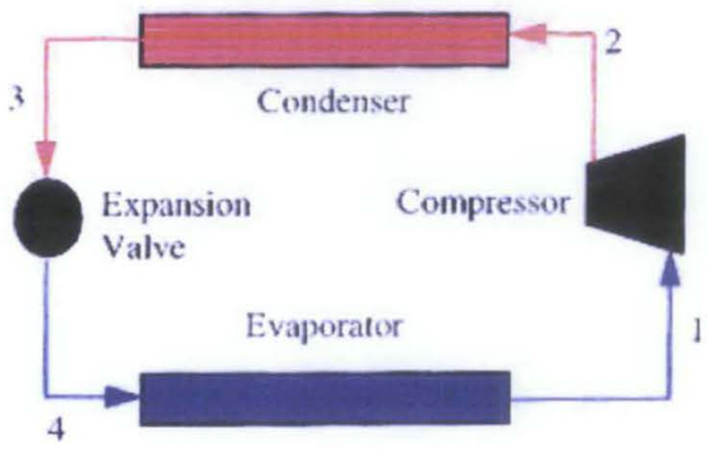


Figure 6: Refrigeration cycle [6]

2.3.2 *TQ Diagrams*

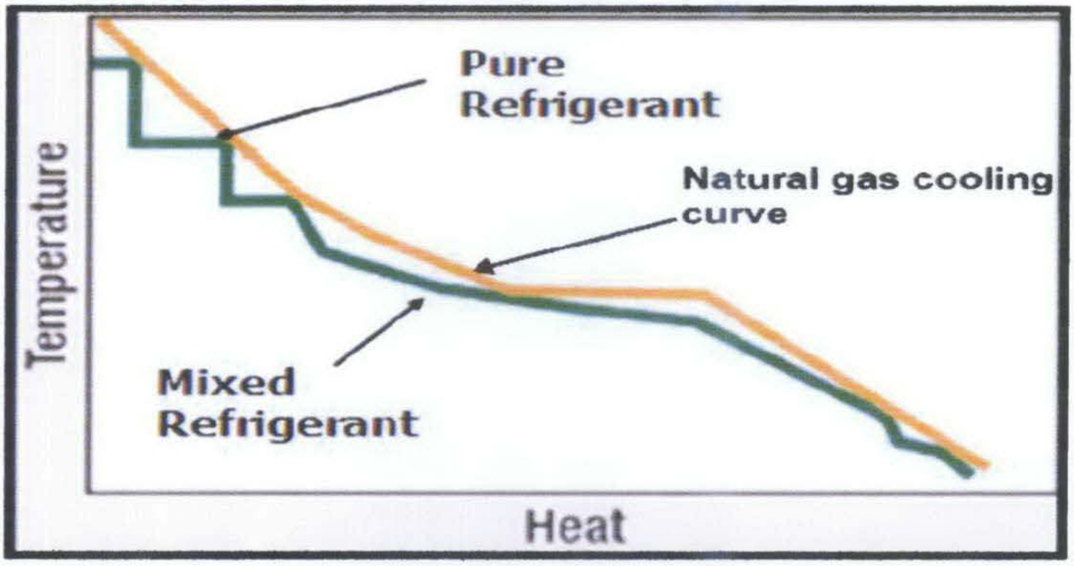


Figure 7: Natural gas refrigerant cooling curve [6]

Above is the example of what a typical temperature-heat diagram or cooling curve, for the cooling of natural gas using both pure and mixed refrigerants would look like. During cooling, it desires to have efficient process. One of the methods to determine the efficiency of a cycle is to review the cooling curve. The closer the line depicting the refrigerants is to the curve of the natural gas, the more efficient is the cycle. Increasing the efficiency of the process reduces the amount of the work done by the heat exchangers. The amount of work done by the heat exchangers is indicated by the spaces between the curves.

2.4 Liquefaction Technologies

As new LNG projects are developed, there are opportunities in the integration of newer technologies for larger baseload facilities and niche markets of modest throughput to the world of LNG. This results in renewed competition among process licensors, causing different stages of developing and marketing new or modified liquefaction process technology. Among the technologies available are:

- Classic cascade
- Optimised cascade
- Mixed fluid cascade

- Single-mixed refrigerant cycle (SMR)
- APCI (Air Products and Chemicals Incorporation) propane pre-cooled mixed refrigerant process (C3MR)
- Shell Double Mixed Refrigerant process (DMR)
- APCI AP-X

Although there are many technologies available, APCI propane precooled mixed refrigerant (C3MR) process still dominates in the world of LNG. However due to the inherent limitations of using a single component refrigerant in precooling in C3MR design, Shell had developed Double Mixed Refrigerant (DMR) process to add an addition degree of freedom. This allows the full utilisation of power in compressors, keeping then at their best efficiency points over a very wide range (up to 50°C) of ambient temperature variations and changes in feed gas composition. The process flow for both technologies is described below.

2.4.1 APCI propane pre-cooled mixed refrigerant process (C3MR)

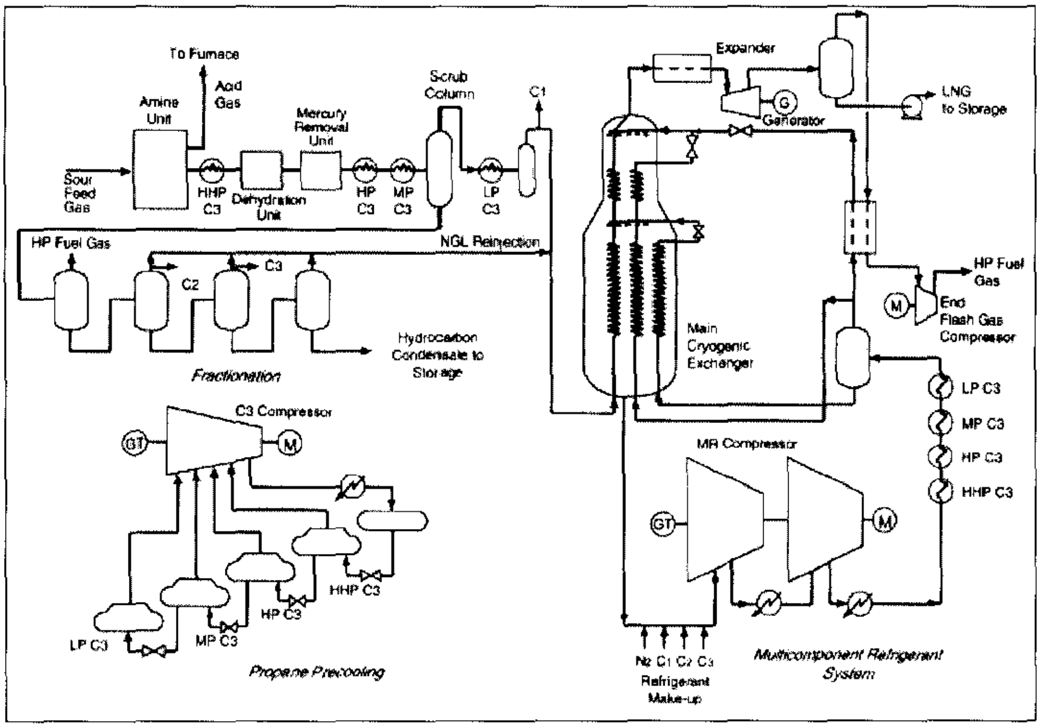


Figure 8: C3MR process flow [5]

There are two main refrigerant cycles in C3MR process as shown above. 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 C3 streams and compresses the vapour 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 vapour 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 vapour 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 lower bundles.

The overall vaporised 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.

2.4.2 Dual Mixed Refrigerant process (DMR)

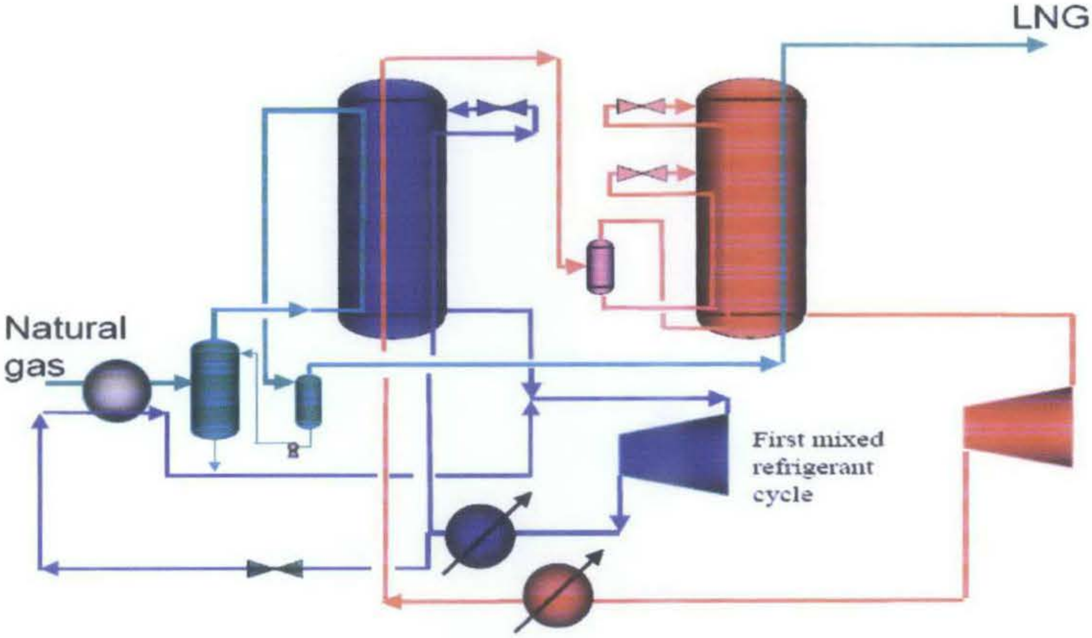


Figure 9: DMR process flow

DMR is very similar to the APCI liquefaction process. it is designed to overcome the inherent limitations of using a single component refrigerant in pre-cooling in the C3MR design; the additional degree of freedom resulting from the use of two mixed refrigerant cycles allows full utilisation of power in a design with two mechanically driven compressors. Furthermore it allows keeping the compressors at their best efficiency point a over a very wide range (up to 50°C) of ambient temperature variations and changes in feed gas composition. [4]

From Figure 9, the natural gas stream is cooled via two stages. The first stage cools natural gas to -50°C while the second column cools natural gas to LNG at -160°C. The composition of the pre-coolant cycle is 50/50 of ethane/propane on molar basis and the coolant composition of the cooling cycle is similar to the composition of APCI. [4]

In this process the heat exchanger tower is divided into two sections and this concept allows the choosing of load on each refrigeration cycle through controlling the two compressors work before each column. [4]

2.5 Process Enhancement Method

Several methods in terms of process and machinery integration advancements can be utilised to accommodate the ever changing requirement of the industry. This can assist in optimising the cost operation of the major equipment, expanding its production in a flexible, reliable and stable manner.

2.5.1 Expansion method

Generally expansion can be isenthalpic via throttling device such as Joule Thomson valve or isentropic which occurs on a work producing expansion turbine.

Joule Thomson effect describes the change in temperature of a thermally insulated non-ideal gas when it is suddenly expanded through a small dole or a porous material (from a high pressure to a low pressure). The ratio of $\Delta T/\Delta P$ is known as the Joule Thomson coefficient. Expansion turbine (turboexpander) is a centrifugal or axial flow turbine expands high pressure gas to produce work, usually used to drive compressor.

Natural gas to be liquefied is first compressed and liquefied at a pressure above at which it is stored or used. This pressurised liquid (usually at boiling point) then has its pressure released through at expansion orifice or valve (commonly known as Joule Thomson nozzle) to atmospheric pressure for storage in a cryostat. In such process, a portion of depressurised liquid vaporises. This vapour is disengaged from fluid and recycled or otherwise disposed of. The residual liquid constitutes the storable liquid.

However, if expansion is achieved isentropically, it will not only increase the amount of liquefied gas, but also obtain useful work or power from the reduction of pressure of liquefied natural gas produced. However the presence of comparatively dense liquid along with light vapour makes turbine expansion of saturated pressurised liquefied gas difficult and at low efficiency. Expansion in piston type expansion engine is difficult because of the effect that liquid has on pistons, seals and valves. Hence it is customary to flash gas directly to atmospheric pressure through Joule-Thompson nozzle, losing the energy which could be conserved.

But if the pressurised gas is flashed to an intermediate pressure, the flash gas produced by initial expansion can be separate from remaining liquid and expanded isentropically. Among the advantages are:

- Increase amount of liquefied gas that is retained when the pressure of the liquefied gas is reduced from process pressure to terminal pressure (A pressure drop across a unit when the maximum allowable pressure drop is reached)for storage
- Obtain useful work from reduction of pressure of liquefied natural gas produced from liquefaction process to storage or terminal pressure
- Provide the flashing of pressurised liquefied natural gas to an intermediate pressure stage, then isentropically expanding the vapour (resulting from flashing) to terminal pressure stage to provide useful work and additional liquefied end product
- Provide flashing of pressurised gas to a point above its terminal pressure, disengaging vapour from flashing and expanding it isentropically in a turboexpander to its terminal pressure
- Decrease amount of liquefied gas that vaporises when pressure is reduced to terminal pressure for storage by expanding pressurised liquefied gas isenthalpically to an immediate pressure, disengaging the flash vapour, there from isentropically expanding the intermediately disengaged vapour to terminal pressure and isenthalpically expanding the intermediately disengaged liquid to terminal pressure
- Isenthalpically expand pressurised liquid in intermediate stages and after each isenthalpic expansion, isentropically expand resulting vapour to the next subsequent stage

2.5.2 *Hydrocarbon extraction*

The extraction of heavier hydrocarbon can help in achieving the lower heating value (HHV) and Wobbe Index required by different countries. There are two types of extraction schemes, extraction before liquefaction unit and extraction integrated with liquefaction unit where there are many process configurations for each. [8]

1. NGL/LPG extraction before liquefaction unit

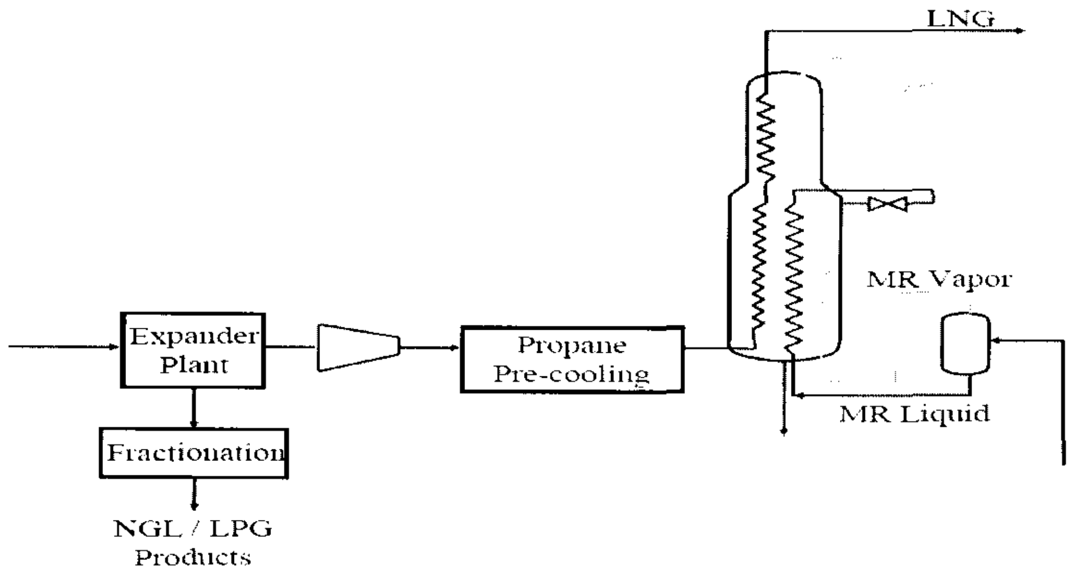


Figure 10: NGL/LPG extraction before liquefaction unit

The upstream expander section separates out heavier hydrocarbons which are sent to fractionation unit for further processing into LPG products as shown in Figure 10. Recompression of natural gas after expander plant and before the liquefaction unit can lead to lower specific power and higher overall efficiency. Besides, it also results in higher LPG extraction and high ethane extraction. However, it requires more equipment.

2. NGL/LPG extraction integrated with liquefaction unit

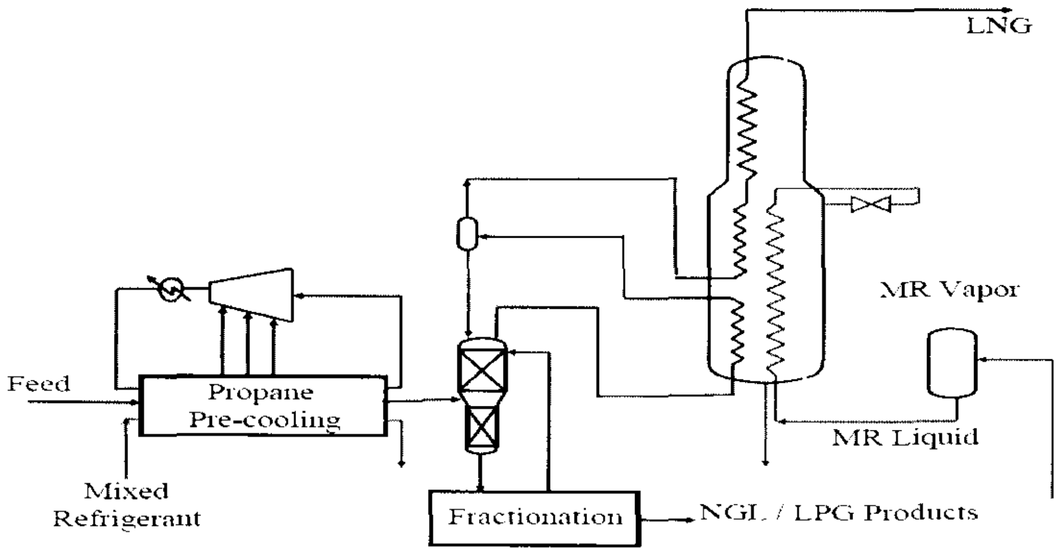


Figure 11: NGL/LPG extraction integrated with liquefaction unit

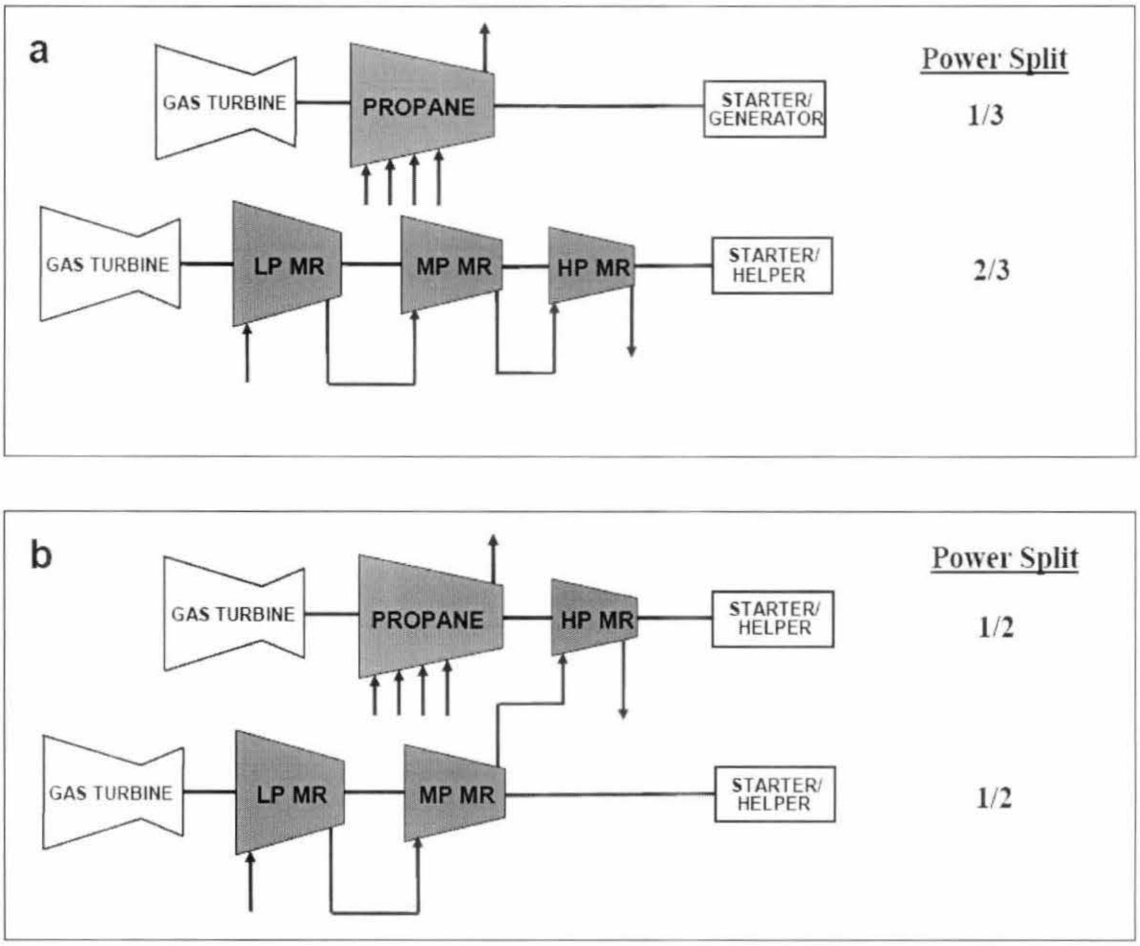
Figure above shows a scrub column is used for LPG extraction and the scrub condenser duty is provided by condensing against mixed refrigerant in the warm bundle of the main cryogenic heat exchanger. By using mixed refrigerant, lower scrub column overhead temperature can be achieved, i.e. -70°C (-35°C for propane). This allows LNG to be produced with lower HHV and Wobbe Index. However, the pressure in scrub column must remain below critical pressure to allow adequate liquid/vapour separation. This extraction method is simpler with less equipment required, but results in less LPG extraction and ethane extraction of less than 20%.

2.5.3 Compressor

About 40% of total operating cost of LNG plant is due to consumption of energy in refrigeration section which mainly involves compressor. Hence, it plays a crucial part in the optimisation of the overall plant process. Below describes some of the process and machinery integration advancements that can minimise the total power cost of refrigerant compressors. [8]

Split MR Technology

Split MR Technology uses a portion of mixed refrigerant compression requirement driven by the same driver as used for propane compression. The power balance becomes evenly split and this allows for full utilisation of gas turbine power and increases train capacity for the same number of drivers and compressors.

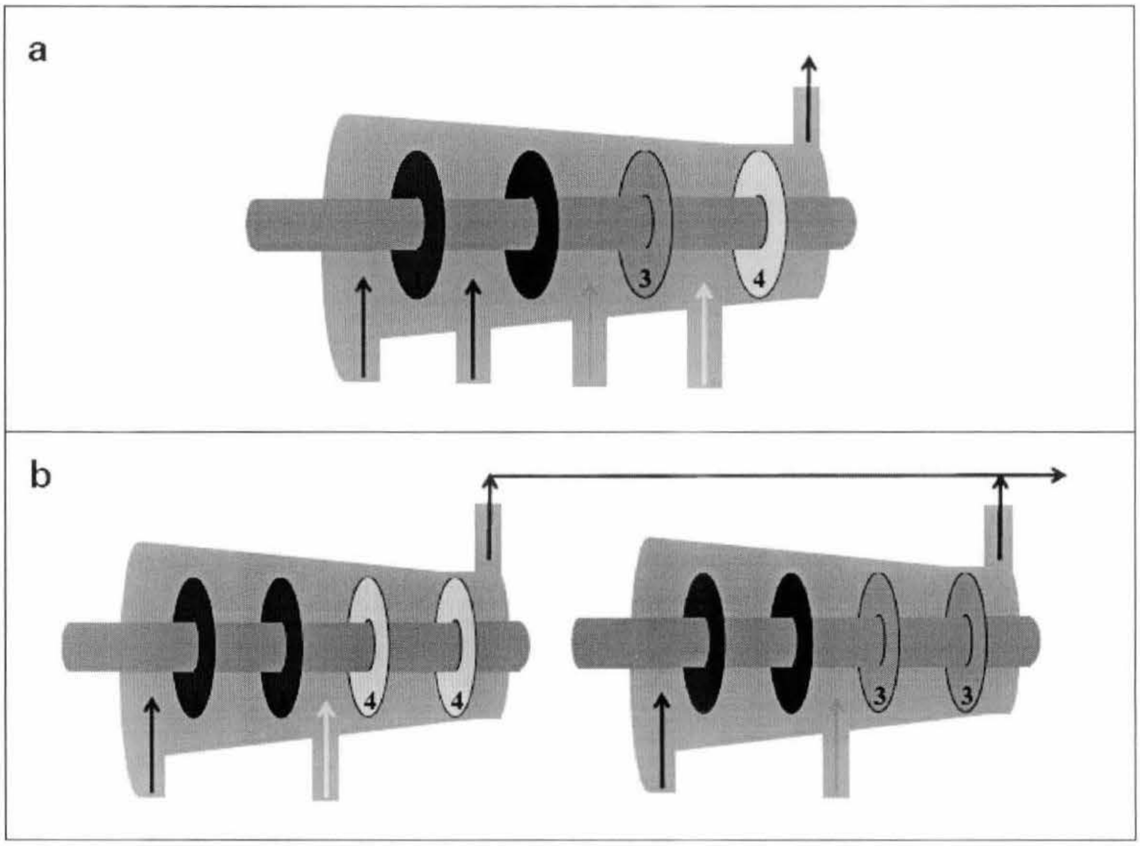


(a) Separate Driver Configuration for Propane and MR Compressors

(b) Split MR Machinery Configuration

Propane Casing Arrangement

A 1,4-2,3 split propane compressor casing in a series arrangement can be used. Stage 1 and 4 are in first casing while stages 2 and 3 in second casing. The inlet pressures to four stages may be different than the single casing compressor design and are adjusted to maximise efficiency. The discharges from third and fourth stages are at the same pressure since they are connected to common condenser. Each stage would usually have multiple impellers. This series arrangement minimises the complexity of suction piping and avoids the potential for imbalances in compressor duties that can occur in parallel compression.



(a) Single Casing Propane Compressor

(b) 1,4-2,3 Split Casing Propane Compressor

Compressor Drivers

Options available include steam turbines, gas turbines (e.g. Frame 5,6,7, 8 and aero-deravatives) and electric motors and its availability is shown in the table below. Most C3MR projects use Frame 7 gas turbines with an ISO power of approximately 86 MW at 3600 rpm. The power and efficiency significantly improved as Frame number increases.

Table 2: Typical component availabilities

Equipment	Average reliability	Scheduled maintenance (avg. hrs/annum)
Centrifugal compressors	**0.998**	**50**
Electric motors	**0.997**	**25**
Gas turbines	**0.994**	**270**
Steam turbines	**0.994**	**45**

Aeroderivative gas turbines have recently been proposed for baseload LNG services. The usage of aeroderivative gas turbine as refrigerant compressors drivers offers several advantages. These include:

- Much higher efficiency with its advantages of reduced fuel consumptions and reduced greenhouse emissions
- Ability to rapidly swap engines and modules thus improving maintenance flexibility
- Excellent starting torque capacity (allows large trains to start up under settle out pressure conditions)
- The engine is essentially zero timed after six years. Maintenance can also be done "on condition", allowing additional flexibility
- Easy installation due to low engine weight

Electric motors are used intensively as starter and helper motors for gas turbines. Electric motors are very efficient, but efficiency of the power source must be taken into consideration when determining overall efficiency. They have higher availability due to less frequent maintenance and shorter outage requirements.

2.5.4 Operating parameters

The control of plurality of parameters include pressure, temperature, flowrate, composition and liquid level at specific location in the process at its desired set point enable the plant to operate in a desired production rate with the highest possible efficiency.

CHAPTER 3
METHODOLOGY

3.1 Research Methodology

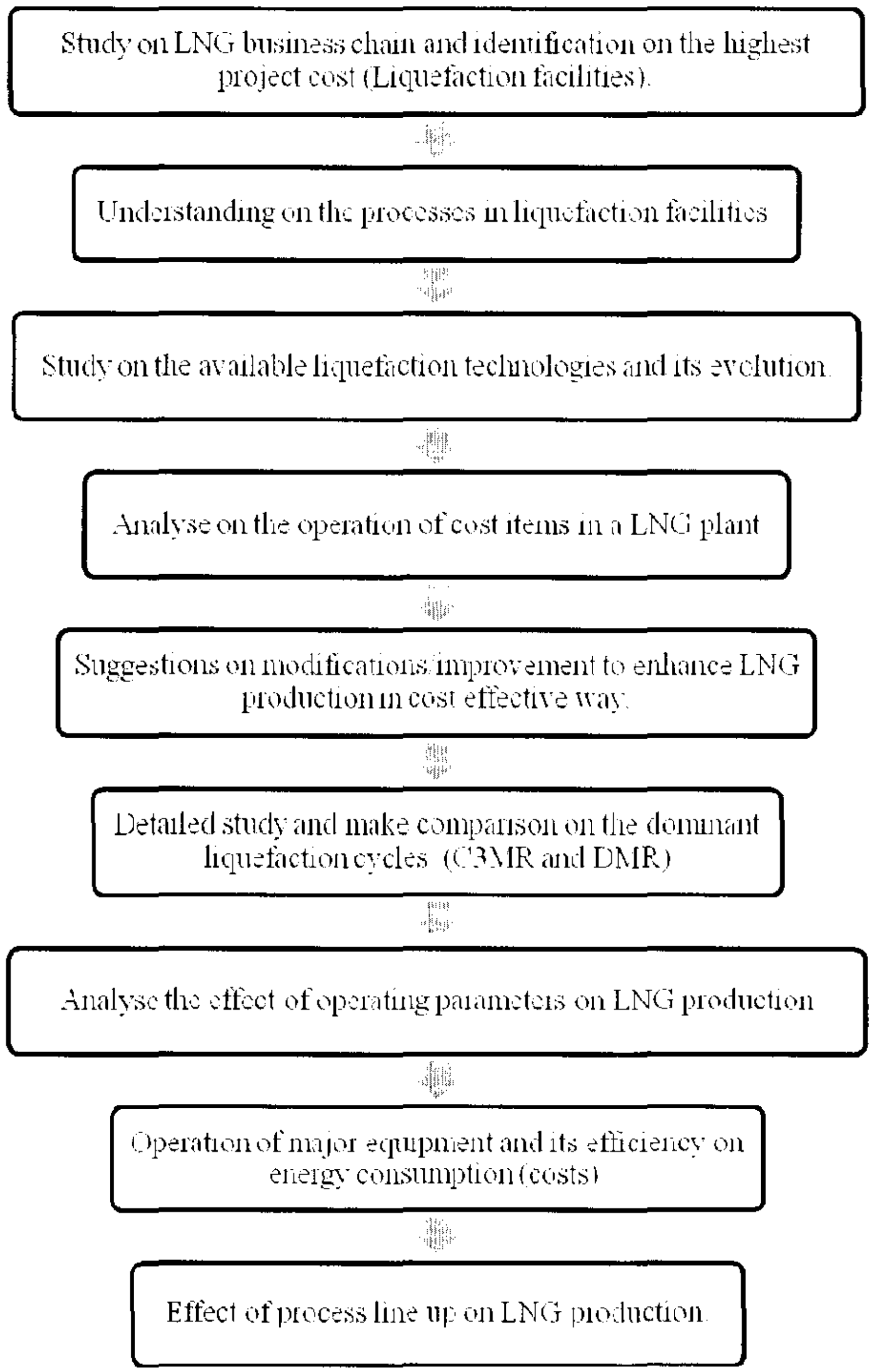


Figure 12: Research methodology

This project depends mostly on research work to obtain information, especially through the internet regarding the LNG industry. The collections of technical details on the development of LNG in regards to its technologies and operating conditions over the years is essential in making comparison and develop innovative methods to increase the capacity of LNG production in a cost effective manner.

3.2 Project activities

Based on the analysis and studies done, the crucial part of this project lies in modelling the process through computer aided simulation to provide a systematic approach in presenting the innovative methods identified. The simulation will be done using HYSYS software. The draft of plan to start simulation can be seen in the figure below.

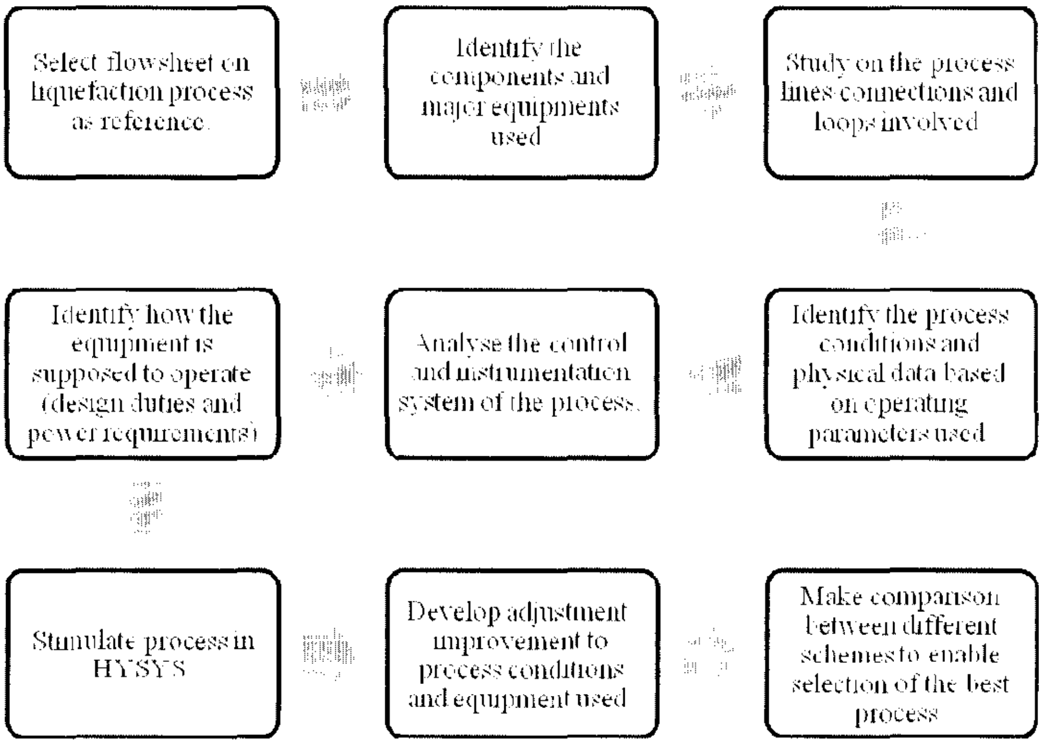


Figure 13: Draft of plan on getting the objectives using HYSYS

Table 3: Information needed for HYSYS

PROCESS VARIABLES	PHYSICAL PROPERTIES	EQUIPMENT DETAILS	PROCESS FLOW
•Stream number •Temperature •Pressure •Composition and state •Total mass flow rate (kg/h) •Total molar flow rate (kmol/h) •Individual components flow rates (kg/h) •Volumetric flow rates (m3/h)	•Molecular weight •Density •Viscosity •Thermodynamic data (heat capacity, boiling point, etc)	•Column (type, operating pressure and temperature, design parameters) •Vessels (diameter, operating conditions) •Heat transfer equipmemt (heat duty, operating temperatures) •Pumps (operating capacity, differential head) •Compressors (operating capacity) •Mixers (type, capacity)	•Control systems •Instrumentation •Process lines •Valves •Recycle streams •Utility lines •Disposal lines (drains, vent)

3.3 Gantt Chart

No.	Detail / Week	1	2	3	4	5	6	7		8	9	10	11	12	13	14	15	16	17	18	19	20
1	Project work commences								MID SEMESTER BREAK								STUDY WEEK	EXAMINATION WEEK				
	- Flowsheet confirmation																					
	- HYSYS simulation																					
2	Submission of Progress Report 1					25th Feb.																
3	Project work continues																					
	- Continuation on HYSYS																					
	- Improvement/ Modifications																					
4	Poster Exhibition/ Pre-EDX/ Progress Reporting												14th, 15th Apr.									
5	Submission of Progress Report 2 (Draft of Final Report)												16th Apr.									
6	EDX (selected)																					
7	Submission of Final Report (CD and Softbound)															7th May						
8	Final oral presentation																					
9	Submission of hardbound copies and CD																					18th June

Figure 14: Project Gantt Chart

CHAPTER 4
RESULTS AND DISCUSSION

The APCI propane pre-cooled mixed refrigerant (C3MR) is the most widely used base-load LNG process which involves two refrigeration cycles, namely propane refrigeration and mixed refrigerant. Thus, it is selected as basis of simulation in HYSYS. For the physical properties of the design flow, the case will be based on average gas and average ambient temperature condition.

4.1 Process Flow

The purposes of liquefaction process include the removal of heavy hydrocarbons from treated natural gas feed and liquefy the remaining natural gas. The C3MR covers three main systems which are natural gas circuit, propane refrigerant and mixed refrigerant system as seen in Figure 15. Among the major equipment utilised in this area are:

- Heat exchangers
- Compressors
- Main cryogenic heat exchanger (MCHE)
- Gas turbines
- Generator (electricity)
- Fin fans

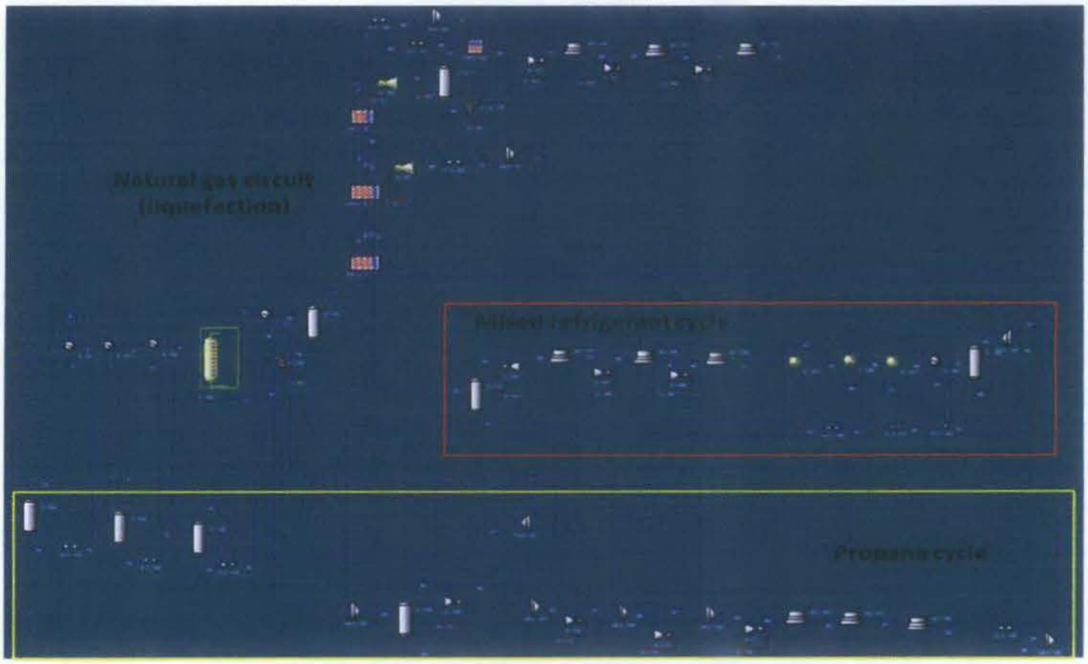


Figure 15: Process flow from HYSYS simulation

4.1.1 Natural Gas Circuit

Following the removal of acid gases, dehydration, and mercury removal, the treated natural gas is precooled against High High Pressure (HHP), High Pressure (HP) and Medium Pressure (MP) propane before being fed into a column. The column is set to remove heavy hydrocarbons (C5+) that could freeze out in the cold part of the Main Cryogenic Heat Exchanger and to recover LPG components (propane and butane) as bottom product of the column which ensure LNG production is on spec at rundown. The overhead gas will generate reflux to the column through partial condensation against Low Pressure (LP) propane to a temperature around -30°C. The scrubbed gas will then be fed into MCHE for further cooling and liquefaction.

In MCHE, natural gas is liquefied and subcooled to a temperature of minus 160°C using mixed refrigerant. Liquefied Natural Gas (LNG) is expanded to a lower pressure using an expander before being flashed to a vessel. LNG will be pumped to storage. The end flash gas is compressed in three stage compressor to raise the vapour pressure to HP fuel gas pressure.

4.1.2 Propane Refrigerant System

Propane is used to precool natural gas and Mixed Refrigerant (MR). Four pressure stages of propane vapour from heat exchangers are routed to vessels before being compressed in a four stage compressor. The desuperheating, condensing and subcooling of propane refrigerant is achieved through fin fans.

4.1.3 Mixed Refrigerant System

Mixed refrigerant consists of a mixture of nitrogen, methane, ethane and propane. It is used for heat extraction of natural gas in liquefaction. Mixed refrigerant exiting MCHE will be compressed and intercooled in stages using fin fans before being sent back to MCHE, forming a close loop mixed refrigerant cooling cycle.

The further cooling and partial condensation is achieved through the boiling of propane refrigerant in heat exchangers at four different pressure levels. It is then separated into light mixed refrigerant (LMR) vapour and heavy mixed refrigerant (HMR) liquid, routed separately to MCHE to be cooled.

4.2 Power consumption in liquefaction unit

Table 4: Description of equipments and desired power

EQUIPMENT	DESCRIPTION	DESIRED POWER (MW)
Compressor	Propane	60.8
	Mixed refrigerant	76.9
	Liquefaction (end flash)	7.073
Gas turbine	Propane compressor	87.7
	Mixed refrigerant compressor	87.7
Generator (electricity)	Propane compressor	10
	Mixed refrigerant	10
	LNG expander	1
	Heavy mixed refrigerant expander	0.8
	End flash compressor motor	7.8
Heat exchanger	NG/LP propane condenser	10.993
	MR/LP propane vaporiser	25.165
	NG/MP propane vaporiser	7.371
	MR/MP propane vaporiser	29.105
	NG/HP propane vaporiser	7.731
	MR/HP propane vaporiser	32.451
	NG/HHP propane cooler	11.508
	MR/HHP propane vaporiser	22.521
	Light MR/endflash	2.276
Main Cryogenic Heat Exchanger (MCHE)		85.679

Fin fan	Propane desuperheater	23.892
	Propane condenser	162.998
	Propane subcooler	26.118
	LP MR compressor aftercooler	45.152
	HP MR compressor intercooler	23.187
	HP MR compressor aftercooler	22.18
	Endflash compressor 1st stage intercooler	1.739
	Endflash compressor 2nd stage intercooler	2.018
	Endflash compressor aftercooler	1.969

LNG plants are energy intensive where a world scale LNG plant usually consumed about 5.5-6 kWh energy per kmol of LNG produced. About 40% of the total operating cost for a base-load LNG is due to the consumption of energy in the refrigerant section.

The data from Table 4 is obtained from a Liquefied Natural Gas Plant in Malaysia on the power consumption of the main equipments for liquefaction process. The plant uses some of the largest compressors in the world in the refrigeration system, driven by frame-type turbines using natural gas as fuel or electric motors.

Table 5 below summarises the total power used by each of the major equipment used in the plant's liquefaction process.

Table 5: Summary of equipments' total power

EQUIPMENT	TOTAL POWER (MW)
Compressor	144.773
Gas turbine	175.4
Generator (electricity)	29.6
Heat exchanger	149.121
MCHE	85.679
Fin fan	309.253

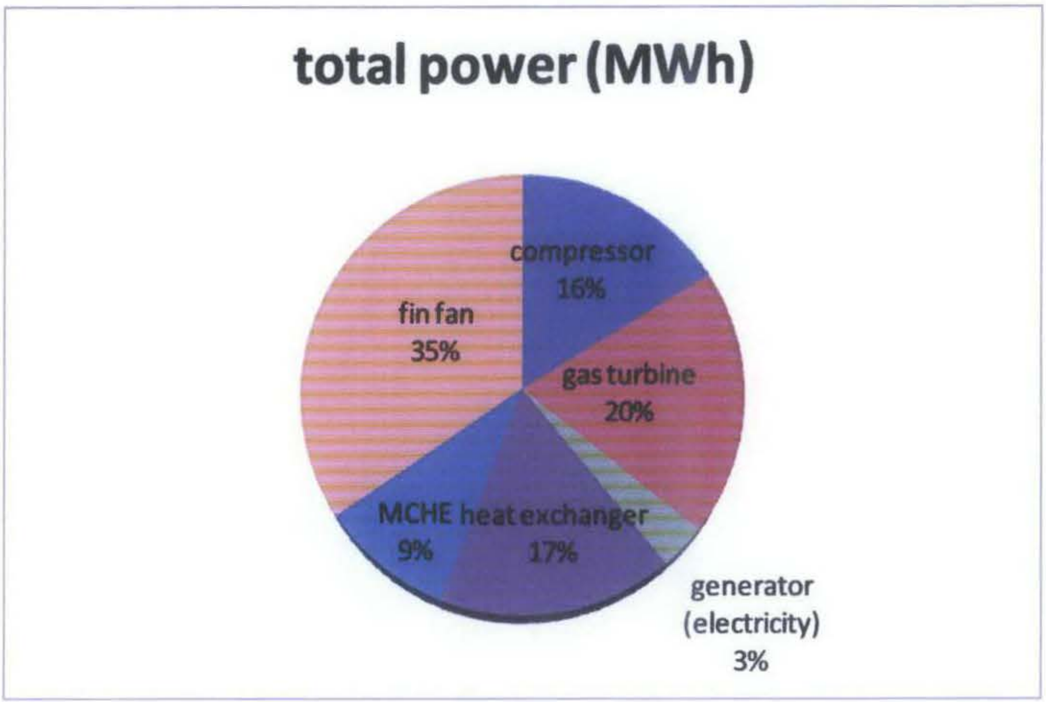


Figure 16: Comparison of power usage in different equipments

Based on the information obtained, we can conclude from figure above that the power used in fin fan for the cooling of refrigerant is the highest compared to the other equipments. From Figure 17, we can see that the compressor used in the mixed refrigerant system accounts for the highest power consumption if compare to propane refrigerant system and liquefaction process.

4.2.1 Comparison of power consumption in compressor

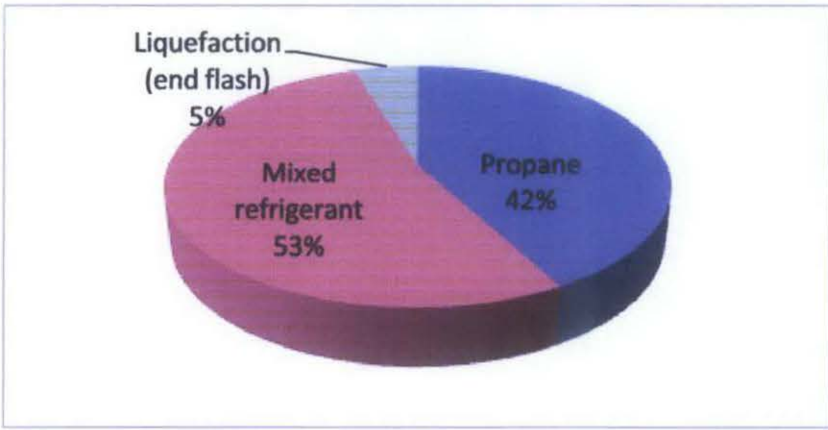


Figure 17: Power consumption of different compressor in liquefaction process

The operating characteristics of compressor depends on the parameters below in determining its preformance and efficiency which will also indirectly lead to the power and operating cost required to run a compressor. The graph below indicates how a compressor will perform based on the parameters.

- Molecular weight
- Temperature and pressure
- Compressibility factor
- Specific heat ratio (k)
- Speed

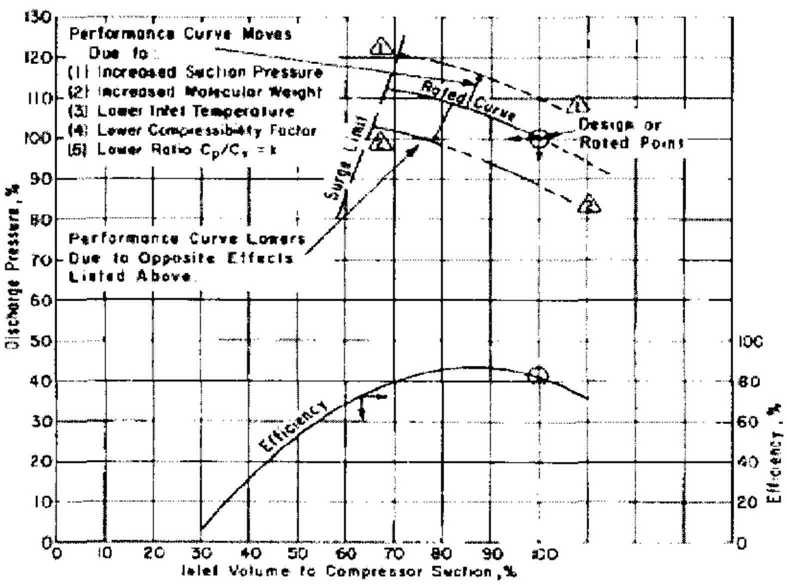


Figure 18: Compressor performance curve

Stonewall represents the maximum stable compressor flow point and minimum head point under stable compressor operation. Surge line is opposite of stonewall which it is the minimum stable compressor flow point and also the highest head point. Nevertheless, lower molecular weight component will lead to a lower discharge pressure as shown in the graph below. This is due to less power needed to compress the component.

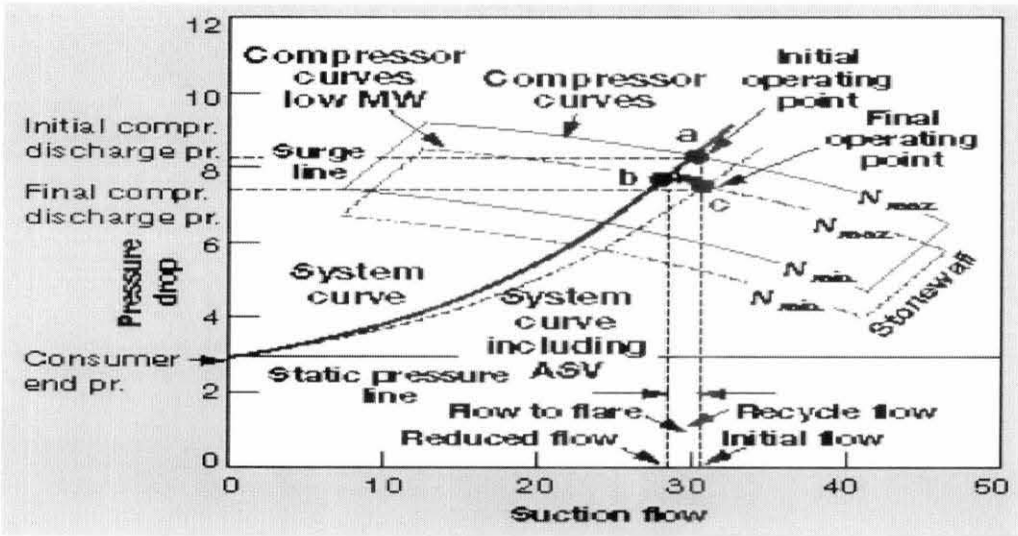


Figure 19: Molecular weight on compressor's operating curve

The scope of study is narrowed down to the compressor used propane refrigeration cycle as propane plays a crucial part in cooling both natural gas and mixed refrigerant using four different stages of pressure, which are LP (210kPa), MP (299kPa), HP (480kPa) and HHP (2073kPa) in liquefaction process. The parameter that is being manipulated is the composition of propane refrigerant as shown in Table 6 to obtain the inlet and outlet temperature at different stages of pressure at compressor and the power used in compressor. The value obtained in power is used to calculate the operating cost of the compressor.

Table 6: Different propane refrigerant compositions used in simulation

composition	1	2	3	4	5	6
C1	0.02	0.015	0.01	0.005	0.0025	0
C2	0.03	0.025	0.02	0.01	0.0075	0
C3	0.85	0.88	0.91	0.95	0.98	1
n-C4	0.05	0.04	0.03	0.0175	0.005	0
i-C4	0.05	0.04	0.03	0.0175	0.005	0
Total	1	1	1	1	1	1

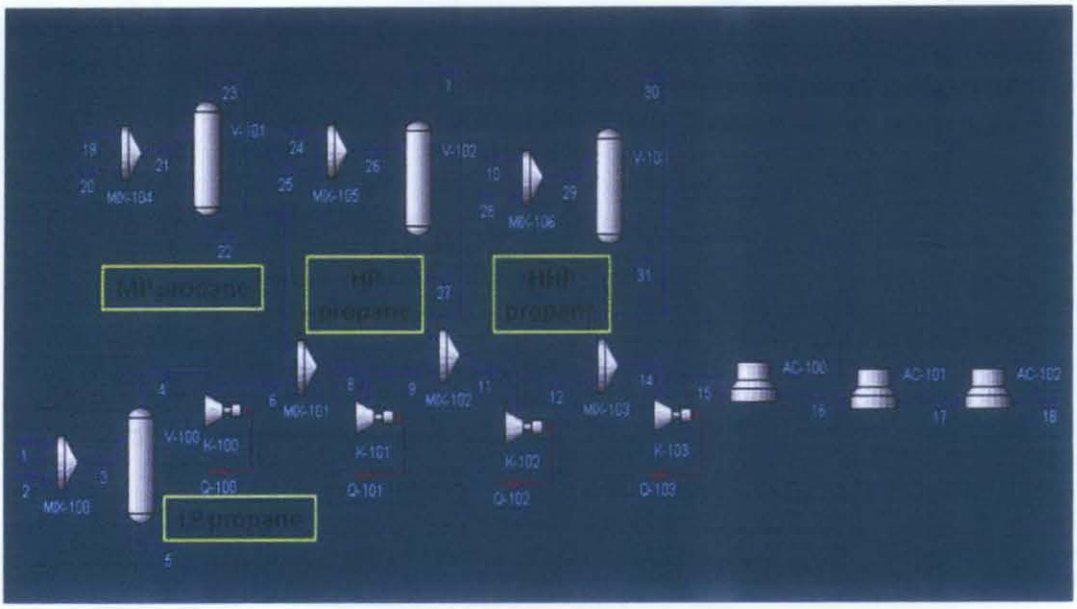


Figure 20: Propane refrigerant system from HYSYS simulation

Below are the data obtained from HYSYS simulation at propane refrigerant system.

- *Inlet temperature (°C) of compressor at different stages and propane refrigerant composition*

Table 7: Inlet temperature of compressor

composition	1	2	3	4	5	6
LP	-37.09	-37.09	-37.09	-37.09	-37.09	-37.09
MP	-17.45	-17.51	-17.55	-17.49	-17.36	-15.29
HP	-3.426	-3.516	-3.582	-3.543	-3.452	0.2818
HHP	14.94	14.67	14.33	13.77	14.04	15.94

- *Outlet temperature (°C) of compressor at different stages and propane refrigerant composition*

Table 8: Outlet temperature of compressor

composition	1	2	3	4	5	6
LP	-14.57	-14.73	-14.88	-15.07	-15.21	-15.29
MP	-1.397	-1.564	-1.707	-1.774	-1.746	0.2818
HP	18.65	18.42	18.22	18.11	18.07	21.75
HHP	88.66	88.1	87.5	86.66	86.78	88.47

- *Change in inlet and outlet temperature (°C) of compressor at different stages and propane refrigerant composition*

Table 9: Change in inlet and outlet temperature of compressor

composition	1	2	3	4	5	6
LP	22.52	22.36	22.21	22.02	21.88	21.8
MP	16.053	15.946	15.843	15.716	15.614	15.5718
HP	22.076	21.936	21.802	21.653	21.522	21.4682
HHP	73.72	73.43	73.17	72.89	72.74	72.53

- *Change in power requirement of compressor (MWh) with the change in propane refrigerant composition*

Table 10: Power requirement of compressor

composition	1	2	3	4	5	6
LP	1.015	1.007	0.9918	0.8638	1.624	1.979
MP	1.405	1.38	1.338	1.108	1.951	1.458
HP	3.5	3.414	3.27	2.612	4.449	2.048
HHP	20.08	20.04	19.91	17.31	26.09	12.11

- *cost of HP fuel gas = RM25.6862/MWh*

Table 11: Cost of HP fuel gas usage in compressor

composition	1	2	3	4	5	6
LP	26.07149	25.866	25.47557	22.18774	41.71439	50.83299
MP	36.08911	35.44696	34.36814	28.46031	50.11378	37.45048
HP	89.9017	87.69269	83.99387	67.09235	114.2779	52.60534
HHP	515.7789	514.7514	511.4122	444.6281	670.153	311.0599

- *cost of compression for propane refrigerant use in compressor*

Table 12: Cost of compression for propane refrigerant at different stages

stage	cost (RM/MWh)
LP	28.22
MP	23.1
HP	17.57
HHP	12.01

Table 13: Cost of compression for propane refrigerant use in compressor

composition	1	2	3	4	5	6
LP	28.6433	28.41754	27.9886	24.37644	45.82928	55.84738
MP	32.4555	31.878	30.9078	25.5948	45.0681	33.6798
HP	61.495	59.98398	57.4539	45.89284	78.16893	35.98336
HHP	241.1608	240.6804	239.1191	207.8931	313.3409	145.4411

Note: The reference to the cost values are obtained from the marginal cost of utilities of Malaysia market.

4.3 Results and Discussion

Based on the data obtained from the simulation in HYSYS, the results had been represented in the graphs shown below. In order to obtain the best choice of composition, comparison had been done.

4.3.1 Inlet and outlet temperature of compressor

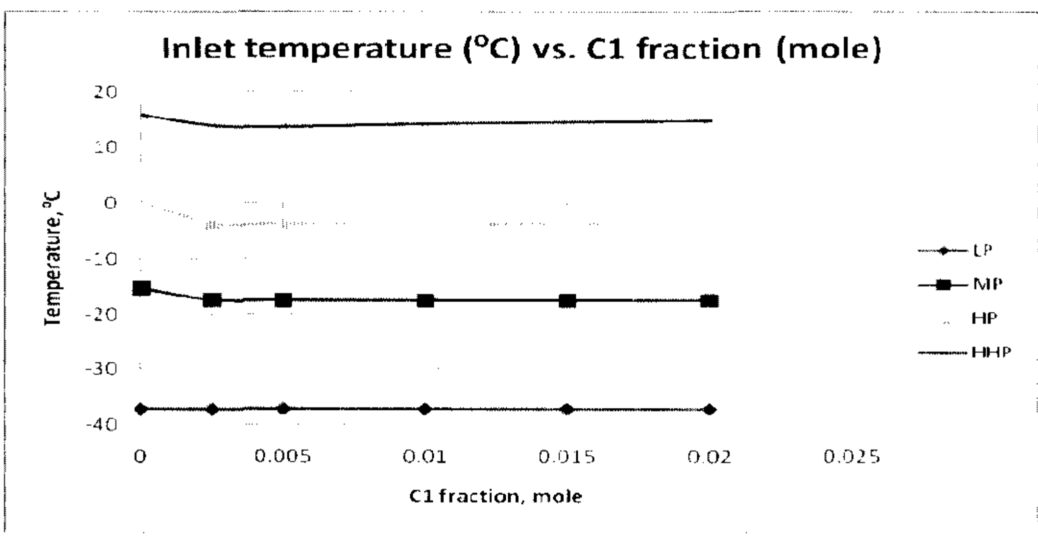


Figure 21: Graph of inlet temperature vs. C1 fraction

In Figure 21, the inlet temperature of compressor at LP pressure for methane (C1) fraction remains constant at -37.09°C, while for the other three stages of pressure; MP, HP and HHP, it shows the trend of slight increment starting at 0.0025 moles. Nevertheless, the higher the pressure, the higher the temperature achieved. However, the inlet temperature is the highest when C1 fraction is equal to 0 moles. This is the composition for pure propane where it is infeasible to achieve in real plant situation. The range of C1 composition is set to be from 0 to 0.02 moles.

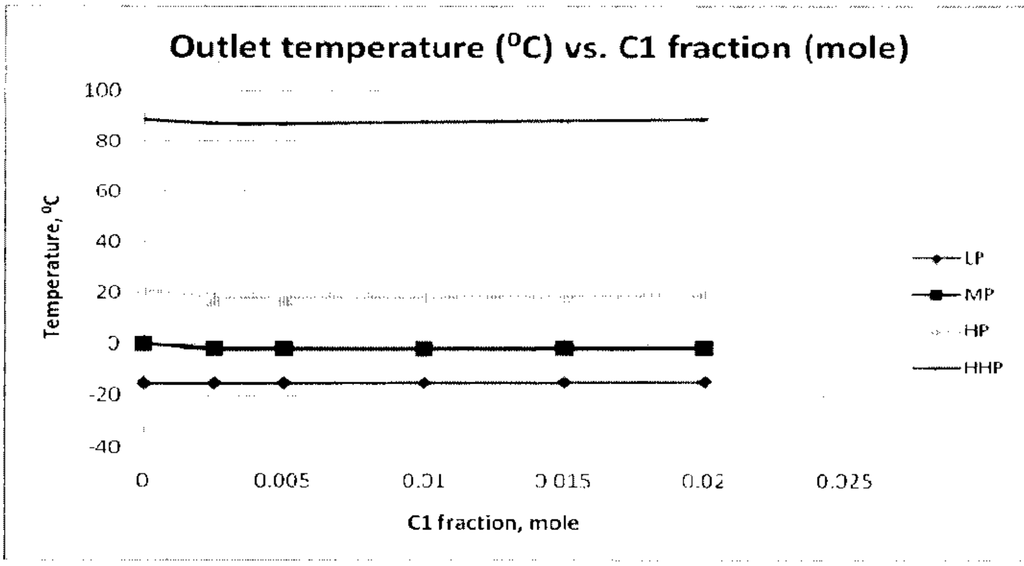


Figure 22: Graph of outlet temperature vs. C1 fraction

As for the outlet temperature of compressor for C1 fraction, the trend remains quite constant throughout the four stages of pressure. Nevertheless, the outlet temperature is the lowest when C1 fraction is equal to 0.0025 moles, while the highest at C1 equal to 0 moles (pure propane). The outlet temperature also projected the same pattern of pressure level, where the pressure increases simultaneously with the temperature as can be seen in Figure 22.

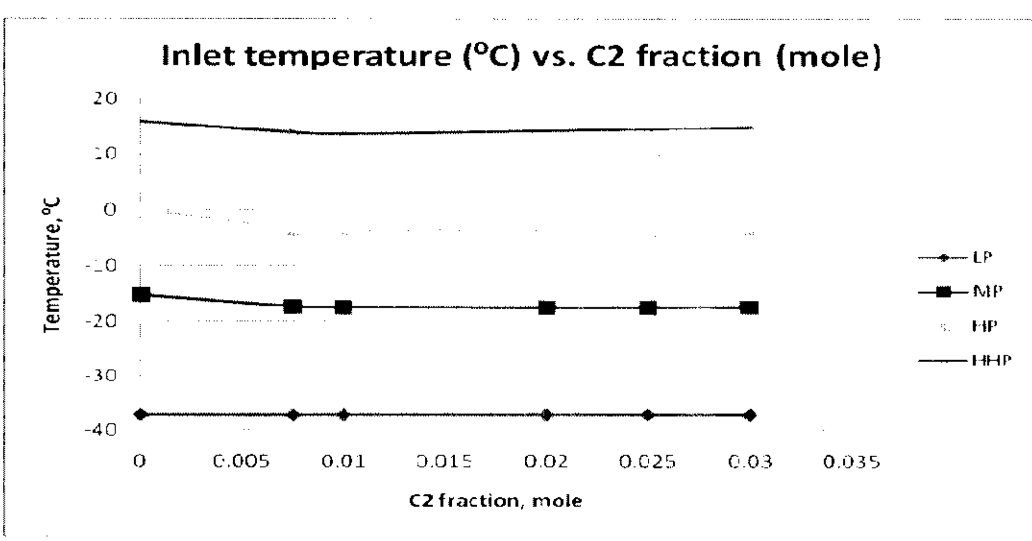


Figure 23: Graph of inlet temperature vs. C2 fraction

The graph from Figure 23 for inlet temperature for ethane (C2) fraction projected roughly the same trend as the graph of inlet temperature vs. C1 fraction. The composition of C2 fraction is in the range of 0 to 0.03 moles. The increment starts when C2 fraction is equal to 0.0075 moles. As the fraction of C2 increases, the inlet temperature also increases.

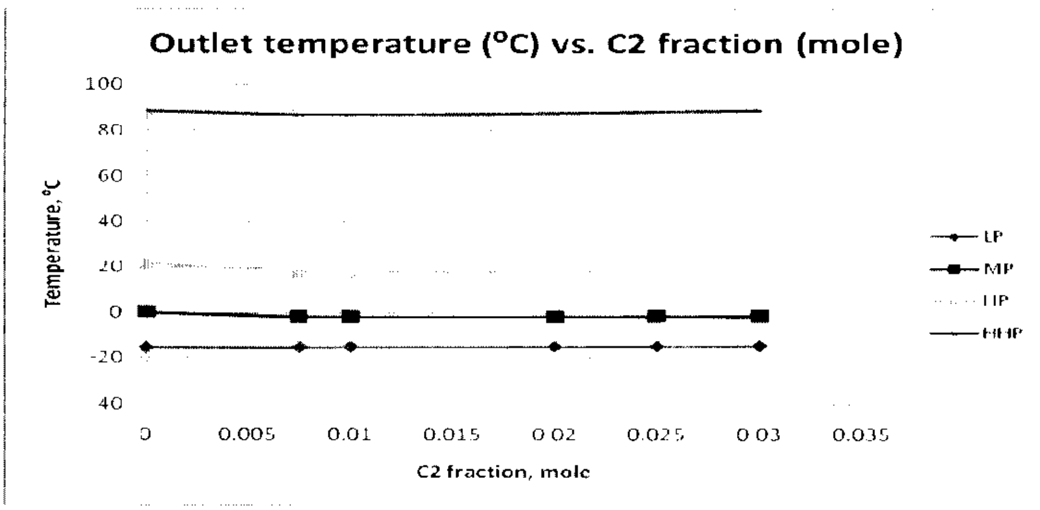


Figure 24: Graph of outlet pressure vs. C2 fraction

The outlet temperature for C2 fraction at all four stages also remains quite consistent throughout the graph in Figure 24. However, the lowest temperature achieved is when C2 fraction is equal to 0.0075 moles.

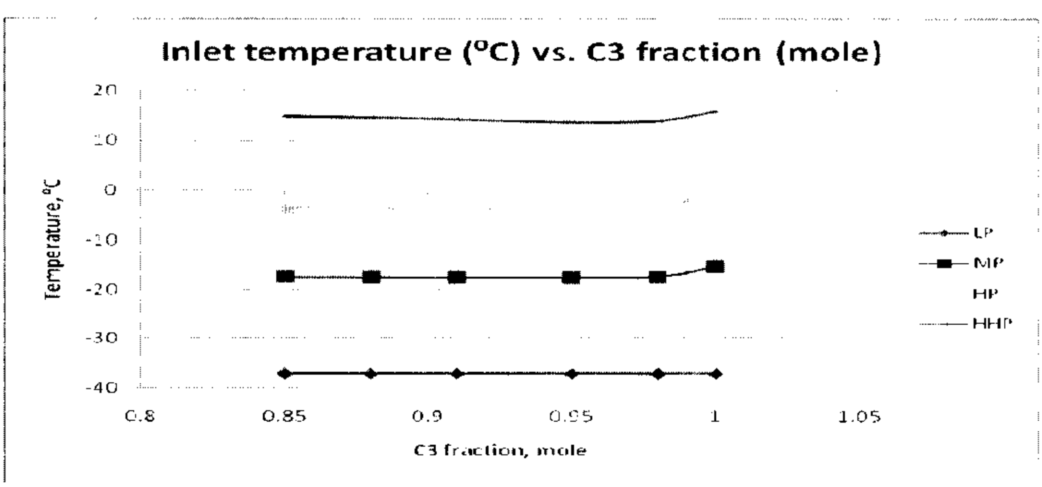


Figure 25: Graph of inlet temperature vs. C3 fraction

The composition of propane (C3) has been manipulated between the range of 85 to 100%, with the highest range compared to the other components. From the graph above, we can see that the inlet temperature slightly decreases as C3 fraction increases from 0.85 to 0.98 moles. However, there is a sudden increase when the fraction is 1, where this is the pure propane composition. Nevertheless, this does not apply for LP pressure as the temperature remains unchanged throughout the increase in C3 fraction.

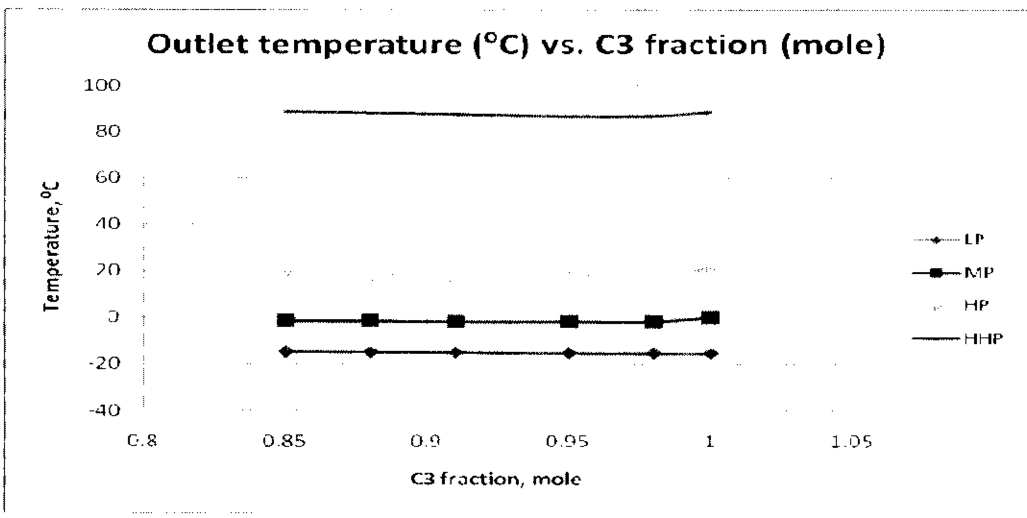


Figure 26: Graph of outlet temperature vs. C3 fraction

As for the outlet temperature for C3 fraction, the trend of graph above remains quite constant throughout the increase in C3 fraction. Only a slight increase can be seen when C3 fraction is equal to 1 (pure propane).

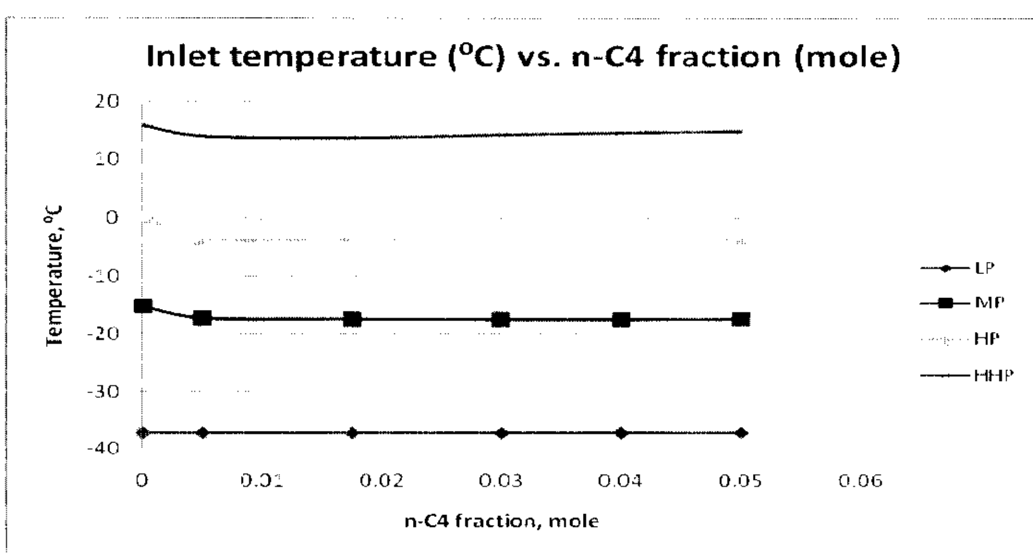


Figure 27: Graph of inlet temperature vs. n-C4 fraction

The trend shown in the graph of Figure 27 is almost the same as shown in the inlet temperature of C1 and C2 fraction. The inlet temperature starts to increase when the composition of n-butane (n-C4) is 0.005 moles and this also does not apply for LP pressure. The range of n-C4 covers from 0 to 0.05.

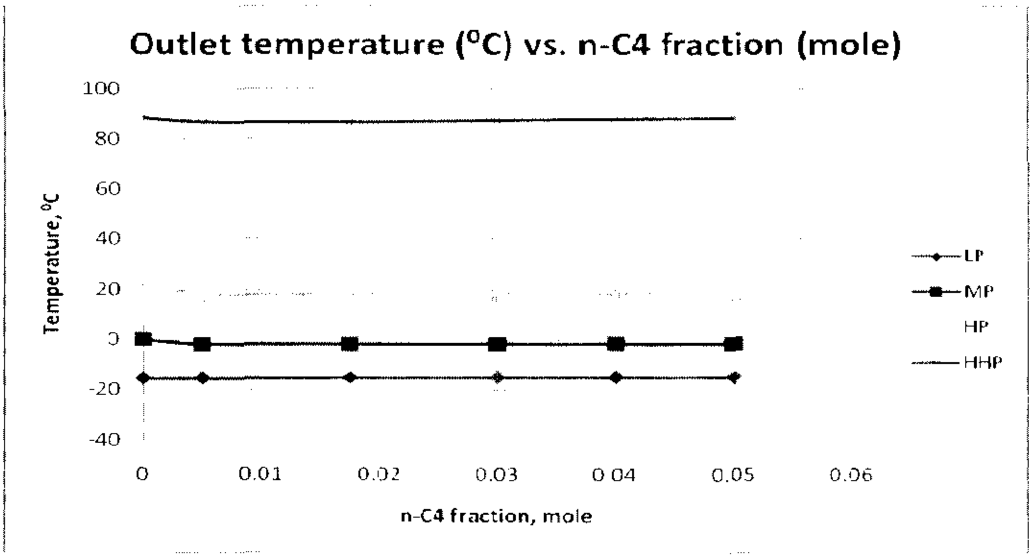


Figure 28: Graph of outlet temperature vs. n-C4 fraction

The projection of graph above shown for n-C4 is also quite similar to C1 and C2's outlet temperature graph. The lowest inlet temperature achieved is when n-C4 fraction is equal to 0.005 moles.

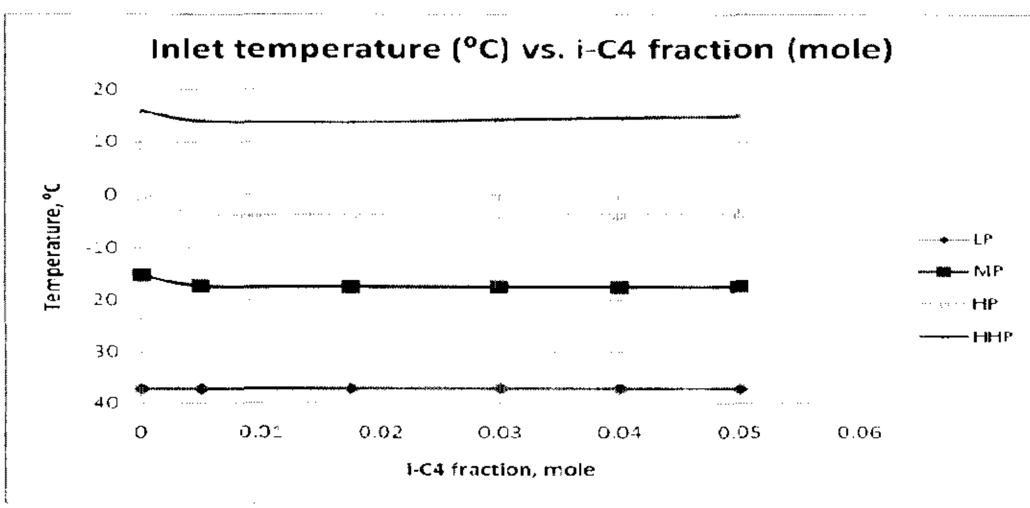


Figure 29: Graph of inlet temperature vs. i-C4 fraction

As for i-butane (i-C4), the graph seen in Figure 29 is exactly the same as n-C4's as the percentage of composition allocated to both components in propane refrigerant is being shared equally.

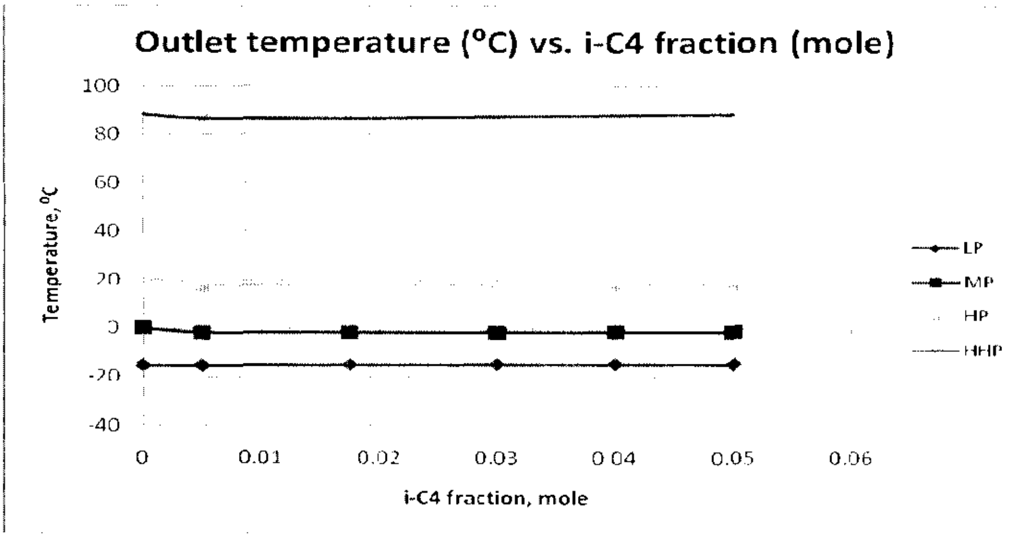


Figure 30: Graph of outlet temperature vs. i-C4 fraction

The scenario in the inlet temperature for i-C4 also applies the same for the graph above.

In conclusion, the inlet and outlet temperature graphs trend for methane (C1), ethane (C2), n-butane (n-C4) and iso-butane (i-C4) are quite similar. Pure propane has the highest inlet and outlet temperature, nevertheless it is impractical to obtain pure refrigerant without any other impurities. However, the trend projected in C3 fraction is the opposite if compared to the other components.

4.3.2 Change in inlet and outlet temperature of compressor

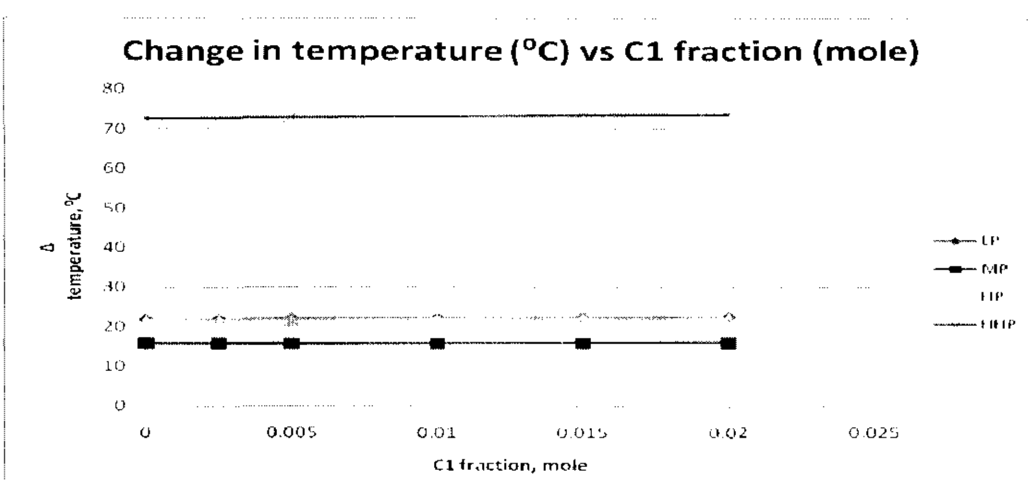


Figure 31: Graph of change in temperature vs. C1 fraction

Figure 31 shows the change in temperature for C1 fraction increases as the fraction increases for all four stages of pressure. However, the changes are the most for HHP pressure, followed by LP, HP and MP.

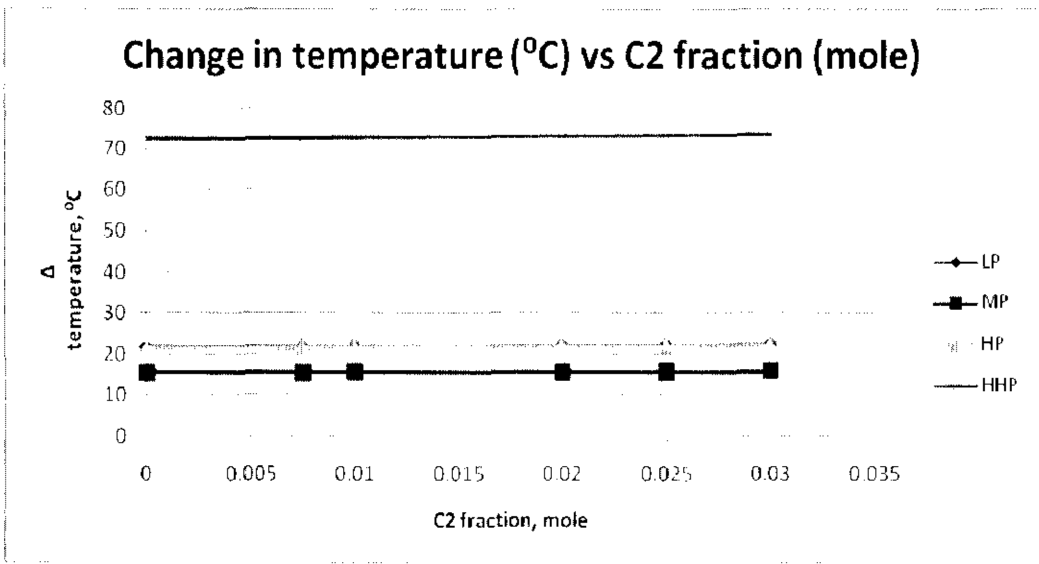


Figure 32: Graph of change in temperature vs. C2 fraction

For C2 fraction, the trend of graph projected above is the same as in C1 fraction.

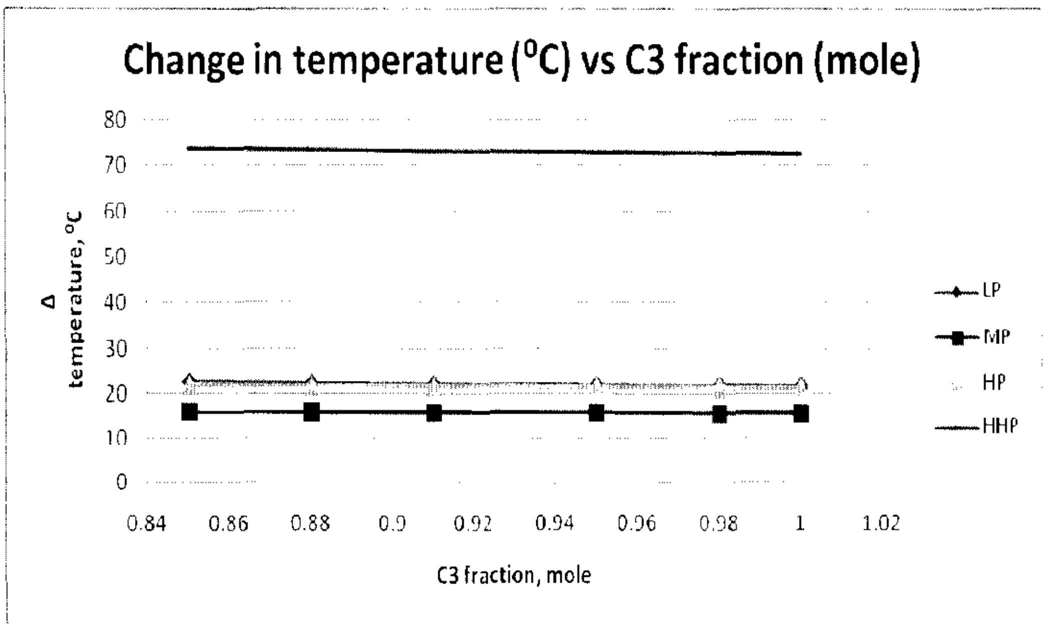


Figure 33: Graph of change in temperature vs. C3 fraction

However, for C3, it is the opposite if compared to the other components. We can see in Figure 33 that as the fraction of C3 increases, the change in temperature decreases.

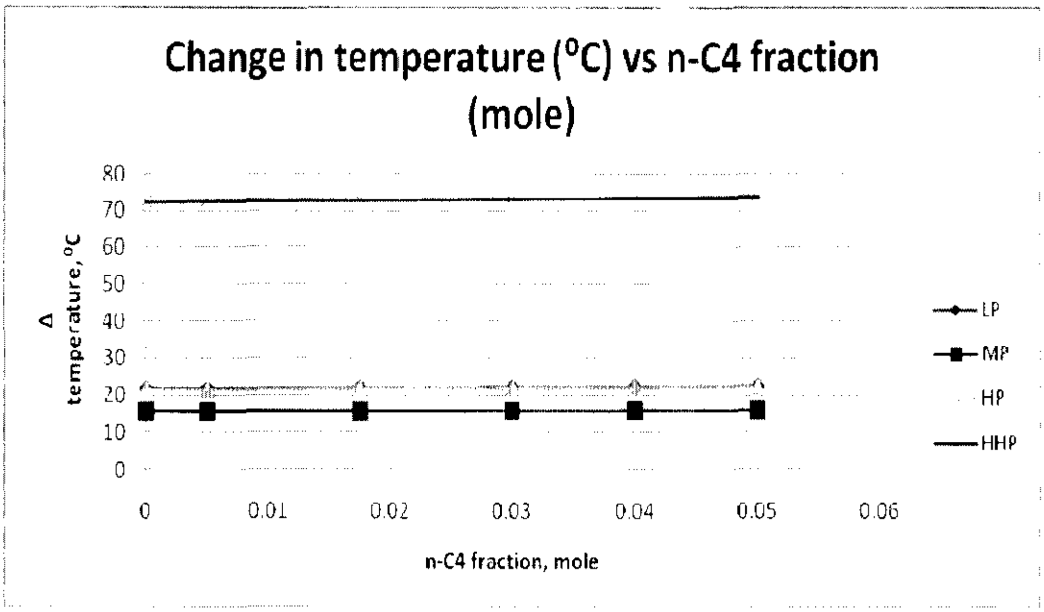


Figure 34: Graph of change in temperature vs. n-C4 fraction

From Figure 34, we can observe the same scenario of C1 and C2 fraction also occurs in n-C4 fraction. As the fraction of n-C4 increases, the change in temperature also increases.

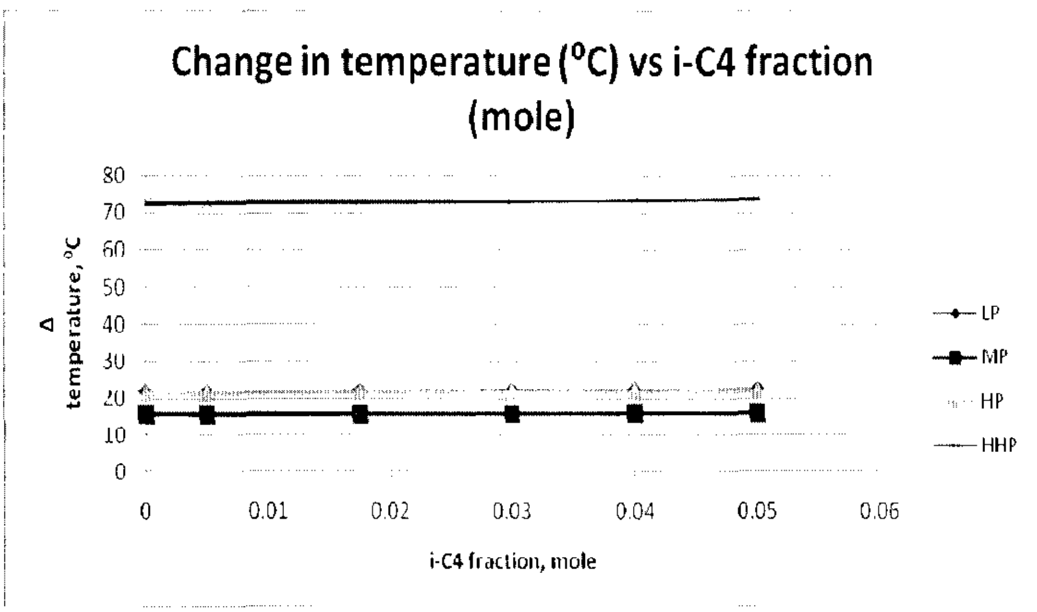


Figure 35: Graph of change in temperature vs. i-C4 fraction

This applies the same in graph of i-C4 fraction above as to what is shown in C1, C2 and n-C4 graphs.

We can conclude that the change of inlet and outlet temperature of compressor for each fraction increases at all stages with HHP with the most, followed by LP, HP and MP as the component's fraction increases, except for C3 component where the change in temperature decreases as C3 fraction increases. This means the purer the propane refrigerant, the least power needed in fin fan to cool down the refrigerant. It is also logical to explain that HHP has the highest change in temperature as it requires the largest change in pressure from HP. This followed by LP where it is the starting point in increasing the pressure of the refrigerant to the required stage. The change of temperature is the lowest for MP as it can be assisted to the required pressure by HP.

4.3.3 Change in power requirement of compressor with the change in propane refrigerant composition

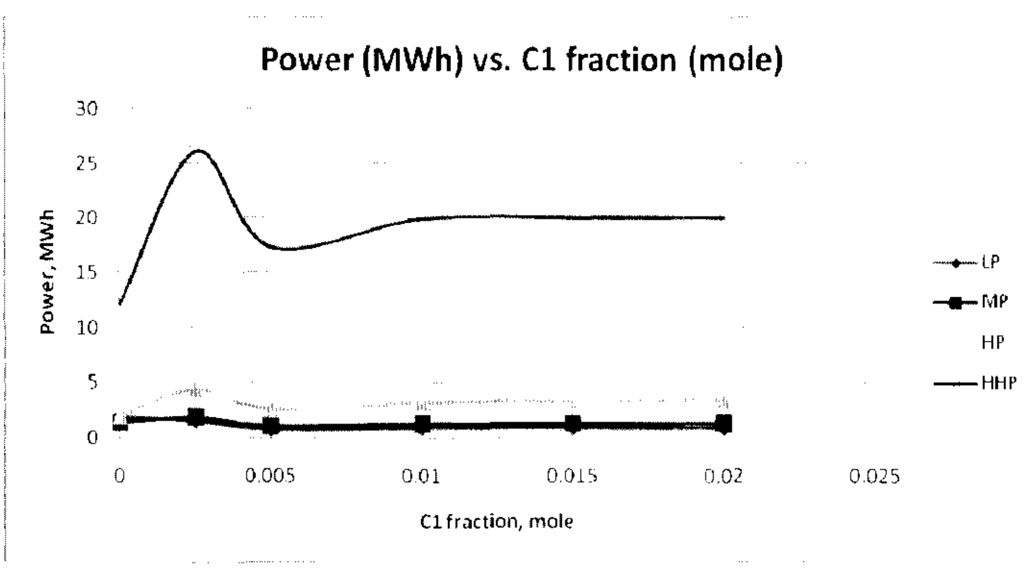


Figure 36: Graph of power vs. C1 fraction

The power requirement of compressor increases and decreases before it increases to remain constant at a certain level as the fraction of C1 increases as shown in the graph above. The change is drastic only in HHP stage, while the rest of the pressure stages are least obvious. The power required is the most in HHP, followed by HP, MP and LP. The highest power required is when C1 fraction is equal to 0.0025 moles and the lowest is at pure propane.

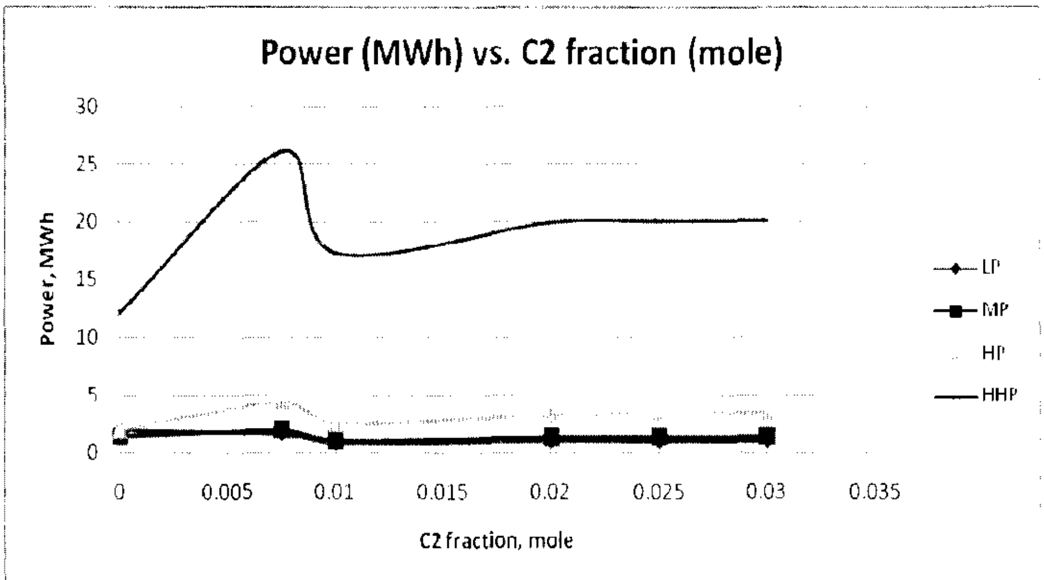


Figure 37: Graph of power vs. C2 fraction

The trend projected in C2 fraction seen in Figure 37 is almost the same as in C1 fraction. The highest power required is when C2 fraction is equal to 0.0075.

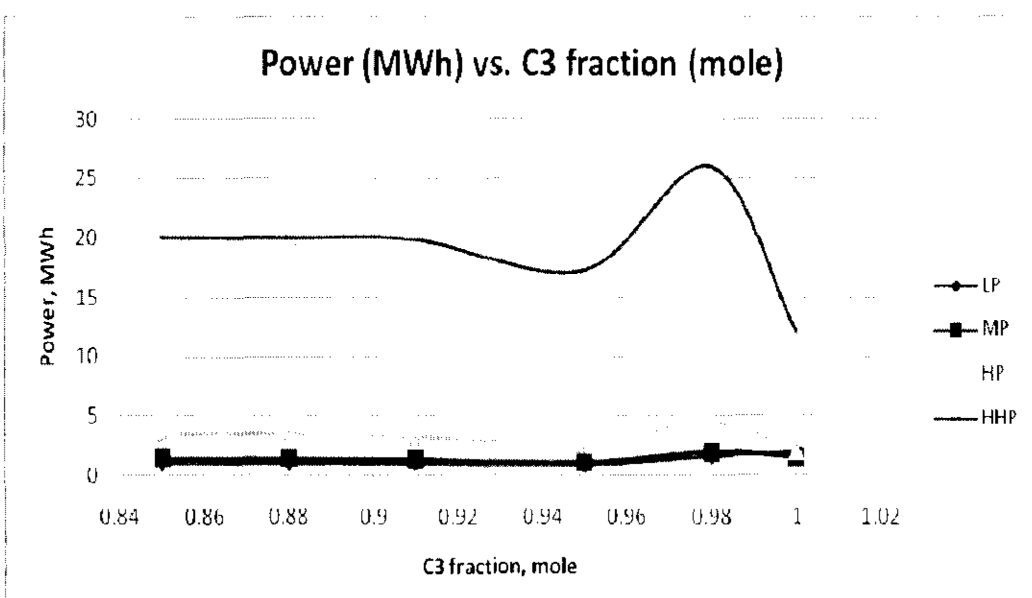


Figure 38: Graph of power vs. C3 fraction

In C3 fraction, the curve of the graph above is the opposite as of C1 and C2 fraction. It remains constant as the fraction of C3 increases before it decreases and increases to its highest level at C3 fraction equal to 0.98 moles. It hit the lowest level of power requirement when the composition of pure propane refrigerant. However, as mentioned above, it is impractical to be achieved in real plant application.

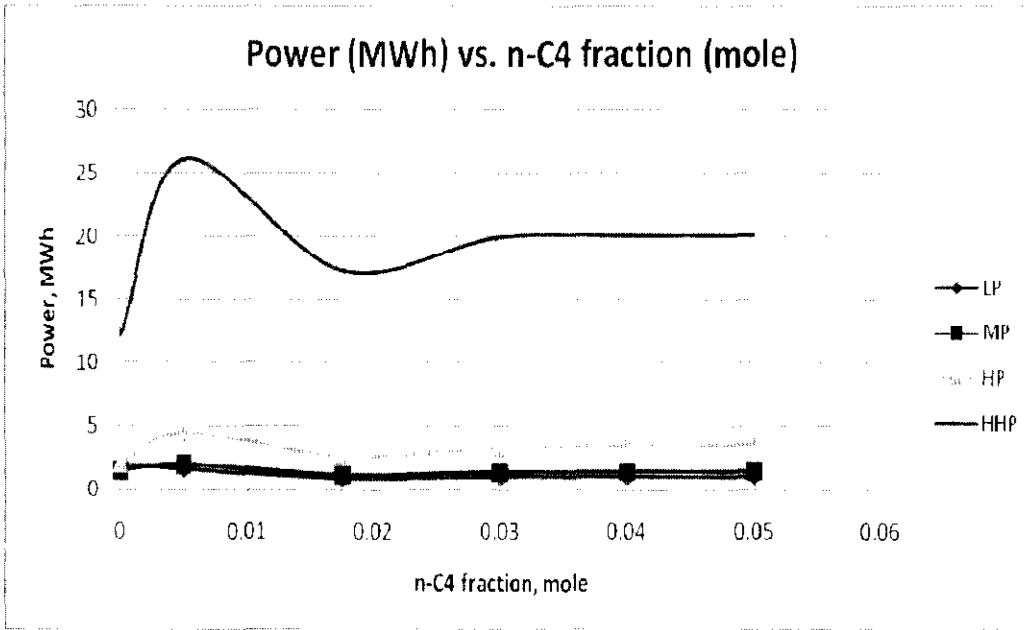


Figure 39: Graph of power vs. n-C4 fraction

The output of the graph above is almost similar to the graph of power requirement of C1 and C2 fraction. The highest power required is when n-C4 fraction is 0.005 moles.

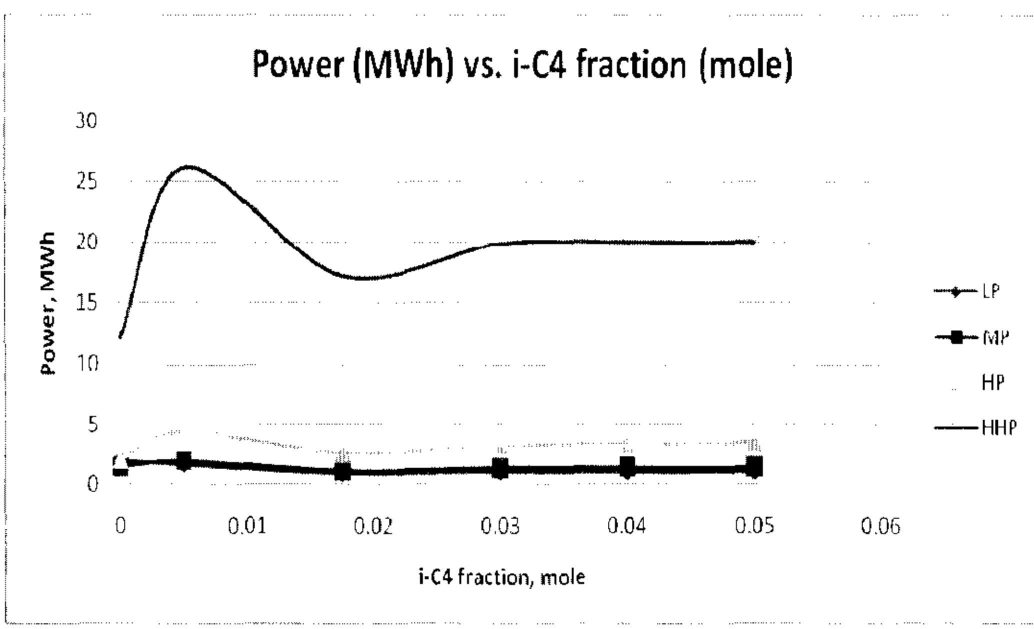


Figure 40: Graph of power vs. i-C4 fraction

The graph above is the same as the graph of power vs. n-C4 fraction, with the fraction of i-C4 equal to 0.005 moles requiring the highest power.

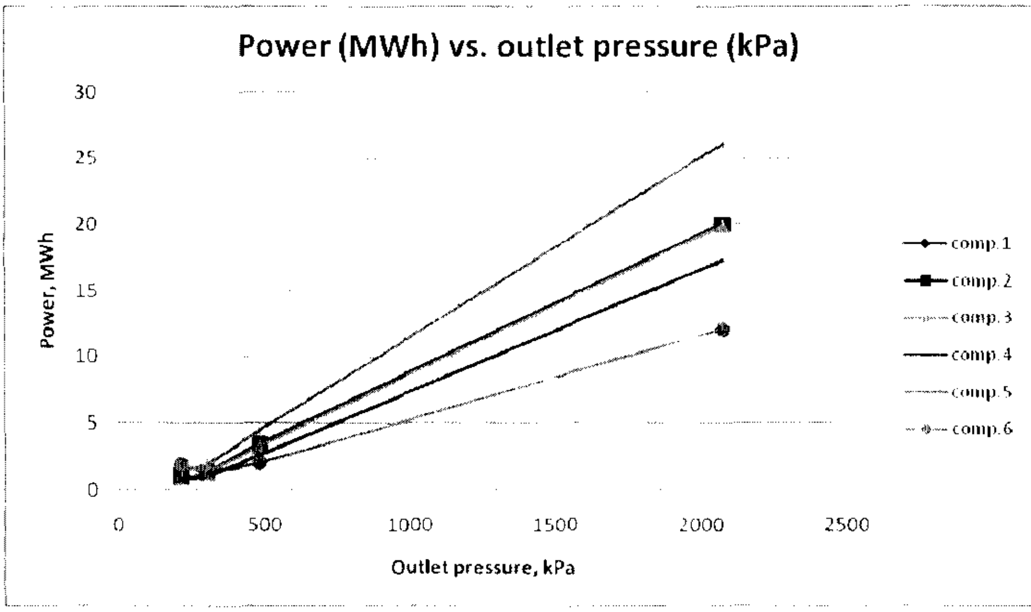


Figure 41: Graph of power vs. outlet pressure at different composition

To conclude, a graph of power requirement of compressor vs. the outlet pressure (four stages of pressure; LP, MP, HP and HHP) based on the six different cases of propane refrigerant composition is plotted to make comparison and choose the best choice of composition with the least power required. From the graph of Figure 41, we can see that as the pressure increases, the power required also increases linearly. The power required to run the propane compressor is the highest at composition 5, where C1 = 0.0025, C2 = 0.0075, C3 = 0.98, n-C4 = 0.005 and i-C4 = 0.005. Though the power required is the lowest when using pure propane, due to its impracticability, the next optimum choice of the lowest power requirement is composition 4, where C1 = 0.005, C2 = 0.01, C3 = 0.95, n-C4 = 0.0175 and i-C4 = 0.0175.

4.3.4 Cost of HP fuel gas usage in compressor

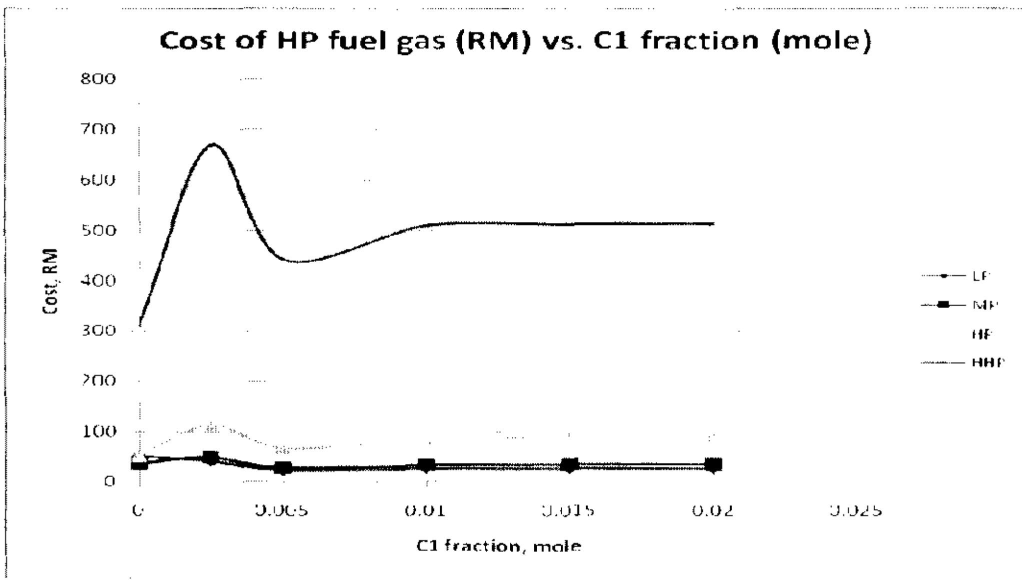


Figure 42: Graph of cost of HP fuel gas vs. C1 fraction

The trend of the graph projected above is similar to the graph plotted in power vs. C1 fraction. When C1 fraction is 0.0025 moles, the cost of HP fuel gas to run the compressor is the highest while lowest when it is pure propane refrigerant which is not practical. Hence the second lowest choice of option is when C1 fraction is at 0.005 moles.

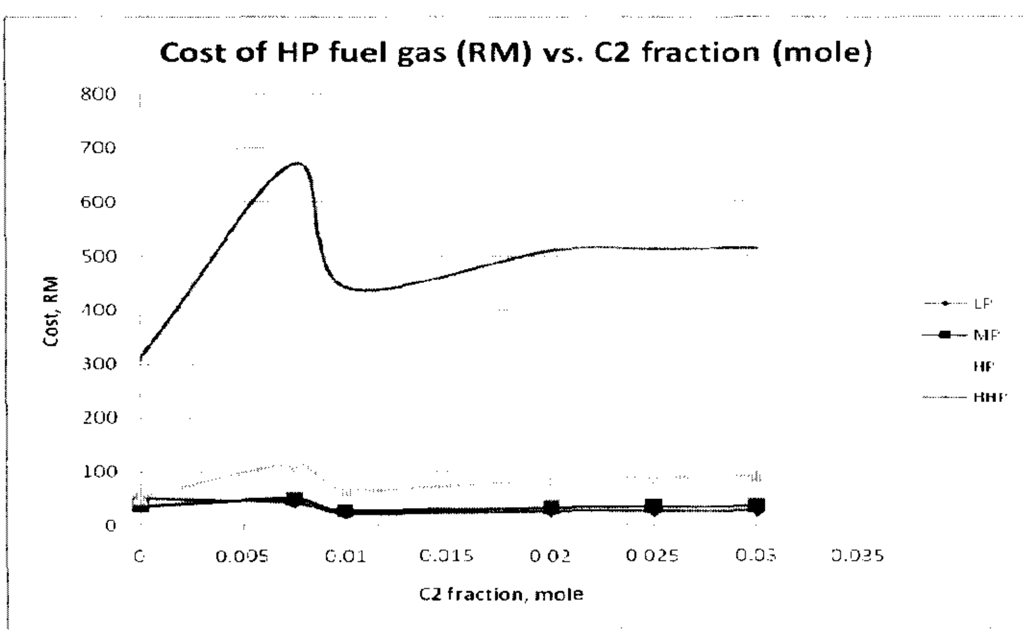


Figure 43: Graph of cost of HP fuel gas vs. C2 fraction

The same situation applies the same for C2 fraction. In Figure 43, the cost of HP fuel gas is the highest when the fraction of C2 is 0.0075 moles and lowest at 0.01 moles of feasible composition.

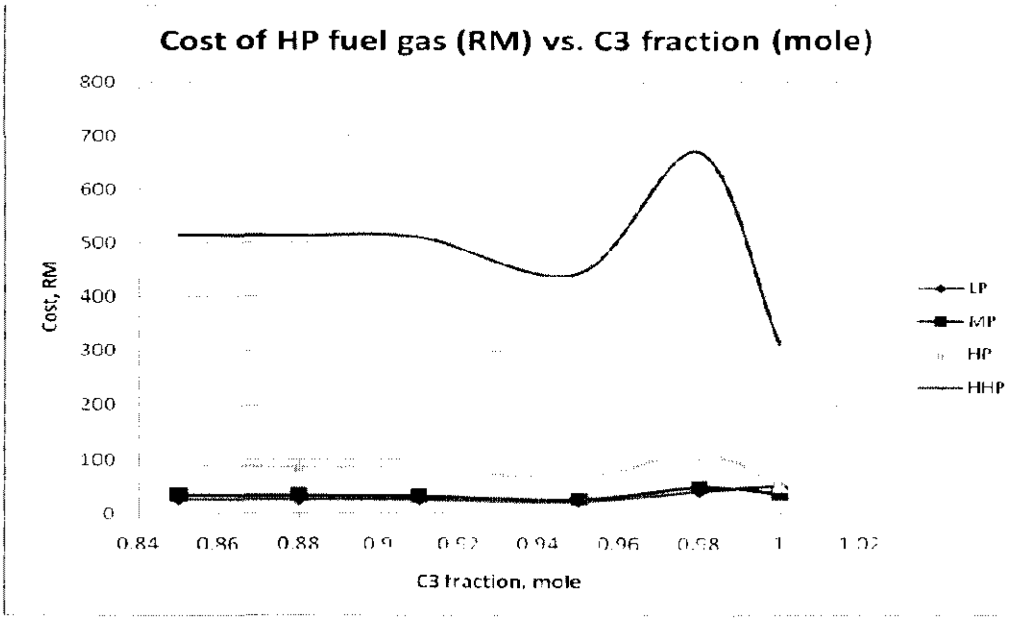


Figure 44: Graph of cost of HP fuel gas vs. C3 fraction

The above graph shows that the cost for HP fuel gas is the highest when C3 fraction is 0.98 moles. The lowest cost of applicable composition is when C3 is at 0.95 moles.

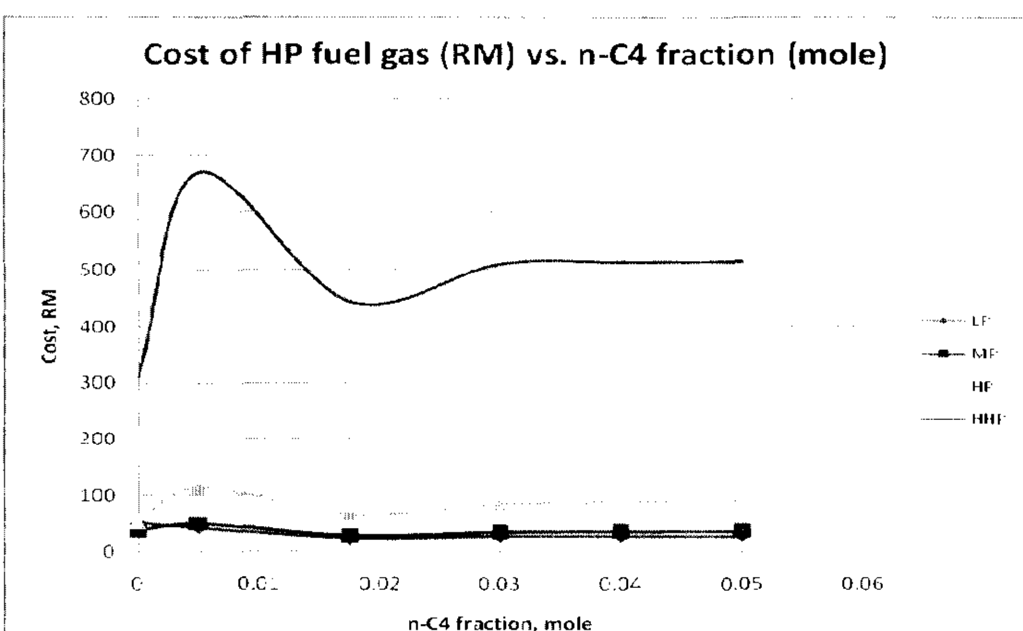


Figure 45: Graph of cost of HP fuel gas vs. n-C4 fraction

The highest cost of HP fuel gas in running the compressor is when n-C4 fraction of 0.005 moles while the lowest is at 0.0175 moles as observed in Figure 45.

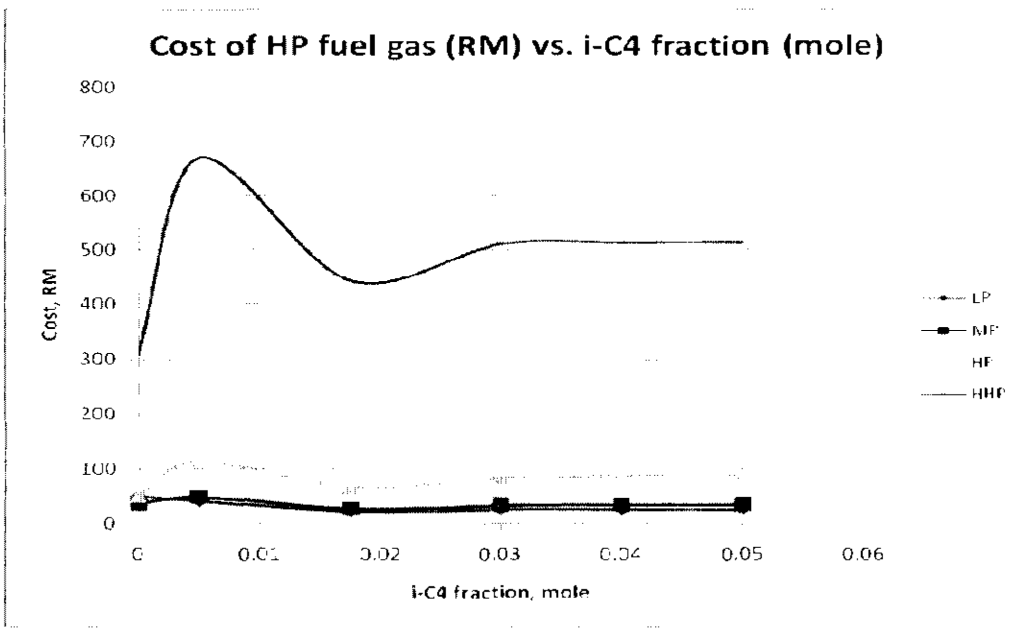


Figure 46: Graph of cost of HP fuel gas vs. i-C4 fraction

From Figure 46, the cost of HP fuel gas is the highest when i-C4 is 0.005 and lowest when its fraction is 0.0175 moles. It is the same as n-C4.

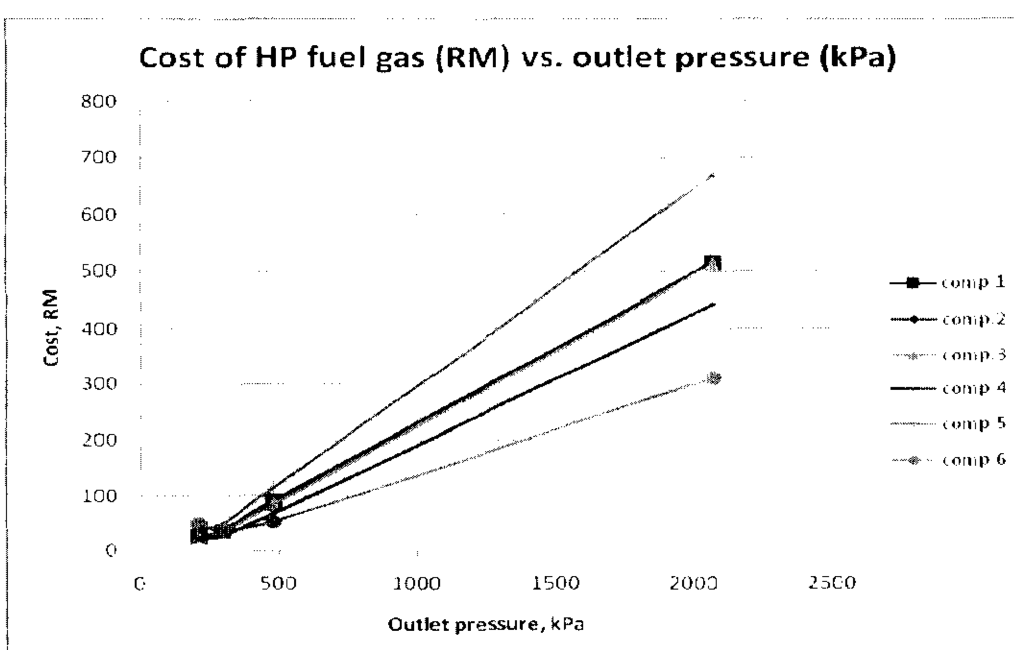


Figure 47: Graph of cost of HP fuel gas vs. outlet pressure at different compositions

In order to select the best choice of composition, a graph of the cost of HP fuel gas vs. the outlet pressure on each of the different cases of composition is plotted. The graph shows a linear increase for all compositions. We can also conclude that the cost of HP fuel gas is the highest at HHP level, followed by HP, MP and LP. From the graph above, we can see that by using pure propane refrigerant, the cost to run the compressor using HP fuel gas is the lowest. Only a slightly higher cost at LP level if compared to the other compositions at the same level of pressure. However, as mentioned before, due to its impracticality, the next optimum choice is composition 4. The cost of HP fuel gas usage is the highest at composition 5.

4.3.5 Cost of compression for propane refrigerant use in compressor

The trend of the graphs for cost of compression for propane refrigerant used at different stages is quite similar to the graphs plotted in power requirement of compressor and the cost of HP fuel gas in running the propane compressor. However, we can see that the cost of LP propane is slightly higher if compared to MP propane. This followed by HP and HHP.

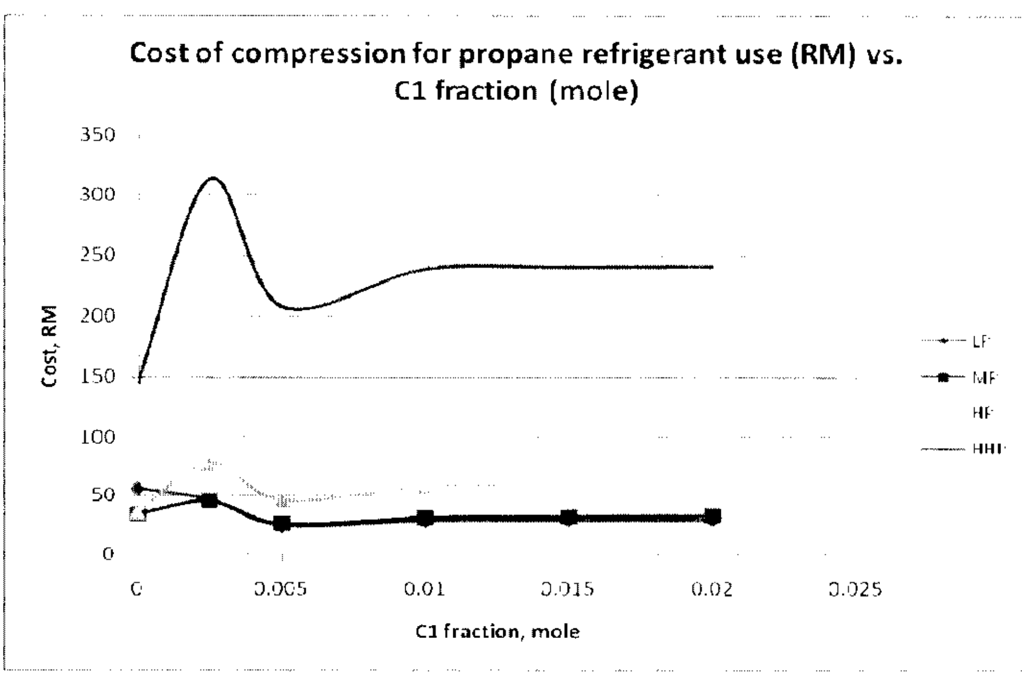


Figure 48: Graph of cost of compression for propane refrigerant use vs. C1 fraction

For the cost of compression for propane refrigerant use in C1 fraction shown above, the cost is the highest when C1 fraction is 0.0025 moles while lowest when its fraction is 0.005 moles.

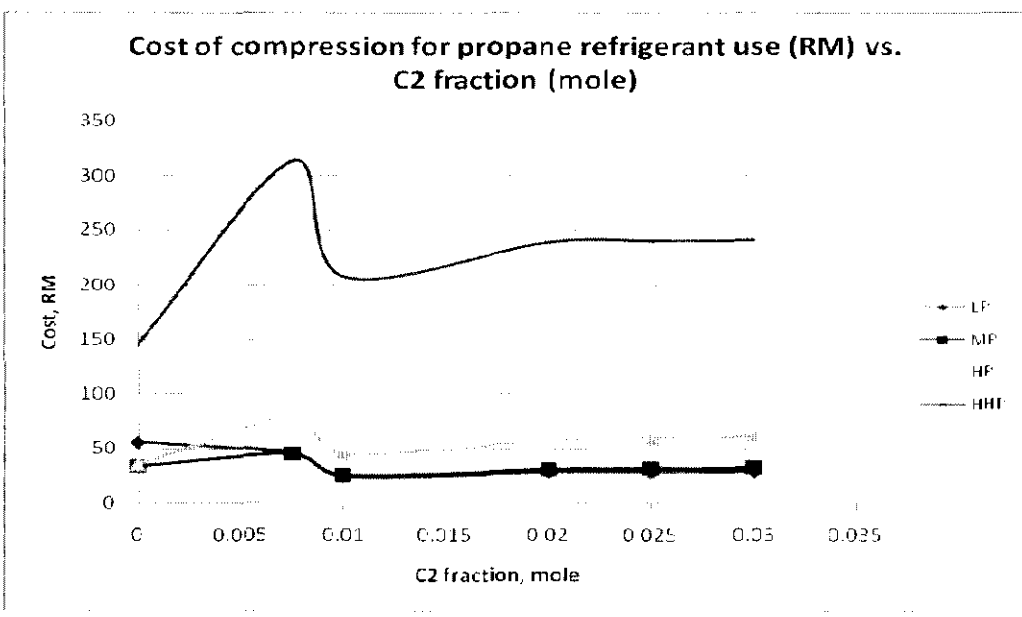


Figure 49: Graph of cost of compression for propane refrigerant use vs. C2 fraction

As for C2 fraction, the cost is the highest when the fraction is 0.0075 moles and lowest at 0.01 moles.

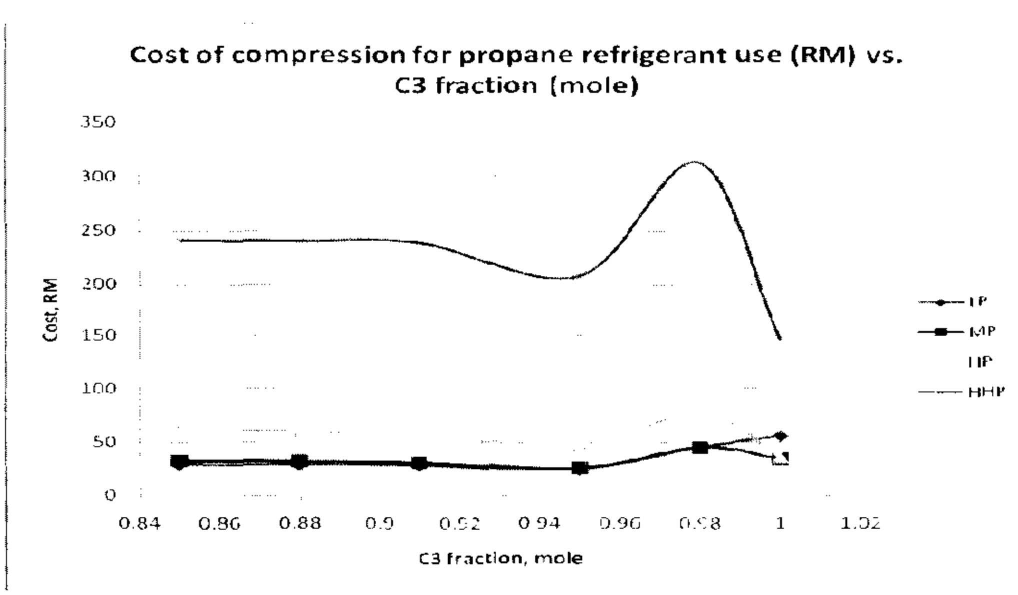


Figure 50: Graph of cost of compression for propane refrigerant use vs. C3 fraction

From the graph above, the lowest cost of compression for propane refrigerant use is when C3 fraction is 0.95 moles. The highest cost is when C3 fraction is at 0.98 moles.

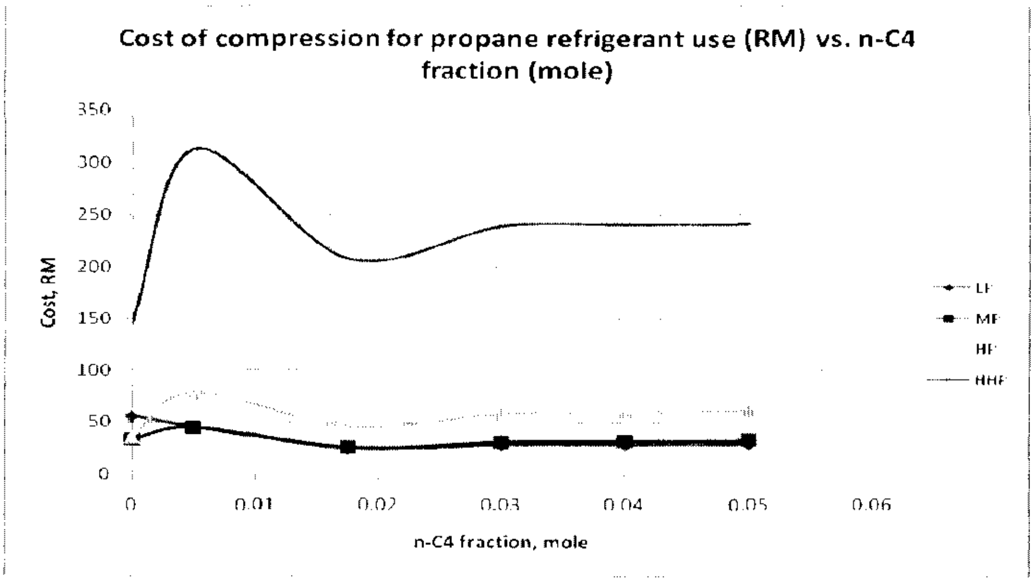


Figure 51: Graph of cost of compression for propane refrigerant use vs. n-C4 fraction

For n-C4, the cost is the highest at the fraction of 0.005 moles and lowest at 0.0175 moles as seen in Figure 51.

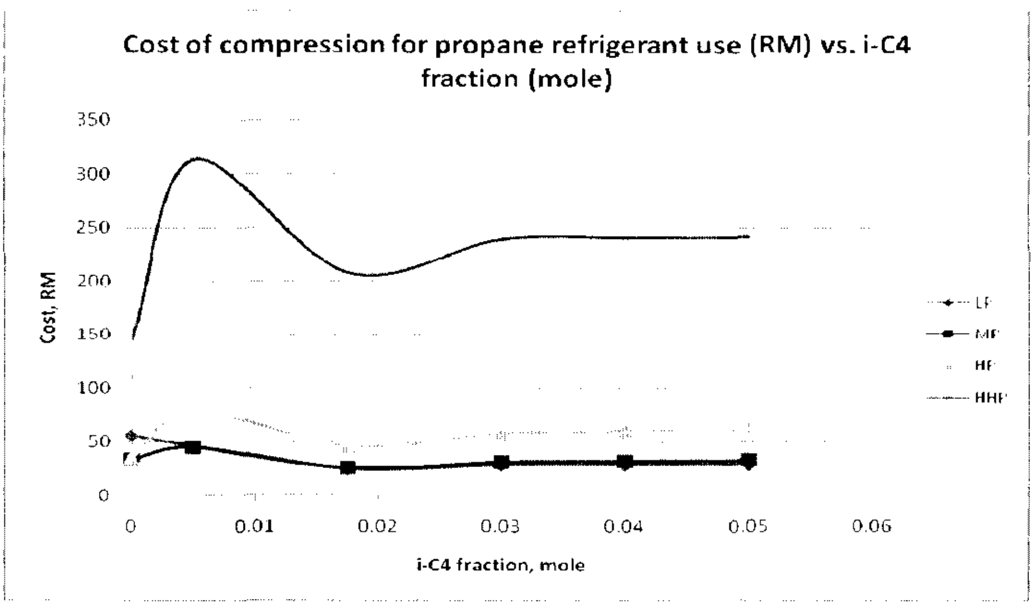


Figure 52: Graph of cost of compression for propane refrigerant use vs. i-C4 fraction

In Figure 52, the fraction of i-C4 is the same as n-C4 at the highest and lowest cost of compression for propane refrigerant use.

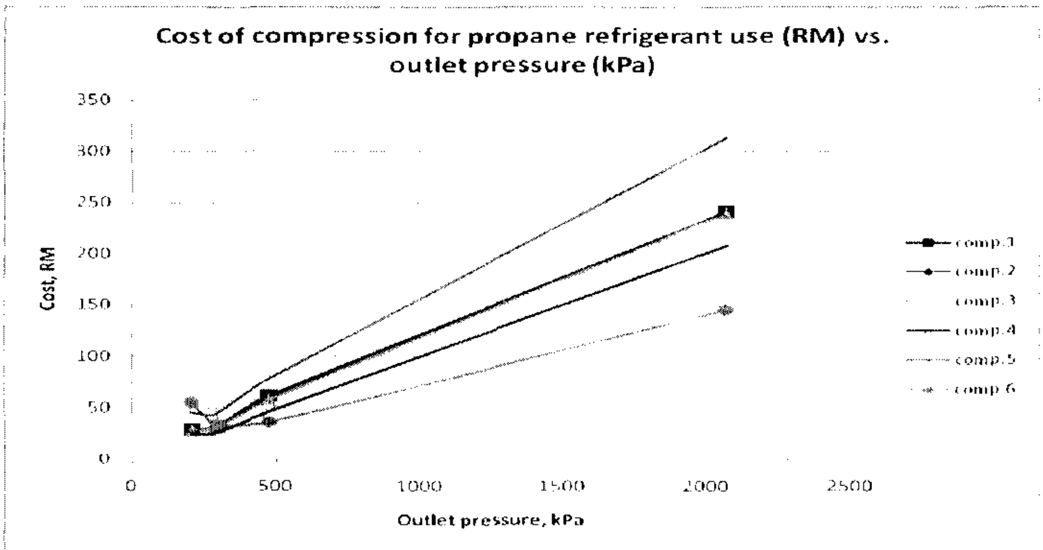


Figure 53: Graph of cost of compression for propane refrigerant use vs. outlet pressure at different compositions

The interpretation of result of the above graph is the same as the one in the cost of HP fuel gas. The cost is the highest at composition 5, while the lowest is at composition 4. As the pressure increases, the cost of compression for propane refrigerant use also increases in a linear trend.

4.4 Summary for the best composition

Table 14: Comparison between different compositions with different study parameters

Composition	1	2	3	4	5	6 = pure propane (infeasible)
Minimum change in temperature					√	
Minimum power usage				√		
Minimum cost of HP fuel gas				√		
Minimum cost of compression for propane refrigerant use				√		

The table above summarises the best composition of propane refrigerant usage. With the minimum change in temperature, it can help to reduce the power required at fin fan to cool down the refrigerant. However, as the focus lies within the selection of the minimum operating cost in the compressor for the plant, composition 4 is the most optimum choice with C1 = 0.005, C2 = 0.01, C3 = 0.95, n-C4 = 0.0175 and i-C4 = 0.0175. The worst choice of composition is composition 5, with C1 = 0.0025, C2 = 0.0075, C3 = 0.98, n-C4 = 0.005 and i-C4 = 0.005 as it requires high usage of power and cost.

From the comparision in terms of cost, the usage of individual propane refrigerant use at different level of pressure requires the lowest cost if compared to using HP fuel gas.

CHAPTER 5

CONCLUSION AND RECOMMENDATIONS

CONCLUSION

Process technology for liquefaction of natural gas is undergoing continuous improvement to meet with the increasing demand of LNG. This has demanded LNG plants to be volume flexible to match with market demand. Thus process evaluation techniques are required to identify process improvements which can be apply on existing plants to lower the operation cost, expand its capacity to increase LNG production. Selecting of the optimum process cycle is indeed crucial to ensure the capacity of the plant will be fully utilised and run in cost effective manner. The unique characteristics of an LNG plant will be operating it at an optimum thermal efficiency without neglecting the economics merits and environmental impacts.

APCI propane pre-cooled mixed refrigerant process (C3MR) has been selected as the basis of simulation in HYSYS. This is due to a study obtained from an LNG journal stating that in 2012, there will be a total of 100 LNG trains throughout the globe with 86 of them that will be utilising the APCI technology. This clearly indicates the reliability and importance of APCI technology in the LNG industry. The focus of study in the power consumption had been narrowed down to propane refrigerant system to determine the best composition for each fraction at which the least power and cost are required. From the study, the best recommended composition of propane refrigerant is composition no. 4, while the worst choice of composition is composition no. 5. Hence, by using composition no. 4, it requires the least power with the lowest cost. Table 12 summarises the fraction of components for composition no. 4 and 5. Nonetheless, it is best to operate the compressor using the individual refrigerant use, as it is the cheapest method if compared to using HP fuel gas.

Table 15: Fraction of components for composition no. 4 and 5

Components	C1	C2	C3	n-C4	i-C5
COMPOSITION 4	0.005	0.01	0.95	0.0175	0.0175
COMPOSITION 5	0.0025	0.0075	0.98	0.005	0.005

RECOMMENDATIONS

To further achieve the objective of the project which is to reduce the operation cost while increasing the production of LNG, this requires the identification of modifications or improvement means to retrofit the existing facilities and only requires addition of minor equipment. This includes optimisation in the operating conditions, enhancement in equipments' efficiencies, machinery addition and others.

Factors that can be taken into considerations for the optimisation in operating conditions are:

- The composition of refrigerant

  In general increasing the molecular weight of the compression fluid increases compressor capacity. This is because at a given mass flow, the volume occupied by higher molecular weight refrigerant is lower than refrigerant of lower molecular weight. Having an additional capacity margin on compressor the operating pressure of the MCHE system can be further reduced to increase LNG production.

- The operation of compressor

  It is important to use the compressor curve to assess the compression performance and the available capacity that can be utilised to increase the cooling capacity of refrigerant. From the compressor's performance curve, we can determine the optimum inlet volume to the compressor to obtain the required discharge pressure and achieve the best efficiency possible for the type of compressor used. Operating at too low pressure results in low suction pressure will cause stonewall operation at which point the compressor becomes inefficient, while high pressure operation may result in surge and trip the system. However, a few parameters can be manipulated to decide on the favourable operating condition of the compressor. These include the molecular weight of the component to be compressed, the inlet temperature, and the compressibility factor of component, the specific heat ratio, and the speed of compressor based on the surrounding weather condition.

Apart from that, in terms of enhancement in equipments' efficiencies, it is crucial to pay more attention at the major and costly equipments, such as the compressors, turbines, heat exchangers and others that are being greatly utilised and of high importance in the whole liquefaction process. By running the equipment in an optimum condition can ensure that the whole plant will run smoothly, thus reducing the unnecessary maintenance cost. Besides, with the latest development in APCI technology, the AP-X technology had shown its credibility in increasing the production of LNG through the addition of nitrogen expander. This machinery addition method can be one of the step taken by the existing plant in improving the overall production of the plant.

REFERENCES

1. Report #:DOE/EIA-0484(2009), EIA – International Energy Outlook 2009 > Natural Gas

2. Adrian J Finn, Grant L Johnson and Terry R Tomlinson, "LNG Technology for Offshore and Mid-Scale Plants", Costain Oil, Gas & Process Limited, Manchester, UK.

3. Anthony E., Rick H., Allyn R., Phil H., Amos Avidan and John D., "Lowering LNG Unit Costs through Large and Efficient LNG Liquefaction Trains – What is the Optimal Train Size?", New Orleans, LA., 2004.

4. Paul Bosma, Rob Klein Nagelvoort, "Liquefaction Technology; Development through History", 2009

5. Dr Tariq Shukri, Foster Wheeler, "LNG technology selection", UK, February 2004.

6. Valerie Rivera, Ayema Aduku, Oluwaseun Harris, "Evaluation of LNG Technologies", 2008

7. James T. Jensen, "The Development of a Global LNG Market", Oxford Institute for Energy Studies, 2004.

8. Dr. Mark Pillarella, Dr. Yu-Nan Liu, Joseph Petrowski and Ronald Bower, "The C3MR Liquefaction Cycle: Versatility for a Fast Growing, Ever Changing LNG industry", Air Products and Chemicals, Inc. Allentown, Pennysylvania.

9. Charles Durr, Christopher Caswell and Heinz Kotzot, "LNG Technology for the Commercially Minded – The Next Chapter".

10. Pierre-Yves Martin, Jerome Pigourier and Beatrice Fischer, "Natural Gas Liquefaction Processes Comparison", France.

11. ChE 3G04 Winter 2004 McMaster University, "HYSYS Tutorials"

12. Doug Yates, P.E, Thermal Efficiency – Design, Lifecycle, and Environmental Considerations in LNG Plant Design, GASTECH 2002

13. F. P. Ross, S. T. Walther, K. T. Cuellar, "Advanced technologies provide improved economics for liquefaction natural gas facilities."

14. APX Energy Viewpoints, "LNG-Prospects for the Future", 2005

15. Mark R. Pillarella, James C. Bronfenbrenner, Yu-Nan Liu, Mark J. Roberts, "Large LNG Trains: Developing the Optimal Process Cycle", Air Products and Chemicals, Inc. Allentown.

16. Operating Manual and Process Flow Sheet (PFS) U91400 Liquefaction Unit for Malaysia LNG Tiga Plant Project, Bintulu, Sarawak, Malaysia.

17. J. S. Swearingen, "Staged Isenthalpic-isentropic Expansion of Gas From a Pressurised Liquefied State to a Terminal Storage State", 1969

18. Yu-Nan Liu, Emmaus; Glenn E. Kinard, Allentown; David A. Zagnoli, Macungie, "Gas Liquefaction Process Control System", 1991

19. Mory Houshmand; Kimberly A. Kruger; Gerald W. Alves; Ricardo Ostaszewski; Noureddine Belhateche, "LNG Production In Cryogenic Natural Gas Processing Plants", 1994

20. Dag-Erik Helgestad, "Modeling and optimisation of the C3MR process for liquefaction of natural gas", December 10, 2009