CERTIFICATION OF APPROVAL

Structural Arrangement of End Flash System for LNG Production

by

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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 the original work contained herein have not been undertaken or done by unspecified sources or persons.

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(CHIA BOON CHUNG)

ABSTRACT

End flash system is important in LNG Chain as it increases LNG throughput by reducing the cooling load in Main Cryogenic Heat Exchanger (MCHE). One of the main problems encountered in industrial installations of end flash system is related in particular to the optimum use of the compression apparatus which represents a significant investment, both in terms of initial purchase and in terms of power consumption. The energy needed for the typical end flash system is approximately 995 kJ for the production of 1 kg of LNG. In this work, simulations on the structural arrangement of end flash system were conducted in the aim of reducing the specific power of production of LNG. The energy required to produce 1 kg of LNG and machinery power consumed by the end flash system are investigated. Two optimization cases, namely pure refrigerant cascade and mixed refrigerant cascade, have been studied and analyzed from technical and economical aspect. Results show that the base case needs 351 kJ to produce 1 kg of LNG while the mixed refrigerant cascade consumes only 273 kJ for the same purpose. Besides that, mixed refrigerant cascade gives higher LNG rundown at 542 t/h compared to the base case which gives only 540 t/h of LNG. In terms of economic analysis, the additional LNG rundown yields additional revenue of USD 13.6 million. The profitability analysis of installation of mixed refrigerant cascade over the base case shows positive feedback as ROI is 20.7% and IRR is 17.8% which is higher than the MARR. The study develops a new process which allows significant increase in LNG production with lower energy consumption as compared to the current end flash system.

ACKNOWLEDGEMENT

The author strongly understands that the success of this project would not be possible without the involvement and contributions of certain parties, organizations and individuals. Therefore, the author would like to take this opportunity to express his deepest gratitude to each and every one of them.

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CHAPTER 1

INTRODUCTION

1. INTRODUCTION

1.1 Background of Study

Natural gas (NG) is mainly made up of methane and varied amount of ethane, propane, butane and pentane. Under atmospheric conditions, it is colorless and odorless in nature. Natural gas is considered as a main source of energy in the near future as it is able to generate large amount of energy through combustion. Table 1.1 shows the compositions of natural gas in molar percentage and weight percentage as suggested by Paradowski (2005). Owing to the fact that natural gas is highly combustible when exposed to flame or sparks, it is a normal practice that gas companies added mercaptan to the natural gas so that individuals are able to detect even the smallest leak of natural gas (Sapuan, 2008).

Total world energy use rises from 524 quadrillion British thermal units (Btu) in 2010 to 630 quadrillion Btu in 2020 and to 820 quadrillion Btu in 2040 (EIA, 2013). Energy Information Administration states that the world natural gas consumption and production are expected to increase by 64 percent from 113 trillion cubic feet in 2010 to 185 cubic feet in 2040. In near future, Asia is expected to become the world top gas consumer overtaking the spot from North America (Ramli, 2009).

Table 1.1: Compositions of Natural Gas

	Molar %	Weight %
Nitrogen	0.10	0.16
Methane	91.40	81.37
Ethane	4.50	7.51
Propane	2.50	6.12
i-Butane	0.60	1.94
n-Butane	0.90	2.90

Due to environmental soundness and multiple application of natural gas across all section, it is expected that natural gas will play an important role in meeting the energy demand of the world in the future. Nevertheless, transportation to distant market is not always economically wise or technically feasible through gas pipelines. Hence, natural gas liquefaction has emerged as plausible economically and technically mature alternative. The economics of liquefying natural gas are obtained by the reduction of natural gas volume upon liquefying and give benefit to storage and transportation in large quantity (Rojey and Jaffrett,1997).

Referring to Figure 1.1, in a typical LNG plant, natural gas is first treated with acid gas removal to remove impurities such as the hydrogen sulfide and carbon dioxide as the presence of these acid gases will cause corrosion to the pipeline, reduce heating value of natural gas, freeze and create solid in cryogenic process. Normally the maximum allowable concentration for carbon dioxide is 50 parts per million by weight (ppmw) while hydrogen sulfide is 4 ppmv. After the feed pre-treatment process, the treated natural gas will be sent to dehydration unit where water is removed from the feed gas. Removal of water is crucial in preventing freezing of water in cryogenic process. The dried natural gas is then sent to mercury removal unit to reduce the concentration of mercury to $0.01 \mu g$ per 1 Nm^3 of natural gas. Removal of mercury is essential to prevent corrosion at the downstream of the process and to ensure a clean product as mercury is poisonous to human being. Liquefaction unit forms the central element in the LNG

supply chain. The natural gas which has primarily methane will be liquefied by Main Cryogenic Heat Exchanger (MCHE). Typically, the liquefaction condition varied from place to place and is normally ranging from -120℃ to -170°C, and pressure of between 1 to 60 bar (Sapuan, 2008).

At the outlet of liquefaction, the liquefied natural gas (LNG) is sent to an end flash system. The idea of end flash system is relatively new and its installation is optional in LNG supply chain. One of the advantages of end flash system is that it improves the heat transfer in MCHE. The installation of end flash system offers expansion cooling to the LNG stream to desired temperature, thus allowing a warmer LNG stream to escape liquefaction unit. Such effort allows more natural gas to be condensed in the MCHE. For instances, installation of extended end flash in MLNG Dua allows the LNG stream to leave MCHE by 7°C warmer, thus more natural gas is allowed to pass through the exchanger. The extended end flash contributes to 8% of LNG production increment for MLNG Dua (Ibrahim, n.d.). Use of end flash drum in end flash system will produce top distillate stream which will be used as fuel gas for the unit. A point to note is that the light components, such as the nitrogen, oxygen and helium, will escape as distillate thus ensure the quality of the product. The bottom product which is primarily sales LNG will then sent to storage.

Figure 1.2 shows the process flow sheet for typical end flash system. LNG from MCHE at -150.5℃ and 48 bar is let down by Joule Thompson (JT) Valve to 1.15 bar. The expanded stream 1 will in line mixing with stream **17** to yield stream **2** at -159.0℃ and 1.15 bar before fed into end flash drum to produce a vapor phase stream **3** and 521 t/h of bottom product, stream **4.** Stream **3**, which contains 45 t/h of end flash gas will first heat exchange with stream **14** and then undergoes a series of compression and heat exchange. At the end of compression, 79% of end flash gas will be tapped off as fuel gas to drive the compressors of the unit while the remaining 21% will be recovered back to the system as stream **14**. Stream **14** will give up its heat to cool down to -98.2℃ and let down by JT valve to -159.1℃ and 1.15 bar. The end flash gas recycled back as stream **17** by in line mixing with stream **1**.
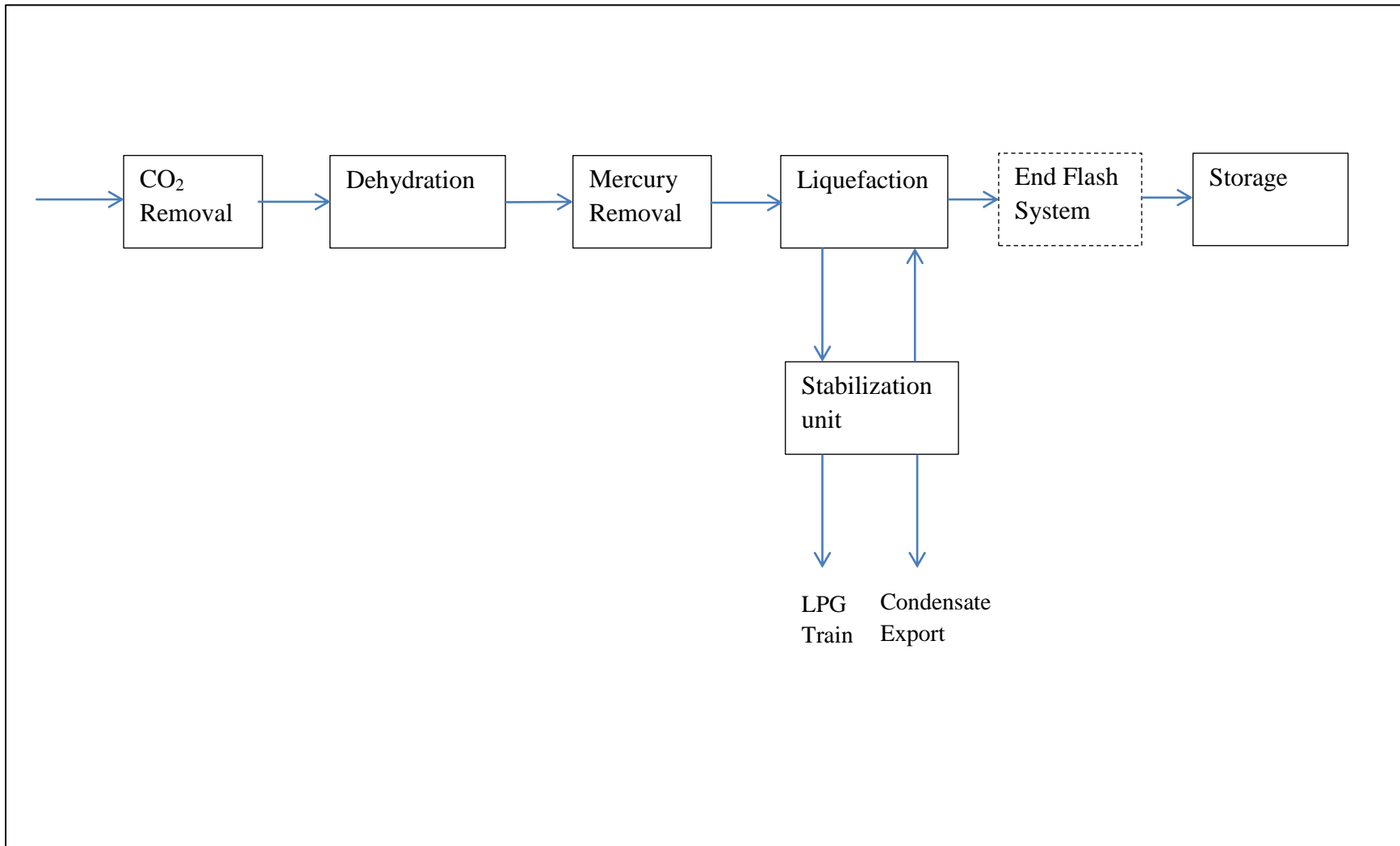
Figure 1.1: Block Diagram for Typical LNG Plant with End Flash System
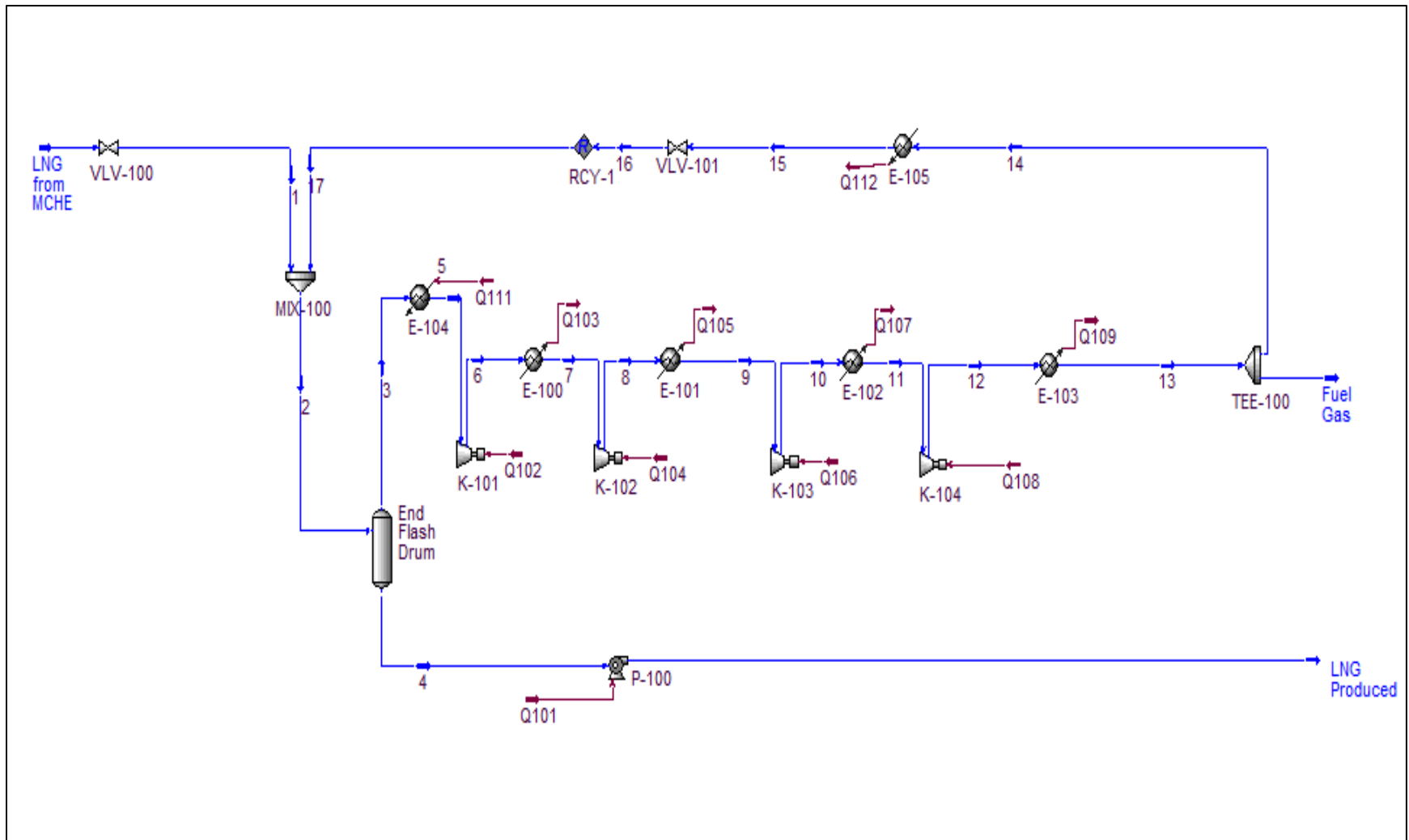
Figure 1.2: Typical End Flash System

1.2 Problem Statement

Though end flash system adds in more value to the LNG supply chain, it is believed that this relatively novel process still offers plenty room for optimization purposes. For instance, 79% of distillate from the end flash drum, which has been treated as fuel gas to drive the compressors in the system, has a composition that is in comparable to the sales LNG. In other words, the current end flash system suffers from the loss of LNG production as 79% of end flash gas, which possess high potential economic value in it, is being used as utility for the plant instead of being sold as main product. The similarity in composition allows the potential sources of fuel gas to be recovered as sales gas, leaving behind the minimum amount of fuel gas needed to drive the compressors in the end flash system.

In addition the current end flash system relies heavily on mechanical cooling which involves few stages of compression and heat exchange, resulting in high cost of machinery power to produce 1 kg of sales LNG. The specific power of LNG production for typical end flash system as suggested by Paradowski (2005) is 995 kJ/kg. The possibility of reducing the duty of compressors by utilizing mixed refrigerant consisting of methane, ethane and propane has been overlooked by process engineers. As a result, the overall cost of operation may be reduced in the long run.

1.3 Objectives

The objectives of this project are:

    i. To maximize the production of sales LNG from end flash system (kg/hr)
   ii. To reduce the specific power of LNG production (kJ/kG)

The engineering project is said to be successful if it is able to increase the amount of recovered end flash gas back to the process as the amount of LNG production increases with increasing recovered end flash gas. The specific power of LNG production measures the amount of energy required to produce 1 kg of sales LNG in 1 hour. Reduction in specific power of LNG production can be done through increasing LNG rundown or reducing the compressor duty.

1.4 Scope of Research Work

In line with the objectives of this project, this paper will focus only in the LNG production and energy consumption in end flash system of LNG plant, particularly in:

- Temperature of LNG at the outlet of MCHE

    - -130^oC
    - -135^oC
    - -140^oC
    - -145^oC
    - -150^oC
    - -160^oC

- Structural Arrangement of End Flash Unit

    - Compressors,
    - Heat Exchangers,
    - Cold Box,
    - Flash Drum

- Composition of Mixed Refrigerant
    - Methane
    - Ethane
    - Propane
    - Butane
    - Pentane

1.5 Relevancy of the Project

In this time of energy crisis where the production is far lesser than the demand, unless other power source is developed, optimization of the LNG plant remains the only remedy to address the issue. All the existing process plants are now operating beyond their capacities (Partho & Ruhul, 2011). In line with the objectives of this project, to increase the production of the LNG produced without any major investment consisting of building another production unit; this paper presents the way and analysis on optimization of the different structural arrangement of end flash unit. This is of paramount importance as the current technology developed has reached its own bottleneck and novel structural arrangement or invention is indeed needed.

The proposed solution comprises the production of LNG through cascade mixed refrigeration rather than mechanical cooling at a temperature of about -160. The end flash gas consists of mixture of hydrocarbon, ranging from methane, ethane, propane to butane, thus results in non-proportionality behavior of enthalpy of end flash gas with increasing temperature. It is critical to reduce the entropy generation through temperature difference between the end flash gas (hot stream) and mixed refrigerant (cold stream) in the cold box as the smaller the temperature difference between the hot stream and cold stream, the lower the input power needed for liquefaction purposes and thus leads to overall energy conservation in the end flash unit. Though pure refrigerant offers the benefits of simple cycle and ease of operation, large number of refrigeration stages is needed to provide the desired chilling effect. There is a trade-off between the complexity of the mixed refrigeration system and the number of stages required using pure refrigerant in order to achieve the liquefaction temperature.

1.6 **Feasibility of the Project Within the Scope and Time Frame**

Almost all the liquefaction plant is operate at its maximum capacities. As the market demand is far higher than the production of LNG in the recent years, the current liquefaction system offers degree of optimization through structural arrangement. This study enables the debottlenecking of LNG plants as a new process unit can be added as a means of significantly increasing the production from existing trains. Installation of end flash system offers expansion cooling which allow a higher throughput of warmer LNG stream to leave the liquefaction unit. The temperature of LNG leaving the liquefaction unit, the structural arrangement of the new process, and the molar composition of mixed refrigerant used are critical parameters that will govern the effectiveness of the said process.

This project is a continuous work which needs high commitment and good analytical skill in evaluating the advantages and disadvantages of each different arrangement of the end flash system. All the scope of study would have been thoroughly studied and evaluated in 8 months' time. The feasibility of the structural arrangement proposed will be justified by HYSYS simulator.

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CHAPTER 2

LITERATURE REVIEW

2. LITERATURE REVIEW

2.1 Liquefaction Process

According to Paradowski (2005), the invented refrigerating liquefied gas method is able to increase the delivery, and therefore its production by increasing the temperature at which the liquefied natural gas is produced. However, increment in the storage temperature would lead to an increase in its storage pressure which will incur more transportation and storage costs. The reason behind this invention is due to increasing market demand of LNG. This has been done through the use of end flash drum to separate the nitrogen-rich gas and liquefied natural gas. Liquefied natural gas will then be pump to tank for storage purpose while the nitrogen-rich gas will then be compressed and cooled before being expanded by a turbine to drive an electric generator.

Simulation is essential to ensure the reliability and profitability of the plant. Cameron *et al.* (2005) states that LNG is a commodity that was complicated to produce, process, ship and distribute. The paper has described the implementation of large-scale dynamic modeling for a large sub-sea production and gas liquefaction plant. This is done by decomposed the said model into the sub-models, distributing these models onto seven computers and running them in a common flow pressure network. Generally, LNG value chain consists of five steps: production, liquefaction, transport, re-gasification & distribution. For instance, the liquefaction process in LNG plant is indeed

very difficult to simulate as the process is tightly integrated, with many recycle streams. In addition, modeling of these processes requires highly accurate thermodynamic properties. However, through dynamic simulation, the risk of all stages of the LNG value chain is thus manageable. It is important to point out that all the simulation work done in this project is in steady state rather than in dynamic state as at this stage of simulation work, the process is assumes to be in smooth running process and is free of heat exchanger fouling, feed and environmental disturbances.

Houser *et al.* (2001) have studied on the way to improve efficiency of open-cycle cascaded refrigeration process and found out that this can be done through installation of liquid expander in the system. Liquid expander is being used to recover energy associated with the flashing of a pressurized liquid stream and employing said recovered energy to compress the flashed vapor streams in the open cycle. This method offers recovery of energy through expansion of pressurized natural gas.

In line with the effort of reducing the amount of refrigerant used for the liquefaction of natural gas, Olszewski (1972) has suggested a method and apparatus for the liquefaction of low boiling gases, such as the nitrogen and natural gas. Parallel refrigerant expansion engines and feed compression were used to reduce the specific heat of the feed to about 1.5-5 times the specific heat of the low pressure refrigerant gas. In result, the refrigerant required to liquefy the same capacity of natural gas feed has been reduced and thus improve the overall efficiency of liquefaction process.

Process efficiency of liquefaction plants has been greatly improved by between 3% and 5% due to the introduction of Cryoturbine^{TM} by Ebara International Corporation. The first machine has already been introduced at the Oman LNG Liquefaction Plant at Qalhat, Oman. (Liquid Expander in LNG Liquefaction Plant, n.d.). The strategy applied in the novel technology is replacing the pressure breakdown (Joule-Thompson) valve with liquid expander turbine. This approach enables to convert the hydraulic energy from cryogenic fluid into electrical power. The high efficiency of the

turbine compared to the zero efficiency of the valve provides a reduction in the temperature of the fluid as it flows through the machine.

In a recent paper done by Sayyadi & Babaelahi (2011), thermodynamic modeling has been performed based on energy and exergy analysis and an exergoeconomic model based on the total revenue requirement (TRR) has been developed in order to maximize the exergetic efficiency of plant and minimize the unit cost of the system product simultaneously. MATLAB has been used to find a set of Pareto optimal solutions. The paper also describes an example of decision-making process for selection of the final solution from the available optimal points of the Pareto frontier. The feature of selected final optimal system is compared with corresponding features of the base case and discussed. The reason for the simulation is in line with re-liquefy the LNG boiled off gas (BOG) back into cargo tank rather than being used up as fuel for the LNG carriers. The results show that the exergetic efficiency and the total product cost in the multi-objective optimum design was 11.11% and 16.7% higher than the exergetic efficiency of the exergoeconomic optimized system, respectively. The paper proved that by recovering the valuable gas, such as the BOG, it is able to improve the efficiency of the process.

In the paper "Optimal Deisgn and Operation of a C3MR Refrigerant System for Natural Gas Liquefaction" done by Wang *et al.* (2012), a new methodology for LNG liquefaction synthesis targeting energy consumption minimization is presented. It is based on thermodynamic analysis, mathematical programming, and rigorous simulation. This paper aims to minimize the energy consumption in LNG liquefaction process design and operation. The optimization results are then examined by Aspen HYSYS to ensure its solution feasibility. A typical case study for a C3MR process shows that a drastic drop of 13% in terms of energy usage in plant. This paper shows the method of simulating liquefaction of natural gas using Aspen HYSYS simulator.

Due to the steadily increasing price of energy source, more incentive have been taken to improve the efficiency and thus the complexity of the LNG plant. Plant

configuration and operation are the most critical factors that can significantly improve the efficiency of the plant. In the paper done by Mehrpooya *et al.* (2010), a novel process configuration for recovery of hydrocarbon liquid from natural gas is proposed, where the required refrigeration method is supplied by self-refrigerating system. The three most important characteristics of the proposed structural configuration are the high performance of the multi-stream heat exchangers, high recovery levels of the hydrocarbon liquids and low required compression power. The results show that the self-refrigeration compression power is 15.5% lower than the base case and the ethane recovery is 1.45% higher. In other words, this paper present an elegant solution by heating up or cooling down the stream using the recovered hydrocarbon stream, resulting in cost saving in utilities cost.

Foglietta (1998) has introduced a new LNG cycle that has developed for base load liquefaction facilities. Figure 2.1 shows the novel liquefaction process suggested. Such process scheme is economically more attractive to oil and gas companies as they are trying to find paths to monetize gas sources more effectively. For the purpose of simulation, a plant size of 75MMscfd is selected for the process design development. Simple energy index has been used to measure the performance of the said process. The parameters to be measured and compared are horsepower per unit of mass liquefied. It is interesting to note that such comparison is consistent with other paper presented on this subject. According to Foglietta (1998), the performance index for traditional process ranging from 0.87-1.64 while the turboexpander cycle is 1.0. By setting the turboexpander cycle as base case, performance index is able to justify whether the case discussed is effective or otherwise. Though turboexpander cycle hold much promise to energy saving in the plant, it is important to note that further optimization is needed as the traditional process is still offers much better performance at performance index of 1.64.

This is indeed supported by the Owen *et al.*(2009) and Eaton *et al.* (2008), who describe the use of liquid turboexpander to expand the feed while at the same time drive the compressor of the device and therefore provide compression for a close loop

propane refrigeration cycle to pre-cool the natural gas stream. As a result, optimization is done through increasing the volume of LNG production for a given amount of installed horsepower or alternatively, to reduce the capital cost and operating cost associated with the production of specific amount of LNG. The reason for the invention is due to the uneconomical situation to develop natural gas reserves as the cost of processing and transporting the gas to distant market are extremely high.
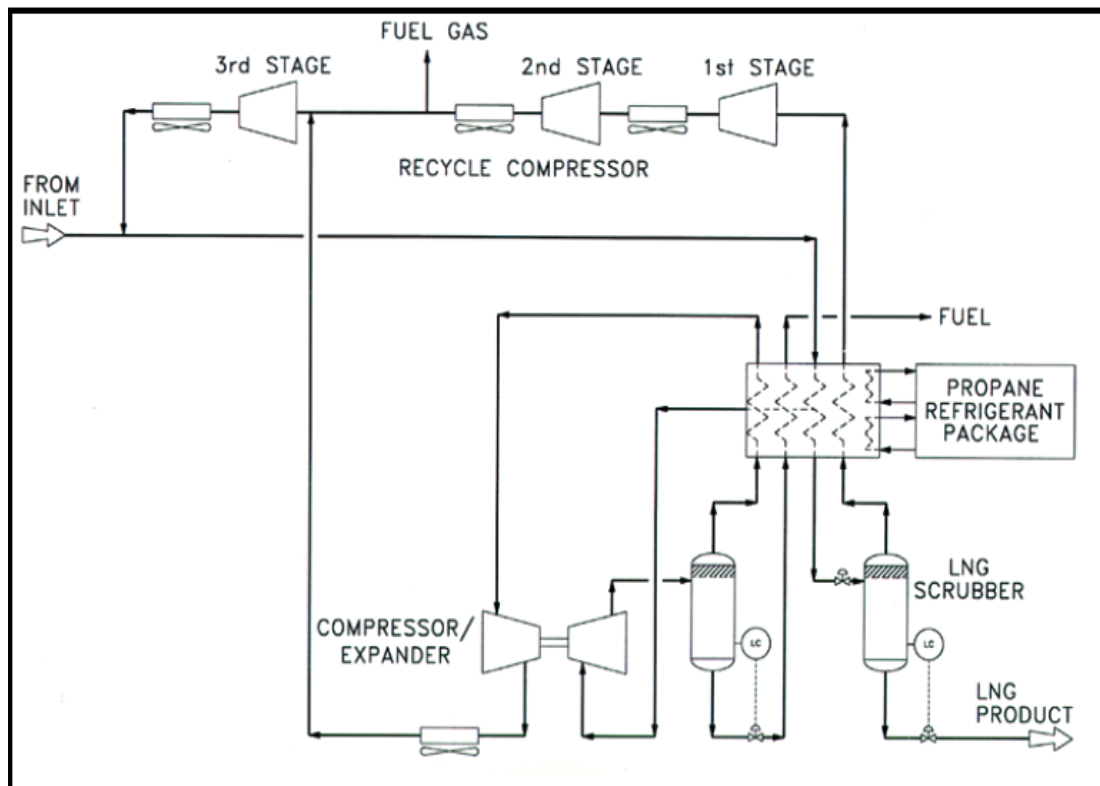
Figure 2.1: Novel Liquefaction Process Proposed by Foglietta

Referring to the paper done by Spilsbury *et al.* (2007), environmentally friendly low nitrogen oxide (NOX) burners for the turbines have a lower tolerance for nitrogen in fuel gas than previously used burners. Thus it is necessary to remove nitrogen from the feed stream and this is done by feeding the LNG through a two stage separation in which it is first fractionated to provide nitrogen enriched overhead vapor streams and a bottom liquid streams. The bottom liquid stream is then fractionated again to provide second nitrogen-enriched overhead vapor stream and a stream of purified LNG as

bottom. Through the method suggested, it is able to generate nitrogen-depleted fuel gas that is able to drive the compressor while produce purified LNG.

Paradowski (2002) studied the method of removal of nitrogen from the nitrogen-rich fuel gas. Figure 2.2 shows the block diagram for the nitrogen removal unit from the vapor of end flash drum. Due to stricter environmental regulations that forbid flaring of end flash gas or associated gas to environment and steadily increasing pressure to exploit on high nitrogen gas resources, there is more incentives and thus more papers, such as Hann (2003) and Butwell *etl al.* (2001), are published on the removal and recovery of nitrogen gas from LNG. The invention is able to effectively separate the nitrogen from the natural gas and at the same time produces nitrogen that is essentially free of hydrocarbon.
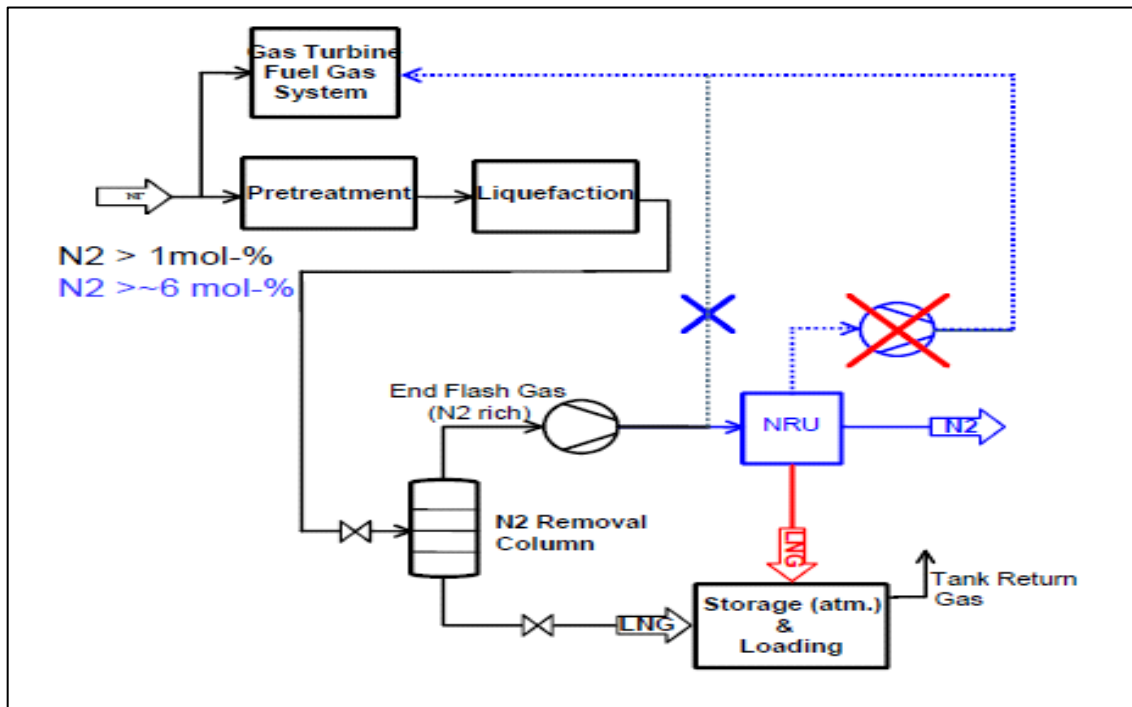
Figure 2.2: Block Diagram of Nitrogen Removal Unit

According to Finn *et al.* (1999), machinery constitutes a major portion of total capital cost. Thus improving thermodynamic efficiency reduces power requirements, machinery size and, hence, capital cost. In order to resolve the issues addressed, exergy analysis, which is a fundamental design tool to reduce costs has been employed (Finn *et*

al, 1990). It is believed that exergy analysis is able to evaluate the process irreversibility and inefficiency. Minimum work to produce a given amount of LNG for a given feed pressure and composition by assuming reversible process, also termed as Carnot work, has been calculated. However, in real processes, the actual work is always greater than the Carnot work since real processes suffers from friction in the compressors, finite temperature differences in the heat exchangers, irreversible flashed across throttling valves, and heat loss to the surroundings (Wang & Economides., 2009). Despite the fact that Carnot work is unable to represent the actual work in a real process, Finn *et al.* (1999), state that Carnot work can be compared with the actual work required by a given process flowsheet or part of a flowsheet, to identify potential process improvements. This is further justified by Paradowski (2005), which states that the Carnot work accounts for 51.5% of the actual work required regardless of the variation in feed in composition and temperatures.

In the paper 'Process to Obtain Liquefied Natural Gas', Migliore *et al.* (2012) propose a novel process to obtain LNG which comprises the use of air as refrigerant in an open or closed cycle. This novel invention is particularly advantageous when located in barges for liquefying gas from small natural gas fields located in distant area, far away from the coast. The process flow as suggested by Migliore *et al.* is shown in Figure 2.3.

Figure 2.3: Conceptual Design of Liquefied Natural Gas Using Air as Refrigerant

2.2 Structural Decomposition Analysis

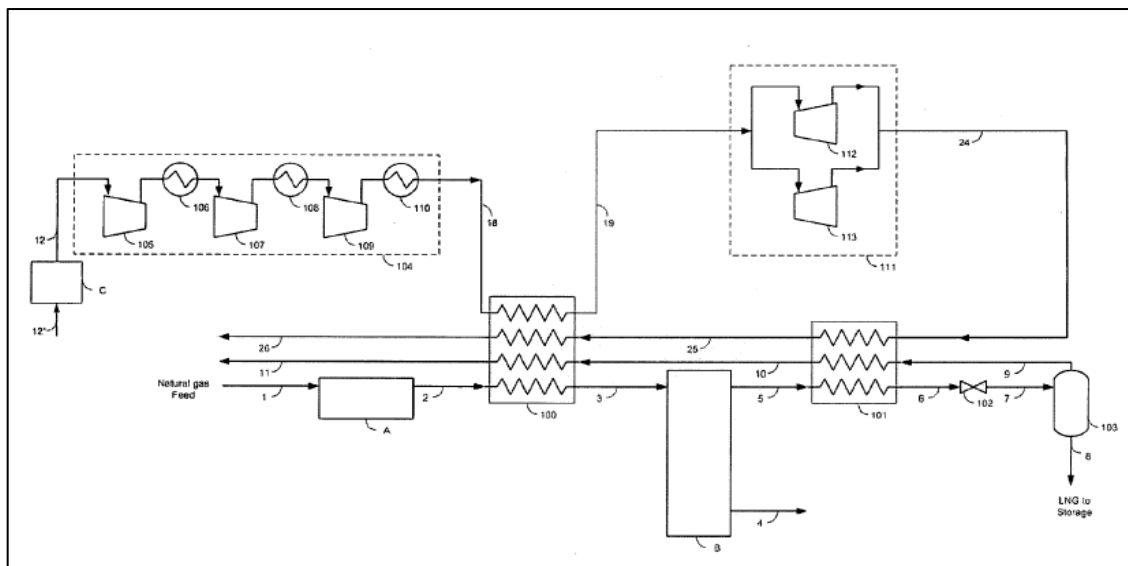This project will study the possible structural optimization of end flash system through HYSYS simulation software. Structural decomposition method has been used heavily in simulation work as the convergence of the whole process is difficult. By structural decomposition, the process flow is broken down into few parts for the ease of simulation.

According to Hubacek *et al.* (2006), structural decomposition has been widely used to explain the changes that occur in any variable over time or space. For instances, structural decomposition has applied in the paper to explain the percentage increment of CO_2 emission from China from the year 1992 to 2002. Hubacek *et al.* (2006) has decomposed the emission of CO_2 to several factor and analyzed it individually. Consumption volume has the highest increment of all which stands at 129% while the consumption structure has increased by 3% over the 10 years gap. However, production structure and emissions efficiency record a negative growth by registering a value of -11% and -62% respectively. Thus, Hubacek *et al.* (2006) concluded that the overall emission of carbon dioxide of China has increased by 59% throughout the time of investigation.

Referring to Milana (2001), the input-output structural decomposition analysis (SDA) is traditionally used to study the observed changes in the level of mix of output and employment. This method is certainly not new as Dietzenbacher and Los (2000), have employed the said method to study the value added growth of Netherlands from the year 1972 to 1986. Structural decomposition techniques are used to break down the changes in one variable into the changes in its determinants.

Nowadays, structural decomposition method has been extended to the simulation of LNG system. In the paper "Simulation & Data Validation of Small-Scale LNG system", Sapuan (2008) has claimed that structural decomposition approach has been chosen as it is difficult to converge LNG exchanger units without enough or complete process data. Thus, structural decomposition has applied on the liquefaction process of mixed-refrigerant cycle (MRC) in order to simplify the simulation on the unit.

CHAPTER 3

RESEARCH METHODOLOGY

3. GENREAL METHODOLOGY

3.1 Base Case Analysis and Design

The process flow as suggested by Paradowski (2005) is taken to be the base case flow sheet for process simulation. For simplicity, the term base case and improved end flash system would be used interchangeably throughout the context of this report. For all the cases studied, the LNG from the outlet of main cryogenic heat exchanger (MCHE) will be assumed to have a mass flow rate of 556,506kg/h. The temperature and pressure for the LNG stream is -147.0℃ and 48.0 bar. The molar composition of the LNG stream is shown in Table 3.1:

Table 3.1: Molar Composition of LNG stream

Species	Molar Composition (%)
Nitrogen	0.10
Methane	91.40
Ethane	4.50
Propane	2.50
n-Butane	0.60
i-Butane	0.90

Referring to Figure 3.1, for the improved end flash system, the LNG **1** from the MCHE will first undergoes an expansion to 1.15 bar before being fed into end flash drum. LNG **6** leaves the bottom of the end flash drum at -159.1℃ and will be pumped to 1.5 bar before sent to storage. The distillate of the end flash drum **5**, which is more commonly known as end flash gas, will be fed into cold box **E-101**. For the ease of operation, end flash gas first leaves the cold box, **E-101** at the temperature of 32℃. The system makes use of the low temperature of the end flash gas to cool down the stream **16**, **25** and **28** respectively. The stream at outlet of the cold box **8** will undergo a plurality of compression stages **9, 11, 14, 16** and series of heat exchange, preferably with water coolers **E-102**, **E-103** and **E-104**. The function of intercooler, which is situated at the outlet of each stage of the compressors, is used to lower the suction temperature of the next stage of compressor in order to reduce the compressor duty. Stream **16** will be fed into **E-101** again to lower down the temperature to approximately 33.0℃. At the end of compression and cooling stages, a small part of the stream **17** will be tapped off as stream **18** which will be treated as the fuel gas. The quantity of fuel gas required is equivalent to the machinery power of all the compressors in the end flash system. A major portion of stream **17** which corresponds to a stream **19** is tapped off. This stream **19** is first compressed to a pressure of 41.5 bar and then passes through water cooler **E-100**, to yield a cooled stream **24** at a temperature of 37.0℃. Stream **24** will be further divided into two different streams. Stream **26**, which is made up of 21% of stream **24**, will be fed into **E-101** and cooled down to -141.0℃ to give stream **31**. Stream **31** will be recycled back to the system by in line mixing with stream **3** to yield stream **4**, which is then introduced into the end flash drum. Stream **25** which consists 79% of stream **24** will first pass through **E-101** to cool down to -60℃, expanded to approximately 9 bar and then giving its heat to stream **16**, **25** and **26** by passing through **E-101**. The outlet stream **29** will then mix with stream **11** before introduced to the medium pressure third stage of compressor, **K-103**. By feeding stream **29** straight to the third stage of compressor instead of the low pressure first stage of compressor, **K-101**, the compressor duty is greatly reduced in **K-101** and **K-102** as the flow rate passing through both stages of the compressors is low.

Stream Name Temperature,	1 -147.00	4 -159.10	6 -159.10	8 32.00	12 55.51	13 37.00	14 119.61	15 37.00	16 120.08	17 32.95	19 32.95	23 324.50	25 37.00	26 37.00	27 -60.00	28 -127.20
Pressure, kPa	4,800	115	115	115	552	552	1,265	1,265	2,900	2,900	2,900	41,500	41,500	41,500	41,500	900
Mass Flow Rate, kg/h	556,506	597,297	539,533	57,763	211,766	211,766	211,766	211,766	211,766	211,766	194,952	194,952	154,012	40,940	154,012	154,012

Figure 3.1: Developed Base Case

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3.2 Required Parameters

In order to analyze the and thus improve on the developed base case, few critical parameters are being defined and are closely monitor throughout the project as the said parameters will govern the profitability and technical feasibility of the novel invention. One of the most important parameter is the specific power of LNG production (k/kg). Increment of production normally will results in substantial increment in terms of the cost. In the context of end flash system for LNG production, the increment in LNG production from end flash unit normally will increase the compressor duty or liquefaction power needed and subsequently such increment will reflect in terms of plant utilities cost. Specific power of LNG production describes the power or energy required to produce 1 kg of sales LNG. The ratio of power consumption of compressors to the production of LNG reflects the effectiveness of the end flash system. For an optimization case to be considered as technically and economically feasible, the said optimization case must show a lower ratio of power consumption of compressor to the production of LNG. This can be achieved either through reduction of compressor duty through structural arrangement of the given flow sheet as suggested by Paradowski (2005) or by substantially compensating the compressor duty by liquefaction power as will be covered in further detail in Chapter 3.3. The current technology, which is shown in Figure 3.1, is matured and thus limited optimization opportunity can be done on the said system. Thus, it is inevitable that a novel structural arrangement or invention is indeed needed.

The second parameter that is closely monitored is the production of sales LNG (kg/h). Basis of the simulation case for feed inlet is 556,506 kg/h which is equivalent to 4 MTPA of LNG. Given the same feed mass flow rate, the optimization case must be able to equal or produce more LNG through the end flash drum as the difference in mass flow rate of LNG produced is the direct indication of the profitability of the optimization case over the base case. The amount of LNG rundown is closely related to the ratio of end flash gas recycled back to the end flash drum as the unrecovered end flash gas will be treated as fuel gas to drive the compressors of the system.

The third critical parameter is the machinery power of the unit (kW). The higher the machinery power, the higher the amount of fuel gas needed to drive the compressors, and thus the amount of end flash gas recycled back to the system would be decrease. The compressors duty can be significantly reduced by cooling down the compressor suction side temperature. It is normal practice to have intercooler at the outlet of multi-stage compressors as shown in Figure 3.1. The function of intercooler, preferably water cooler, is to lower down the suction side temperature of higher stage of compressor and thus reduce the compressor duty needed.

For the simulation purpose, the operating conditions are identical for both the base cases and the optimization case. Fluid packages chosen for all the simulation cases are Peng-Robinson. Theses natural gases are deliberately free of pentane and higher hydrocarbons, so as to simplify the simulation on the liquefaction of C5+ components (Paradowski, 2005). The feed gases are furthermore assumed to have been pre-treated for the removal of acid gases and other trace components.

3.3 Structural Modification

Two optimization cases, which are able to perform the similar function as the improved end flash system, are proposed. In Figure 3.2, for the first optimization case, namely the pure refrigerant cascade, the LNG **1** from MCHE at -147℃ and 48.0 bar is first expanded and cooled in an liquid hydraulic expander to recover some of the energy loss before it is expanded and cooled again in a Joule-Thompson (JT) Valve to achieve a temperature of -159.0℃. The expanded stream **3** will mix with recovered end flash gas stream **13** to yield stream **4** before being fed into the end flash drum. The bottom product which is the LNG stream will then be pumped and sent to storage. The distillate of the end flash drum, which is also known as end flash gas, is then introduced into compressor **K-101** to increase its pressure to 3.0 bar before passes through cold box **E-101** to heat exchange with subcooled pure methane refrigerant. The outlet stream **9** at temperature of -151.4℃ and pressure of 2.5 bar is then split into two different stream where 87.4% of the stream will be used as fuel gas for the unit. Stream **10** which consists of 12.6% of stream **9** will be recycled back to the end flash drum as stream **13**.

Pure methane has been used to heat exchange and liquefy the end flash gas. Stream **14**, which carries approximately 201 t/h of subcooled pure methane is introduced into cold box **E-101**, absorbs the heat from stream **8** and vaporizes to give stream **15**. The vaporized stream will be compressed by compressor **K-102** to 40 bar and then cooled down to -87.8℃ in **E-102** to give stream **17**. Stream **17** will then pass through JT valve to give subcooled stream **14**. For the cooling down of methane refrigerant in **E-102**, pure ethane has been used as the refrigerant. Approximately 397 t/h of ethane in subcooled stream 18 is first heat exchange with methane refrigerant in **E-102** and then compressed to 19.5 bar. The compressed stream is then fed into **E-103** to cool down to -31.0℃ and let down by another JT valve to 1.53 bar. Similarly, 678 t/h of propane refrigerant is being used to cool down the ethane refrigerant in **E-103**. The vaporized stream **23** is first compressed to 7.7 bar before cooled down to 12℃ by a chilled water cooler to give stream **25**. Stream **25** will undergoes let down by JT valve to give stream **22** at 1.45 bar. The inlet temperature for the chilled water is 5℃ while the outlet temperature is 12℃.
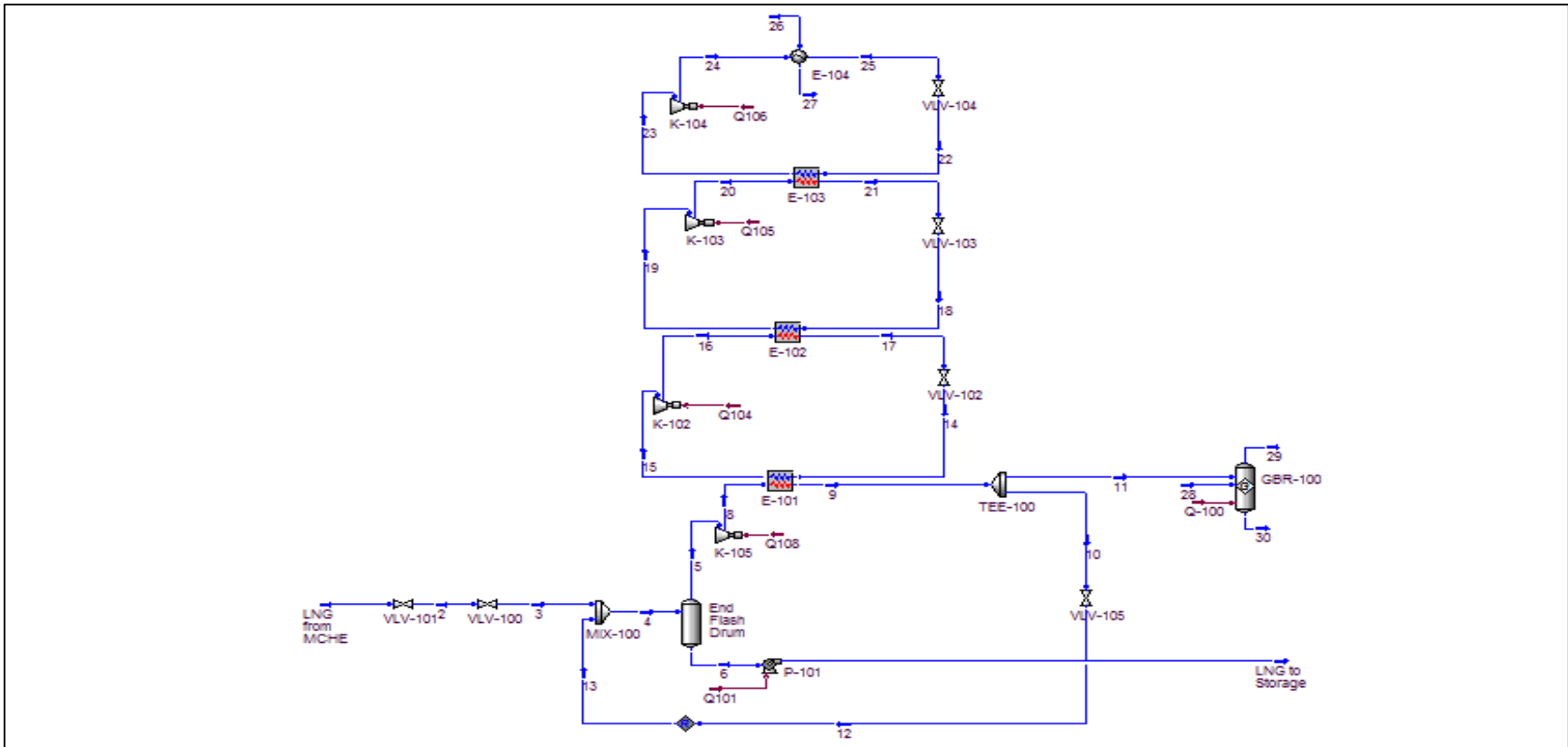
Stream Name	1	4	7	13	14	15	16	17	18	19	20	21	22	23	24	25
Temperature,	-147.0	-159.0	-159.0	-161.0	-157.5	-159.5	61.0	-87.8	-96.6	-96.0	80.0	-31.0	-33.64	-30.0	50.0	12.0
Pressure, kPa	4,800	115	150	115	140	120	4,000	3,950	153	133	1,950	1,900	145	125	770	720
Mass Flow Rate, kg/h	556,506	562,515	514,367	6,008	201,081	201,081	201,081	201,081	396,685	396,685	396,685	396,685	678,079	678,079	678,079	678,079

Figure 3.2: Developed Optimization Case of Pure Refrigerant Cascade

For the second optimization case, namely the mixed refrigerant cascade, the LNG stream at -147℃ and 48.0 bar is expanded through a hydraulic expander and then through JT valve to a pressure of 1.15 bar. The expanded stream **4** is then fed into end flash drum to produce a top distillate stream **5** and bottom product of stream **6**. Stream **6** will be pumped to 1.5 bar and sent to storage. The top distillate of end flash drum will then be compressed to 3.0 bar and then heat exchange with first closed loop of mixed refrigerant. The composition of the first closed loop of mixed refrigerant is shown in Table 3.2:

Table 3.2: Molar Composition of Refrigerant in E-101

	Molar Composition (%)
Methane	92.00
Ethane	8.00

Stream **9** at temperature of -149.2℃ is then heat exchange in **E-102** with second closed loop of mixed refrigerant. The molar composition of the second closed loop of mixed refrigerant is shown in Table 3.3:

Table 3.3: Molar Composition of Refrigerant in E-102

	Molar Composition (%)
Nitrogen	22.00
Methane	78.00

Use of different composition of refrigerants in two different loops is aim to reduce the quantity of refrigerant needed to liquefy the end flash gas and thus reduce the mechanical load on the compressors. Further justification and explanation can be reviewed in Chapter 4. Stream **10** at the outlet of **E-102** is then split into stream **11** and stream **12**. Stream **11**, which is made up of 25.6% of stream **10**, will be treated as fuel gas to compensate for the machinery power required in the unit. The remaining 74.4% in stream **12** will then pass through a JT valve and then recycle back as stream **14**.

Stream **15**, which has 191 t/h of mixed refrigerant with the composition as shown in Table 3.2, is used to precool the end flash gas to -149.2℃. At his temperature, the end flash gas still exist in vapor state; further cooling will result in phase change and thus the hot composite curve will exhibits a vastly different behavior which could not be assimilated by the cold composite curve beyond the temperature. The resolution to the issue aforementioned is by introducing a second closed loop of different composition of mixed refrigerant which exhibits almost the same cooling curve as the hot composite curve. The heated stream **16** at temperature of -124.1℃ and pressure of 1.1 bar passes through compressor **K-102** to yield stream **17** at 40.0 bar. The compressed stream is then first heat exchange with water cooler **E-106**, cold box **E-103** and lastly cold box **E-104** to achieve a subcooled stream **20** at temperature of -81.3℃. Stream **20** will then passes through JT valve to undergo further subcooling to -152.5℃.

For the second closed loop mixed refrigerant, stream **21**, which has the composition as shown in Table 3.4, is first passes through cold box **E-102** and compressor **K-103** to yield stream **23** at temperature of 3.0 and 35.0 bar. Stream **23** will undergo a series of heat exchange through **E-103**, **E-104**, and **E-105** to yield stream **24**, **25** and **26** at -40.0℃, -88.0℃ and -111.9℃, respectively. Expansion cooling through JT valve is done on stream **26** to yield stream **21**, which has a temperature and pressure of -154.9℃ and 2.8 bar, respectively.

The following three different closed loops have no direct heat exchange with the end flash gas; however their presence is critical to complete the cooling cycle in the system. For instances, stream **27** at -57 is responsible to partially cool down the stream **18**, **23** and **36**. The outlet stream **28** from cold box **E-103** at temperature of -3.7℃ is then compressed to 40.0 bar using compressor **K-104**. The compressed stream is then heat exchange with chilled water cooler to 12.0℃ and then passes through JT valve to yield stream **27** at 1.7 bar. The molar composition of the mixed refrigerant in the closed loop is shown in Table 3.4:

Table 3.4: Molar Composition of Refrigerant in E103

	Molar Composition (%)
Ethane	60.00
Propane	27.50
n-Pentane	12.50

Similarly, stream **31** is used to cool down the stream **19** and **24**. The inlet temperature for stream **31** is -91.8℃ and after it passes through cold box **E-104**, the outlet temperature of stream **32** is -49.8℃. Compressor **K-105** is used to increase the pressure of the stream **32** to 70.0 bar. The compressed stream is then heat exchange with chilled water cooler to achieve an outlet temperature of 12.0℃ and let down by JT valve to a pressure of 1.25 bar. The molar composition of refrigerant used in this closed loop is shown in Table 3.5:

Table 3.5: Molar Composition of Refrigerant in E104

	Molar Composition (%)
Ethylene	83.00
Propane	9.00
n-Butane	8.00

In cold box **E-105**, mixed refrigerant stream **38** at -113.4℃ and 2 bar is used to subcool stream **25** from -88.0℃ to -111.9℃. The outlet stream **39** at -104.4℃ is then introduced to compressor **K-106** to raise up its pressure to 40 bar. The compressed stream is then heat exchange with cooling water stream **46**. The outlet stream **36** assumes the same temperature as the cooling water outlet temperature, which is at 37.0℃. Stream **36** is then fed into cold box **E-103** to further cool down to -51.8℃ before passes through JT valve to undergo expansion cooling to -113.4℃ to yield stream **38**. The molar composition for the mixed refrigerant in stream **38** is shown in Table 3.6. All the inlet temperature of cooling water is 30℃ and the outlet temperature is assumed to be 37℃. For instances, stream **40**, **42**, **44**, and **46** are cooling water from utilities site

which are being used as cooling agent to cool down stream **17**, **29**, **33**, and **36** to 37℃. The machinery power required in this system is the summation of power required to drive the compressor **K-101**, **K-102**, **K-103**, **K-104**, **K-105**, and **K-106**. A point to note is that though the equipment costs of the unit will be affected by the quantity and capacity of the compressors, the number of compressor is relatively insignificant to the operating cost of the unit, particularly in the utilities cost. Detailed economic analysis on the installation of mixed refrigerant cascade over the improved end flash system will be studied thoroughly in Chapter 4.

Table 3.6: Molar Composition of Refrigerant in E105

	Molar Composition (%)
Methane	55.00
Ethylene	45.00
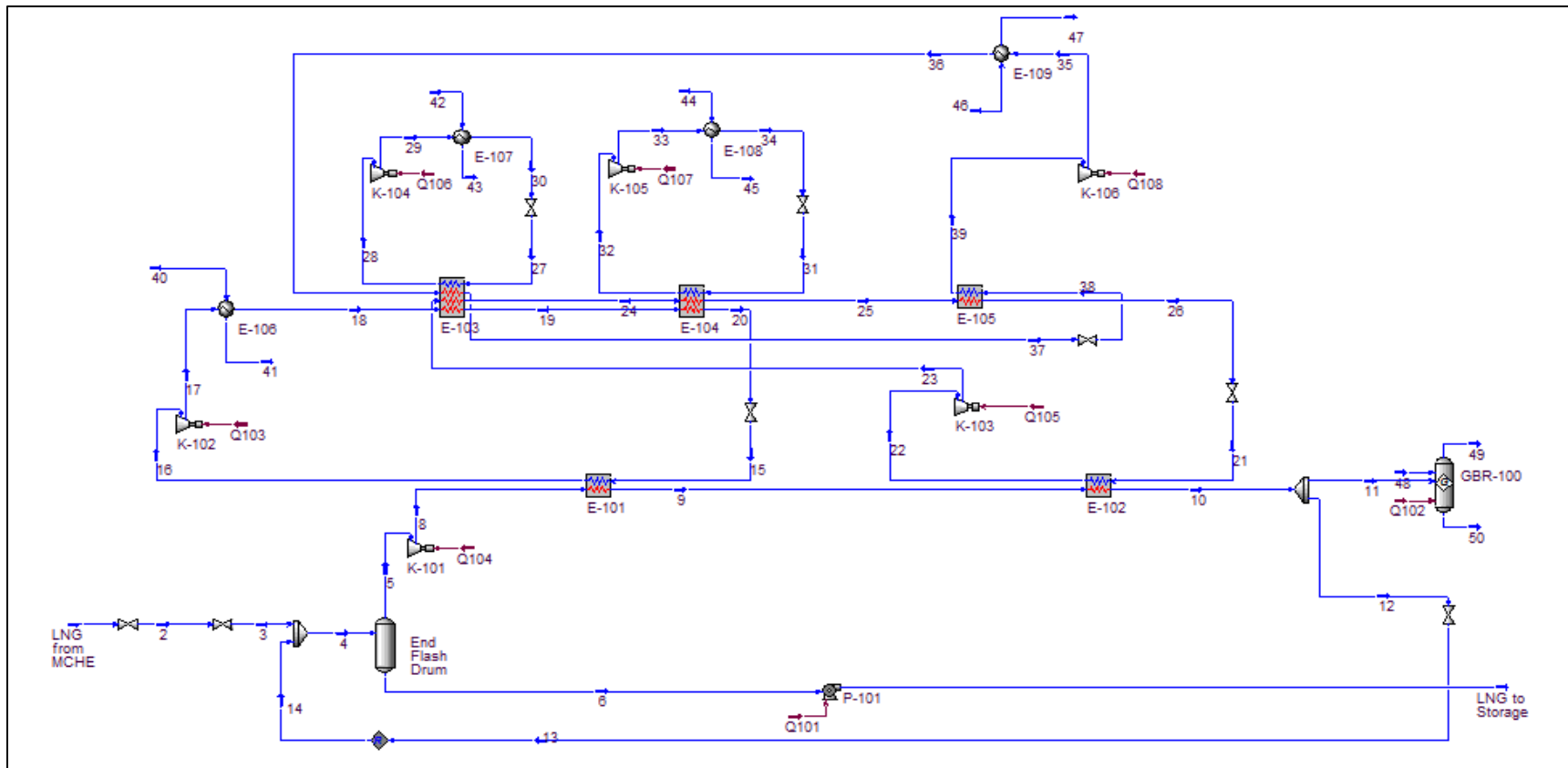
Stream Name	1	4	7	8	10	15	16	17	18	19	21	22	23	24	25	26
Temperature,	-147.0	-159.2	-159.2	-117.5	-151.0	-152.5	-124.1	134.8	37.0	-81.3	-154.9	-152.5	3.0	-40.0	-88.0	-111.9
Pressure, kPa	4,800	115	150	300	250	130	110	4,000	3,950	3,900	280	260	3500	3450	3400	3350
Mass Flow Rate, kg/h	556,506	597,152	542,437	54,715	54,715	19,133	19,133	19,133	19,133	19,133	117,603	117,603	117,603	117,603	117,603	117,603

Figure 3.3: Developed Optimization Case of Mixed Refrigerant Cascade

3.4 General Methodology

For the purpose of this project, the process flow as suggested by Paradowski (2005) is taken to be the base study of simulation. Required parameters to be optimized are identified and are listed before the simulation process has been conducted. Structural modifications on the said process as well as modification on the operating parameters are being done in order to achieve the objectives of this project. Feasibility check serves as the final judgment on the said modified structural configuration is technically feasible and economically affordable compared to the base design. The optimization case is said to be technically feasible when the specific power of LNG production, P_o is lower than the specific power of LNG production of base case, P_B. Failure to achieve the required parameters will lead to rejection of the simulation work. Only the simulation works that satisfy the parameters will proceed to the next stage of feasibility check, namely the economic justification. For an engineering project to be considered successful, it must be technically feasible and economically wise to operate. The engineering project is said to be economically wise to implement when its interest rate of return (IRR) is higher than the minimum acceptable rate of return (MARR). Failure to satisfy either one of the feasibility check will lead to rejection of simulation work. Besides that, this simulation project is considered as sequential work as the work that satisfy the first feasibility check only, namely the specific power of LNG production, would be considered for the economic justification. Upon the completion of this project, a new design will be proposed at the later part of this project. Figure 3.4 shows the summary of methodology of this project.
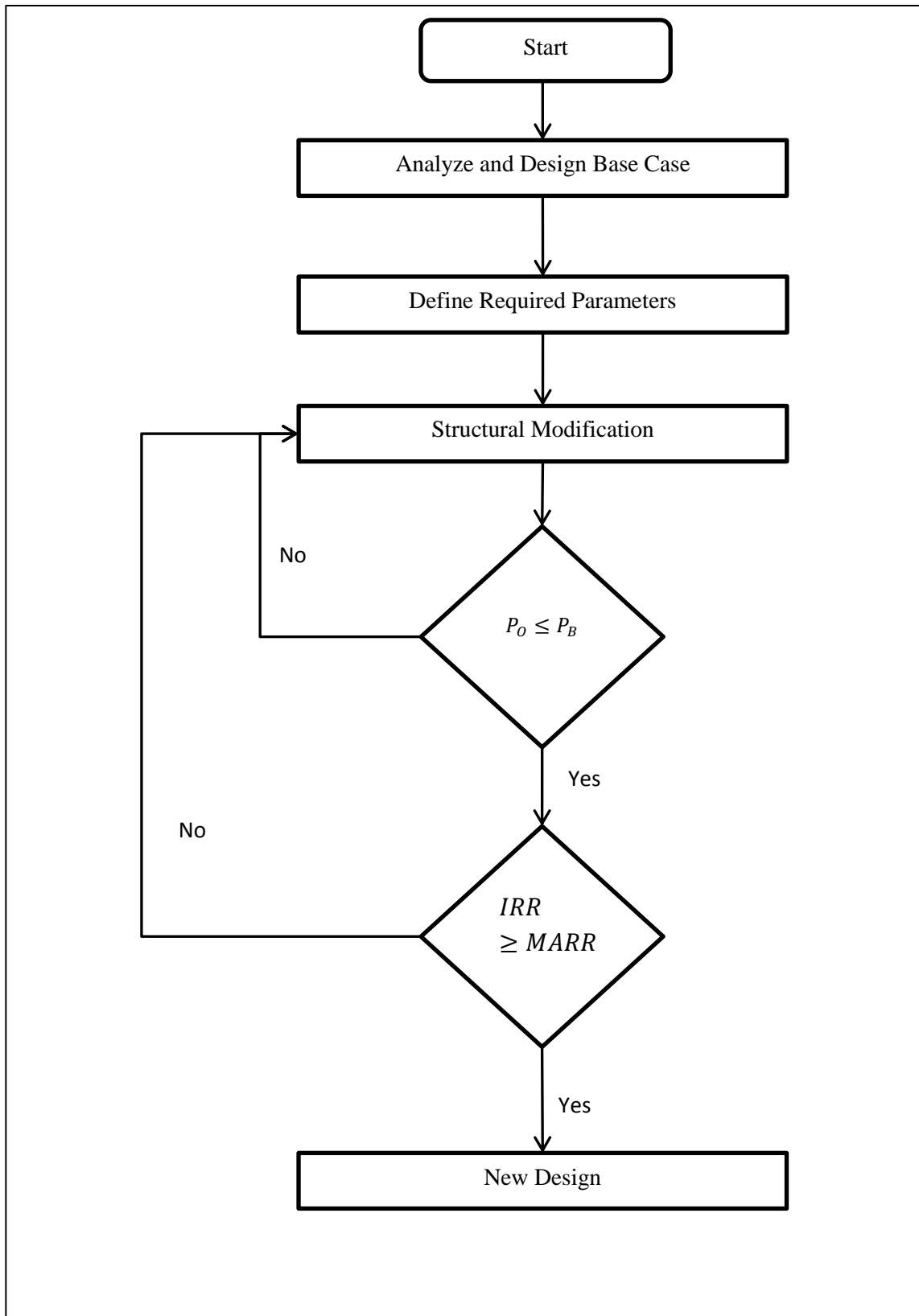
Figure 3.4: Summary of general methodology

3.5 Key Milestones

Key milestones play the role as stepping stones on the critical path to the completion of any project. The key milestones identified are indeed measurable and achievable. Several key milestones have been outlined prior to the start of this project for FYP II in order to ensure the completion of the said project within the given time limit:

Table 3.7: Key milestone of FYP II

Key Milestones	Expected Time of Completion
Submission of Progress Report	7th Week
Pre-SEDEX	10th Week
Submission Draft Report	11th Week
Submission of Dissertation	12th Week
Submission of Technical Paper	12th Week
Oral Presentation	13th Week
Submission of Project Dissertation (Hard Bound)	14th Week

3.6 Gantt Chart for FYP II

No.	Activity/Week	1	2	3	4	5	6	7	8	9	10	11	12	13	14
1.	Optimization Case I	■	■	■	■										
2.	Analysis on Optimization Case I			■	■	■									
3.	Optimization Case II					■	■	■	■						
5.	Analysis on Optimization Case II							■	■	■					
6.	Submission of Progress Report							A							
7.	Pre-SEDEX										A				
8.	Submission of Draft Report											A			
9.	Submission of Dissertation (soft bound)												A		
10.	Submission of Technical Paper												A		
11.	Oral Presentation													A	
12.	Submission of Project Dissertation (Hard Bound)														A

■	Process		A	Key Milestone

Figure 3.5: Gantt Chart

CHAPTER 4

RESULT AND DISCUSSION

4. RESULTS AND DISCUSSIONS

4.1 Technical Feasibility Study

As aforementioned, the natural gas is mainly composed of methane and small amount of ethane, propane and butane, therefore the phase change is expected to occur at non-constant temperature. According to Smith (2005), it is preferred to have a refrigerant cooling curve that is similar to the natural gas cooling curve as such effort will reduce the energy consumption. If pure refrigerant is being used as the refrigerant, more cycles are needed and thus the overall process is more efficient, however, at the expenses of higher operating costs. Another option is to use a mixed refrigerant which exhibits a cooling curve that is as near as possible to the natural gas (hot stream) cooling curve. The vertical spacing between the hot and cold streams is a measure of the temperature driving force for heat transfer. A minimum approach temperature of 3℃ is being chosen for the system. If the cold composite stream is shifted in such a way that it overlaps or has temperature difference less than the minimum approach temperature with the hot composite stream, then there is no integrated heat exchange between the two streams. However, if the temperature difference is much higher than the minimum temperature difference allowable, the heat exchange is deemed to be inefficient as more refrigerant is needed to achieve the desired heat exchange, resulting in large size and large amount of equipment needed. All the processes depicted in this paper are designed to closely

approach the cooling curve of the end flash gas. This is done by monitoring the cooling curve at the different stages of liquefaction process to achieve high refrigeration efficiency and reduce energy consumption.

The improved end flash system as suggested by Paradowski (2005) has been taken to be the base case of this project. Two optimization cases, namely Pure Refrigerant Cascade and Mixed Refrigerant Cascade, have been suggested and will be studied thoroughly in terms of production and energy consumption as shown in Table 4.1. For the production of 4 MTPA LNG train, LNG rundown from the improved end flash system is 540t/h while the pure refrigerant cascade and mixed refrigerant cascade are 520t/h and 542t/h, respectively. However, the specific lower heating value (LHV) for the base case and both the optimization cases show little difference as the improved end flash system and both the optimization cases record a value of 49.4 MJ/kg. The significant difference in LNG rundown results in difference in total lower heating value (LHV) of the sales LNG. The total LHV of the improved end flash system is 7.40 GW while the pure refrigerant cascade records a value of 7.12 GW as this is due to the significant reduction of LNG rundown as compared to the base case. The mixed refrigerant cascade shows an increase of 0.41% as both the LNG rundown and specific LHV are higher than the improved end flash system.

All of the cases discussed above are designed in such a way that the unit itself is operating independently, in which a portion of the end flash gas will be used as fuel gas for gas turbine to drive the compressors in the end flash system. The remaining end flash gas will be recycled back to the end flash drum in order to improve LNG recovery. In other words, the compressor duty will have a significant impact on the quantity of the end flash gas recovered and the performance of the optimization cases, ultimately. Referring to Table 4.1, the quantity of fuel gas needed to power the compressors is 17t/h while the pure refrigerant cascade and mixed refrigerant cascade need 37t/h and 14t/h, respectively. The relatively high amount of total LHV of fuel gas in the pure refrigerant cascade reflects the loss of LNG recovered as more energy has been used to drive the compressors. As compared to the base case, the pure refrigerant cascade utilizes

0.28GW more while the mixed refrigerant cascade consumes only 0.19GW for the same purpose.

The machinery power in this context is referring to the power required to drive all the compressors in the system. The number of compressors used in the improved end flashes system, pure refrigerant cascade and mixed refrigerant cascade are 5, 4 and 6, respectively. Although the mixed refrigerant cascade uses the most compressors in the unit, the machinery power is the lowest among all the cases discussed owing to the fact that the use of mixed refrigerant successfully closing the gap of the cold composite curve and the end flash gas cooling curve. The merit of this approach allows the quantity of refrigerant required to be significantly reduced and thus the compressor duty will be significantly lower as well. The machinery power required in the mixed refrigerant cascade is 41.2MW which is 21.72% lower than the machinery power required in the base case. Although the pure refrigerant cascade utilizes the least compressor which indicates the lowest capital investments, the machinery power required is recorded at 135.2MW, which is the highest among all the cases discussed.

The specific power of production of LNG reflects the effectiveness of the system by describing the amount of energy needed for the production of one kilogram of LNG. The system is said to be relatively effective if the specific power of production of LNG is lower than the base case. The specific power of production of LNG for the base case is 351.38kJ/kg while the mixed refrigerant cascade is 273.74kJ/kg. The reduction of approximately 22% in terms of specific power of production for the mixed refrigerant cascade is due to the increment in LNG rundown and reduction in machinery power required as compared to the base case. For the pure refrigerant cascade, large portion of the end flash gas is used as fuel gas to drive the compressor, thus there is a significant reduction of 20t/h of LNG rundown and consequently the specific power of production of LNG is therefore, highest among all the cases of interest.

Table 4.1: Comparison of Simulation Results for Base Case and Optimization Case

Rundown	Improved End Flash (Base Case)	Pure Refrigerant Cascade	Mixed Refrigerant Cascade
Flow rate, t/h	540	520	542
Specific LHV, MJ/kg	49.4	49.4	49.4
Nitrogen Content, mole %	0.054	0.033	0.056
Total LHV, GW	7.40	7.12	7.43
Total LHV, %	100.00	96.21	100.41

Fuel Gas

Flow rate, t/h	17	37	14
Specific LHV, MJ/kg	48.7	49.2	48.7
Total LHV, GW	0.23	0.51	0.19

Machinery Power

Fuel gas compressor, MW	52.7	135.2	41.2

Performance

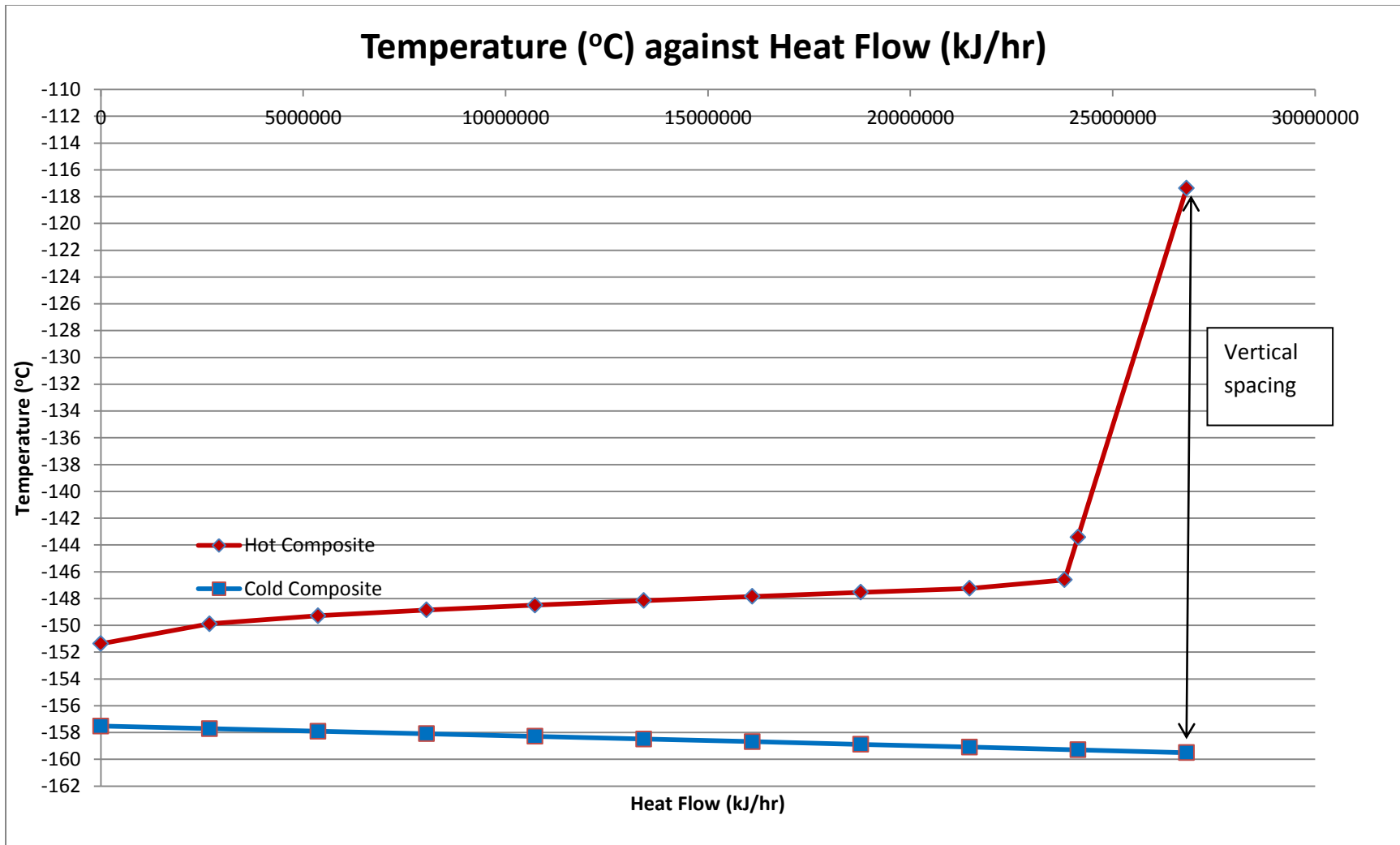|			
Additional production of LNG, kg/h	-	(20,151)	2,644
Specific power of production of LNG, kJ/kg	351.38	977.18	273.74

Figure 4.1: Temperature against Heat Flow for Pure Refrigerant Cascade in E-101
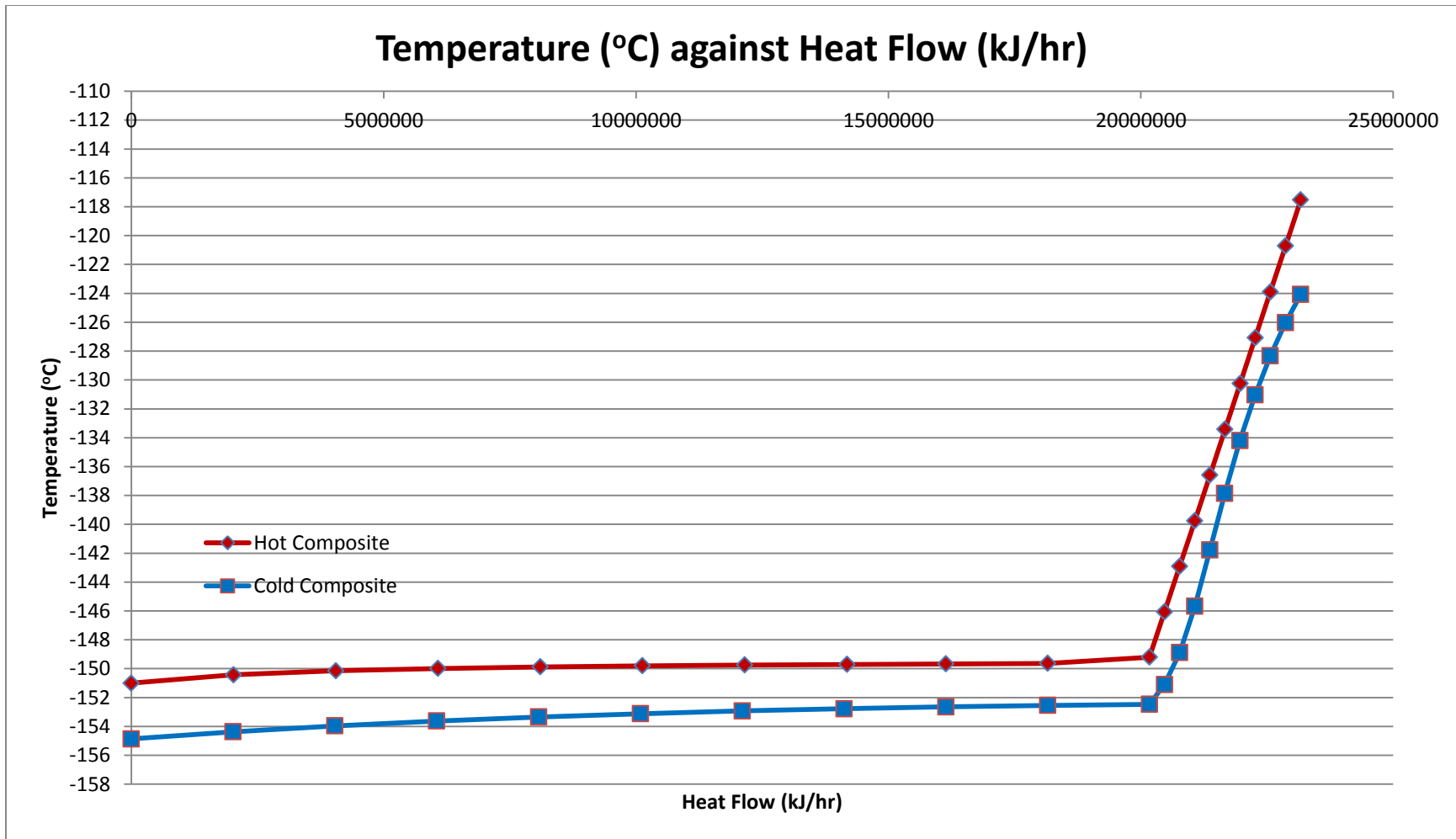
Figure 4.2: Temperature against Heat Flow for Mixed Refrigerant Cascade in E-101 and E-102

Graph of temperature against heat flow for both the cases of pure refrigerant cascade and mixed refrigerant cascade are plotted respectively in Figure 4.1 and Figure 4.2. Referring to Figure 4.1, end flash gas at the outlet of the end flash drum at a temperature of -117.4℃ is being subcooled to -146.6℃ before undergone phase change to liquid state and cooled again to -151.4℃. The sharp bending point of the hot composite indicates the phase change of end flash gas in vapor state into liquid state. The cold composite shows the temperature profile of pure methane, which has been used as the refrigerant for the purpose of heat exchange with the end flash gas stream. The huge vertical spacing between the composite curves reflect the ineffectiveness of heat exchange in E-101 for pure refrigerant cascade, resulting in large amount of refrigerant needed and thus increases the heavy load imposed on the compressors in end flash system. In Figure 4.2, the hot composite curve undergoes almost the identical shape as the hot composite curve in Figure 4.1 as the molar composition of each component in the stream is quite similar. While for the cold composite curve, it shows a sharp bending point at 152.2℃ as it is made up of two types of refrigerant with different molar composition. The molar composition for first type of refrigerant is 78% of methane and 22% of nitrogen while the second type of refrigerant is 92% of methane and 8% of ethane. The purpose of using two different set of refrigerants is aimed at closing the gap of the cold composite curve and the hot composite curve. The vertical spacing between the curves has a direct impact to the heat transfer efficiency and indirectly imposes extra load on compressor duty. A point to note is that the gap between the cold composite curve and hot composite curve is always more than 3℃ as to ensure optimum heat transfer across the streams.

It is not the ineffectiveness of cold box E-101 alone that contributes to the significant machinery power difference between the two optimization cases discussed. Instead, the similarly reasoning can be applied to the cold box E-102and E-103 in pure refrigerant cascade as well as E-103, E-104 and E-105 in mixed refrigerant cascade. The graph of temperature against heat flow of E-102 and E-103 of pure refrigerant cascade are shown in Figure 4.3 and Figure 4.4 while the graph of temperature against heat flow of E-103, E-104 and E-105 of mixed refrigerant cascade are shown in Figure 4.5, Figure 4.6 and Figure 4.7.

Figure 4.3: Temperature against Heat Flow for Pure Refrigerant Cascade in E-102

Figure 4.4: Temperature against Heat Flow for Pure Refrigerant Cascade in E-103

Figure 4.5: Temperature against Heat Flow for Mixed Refrigerant Cascade in E-103

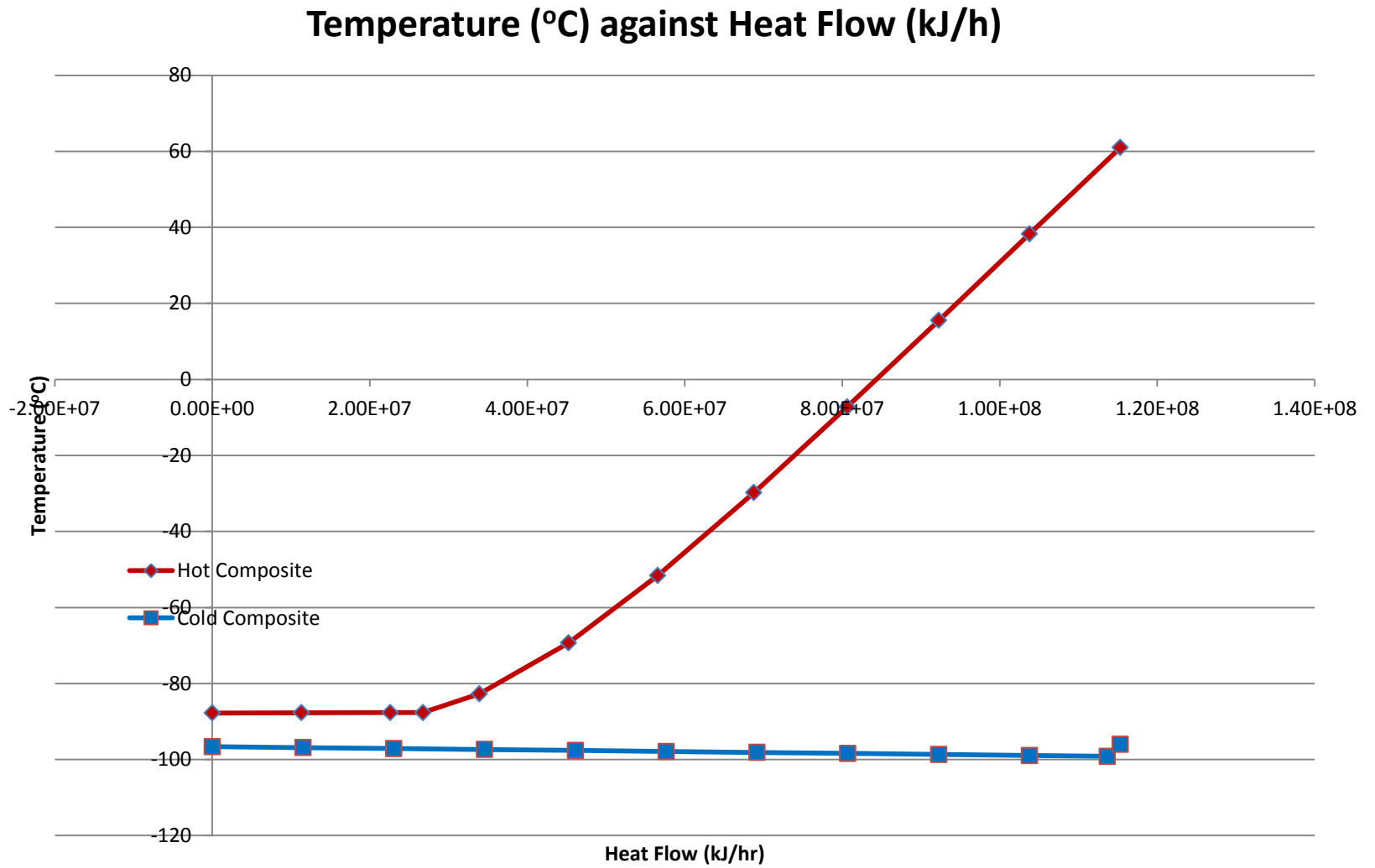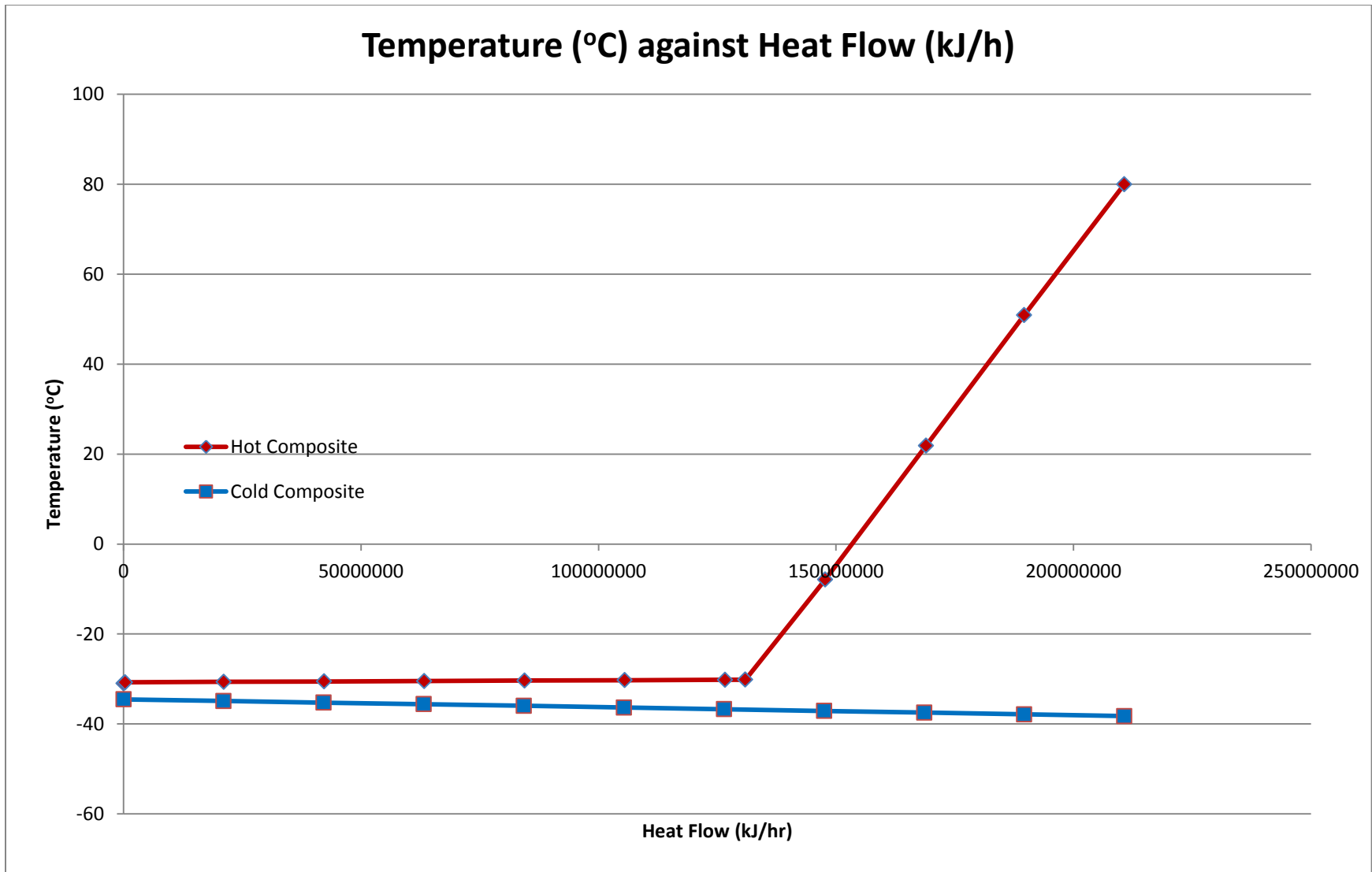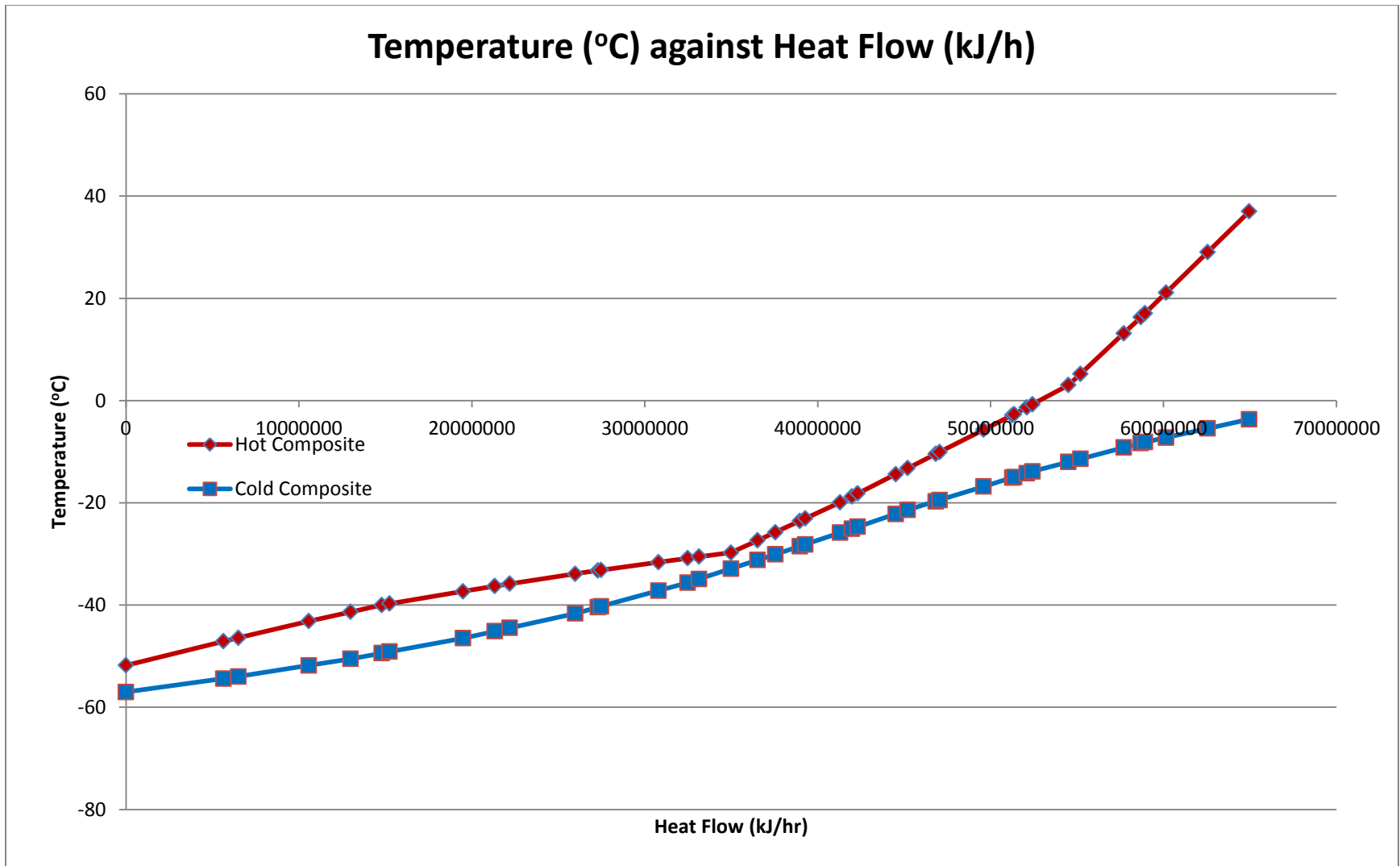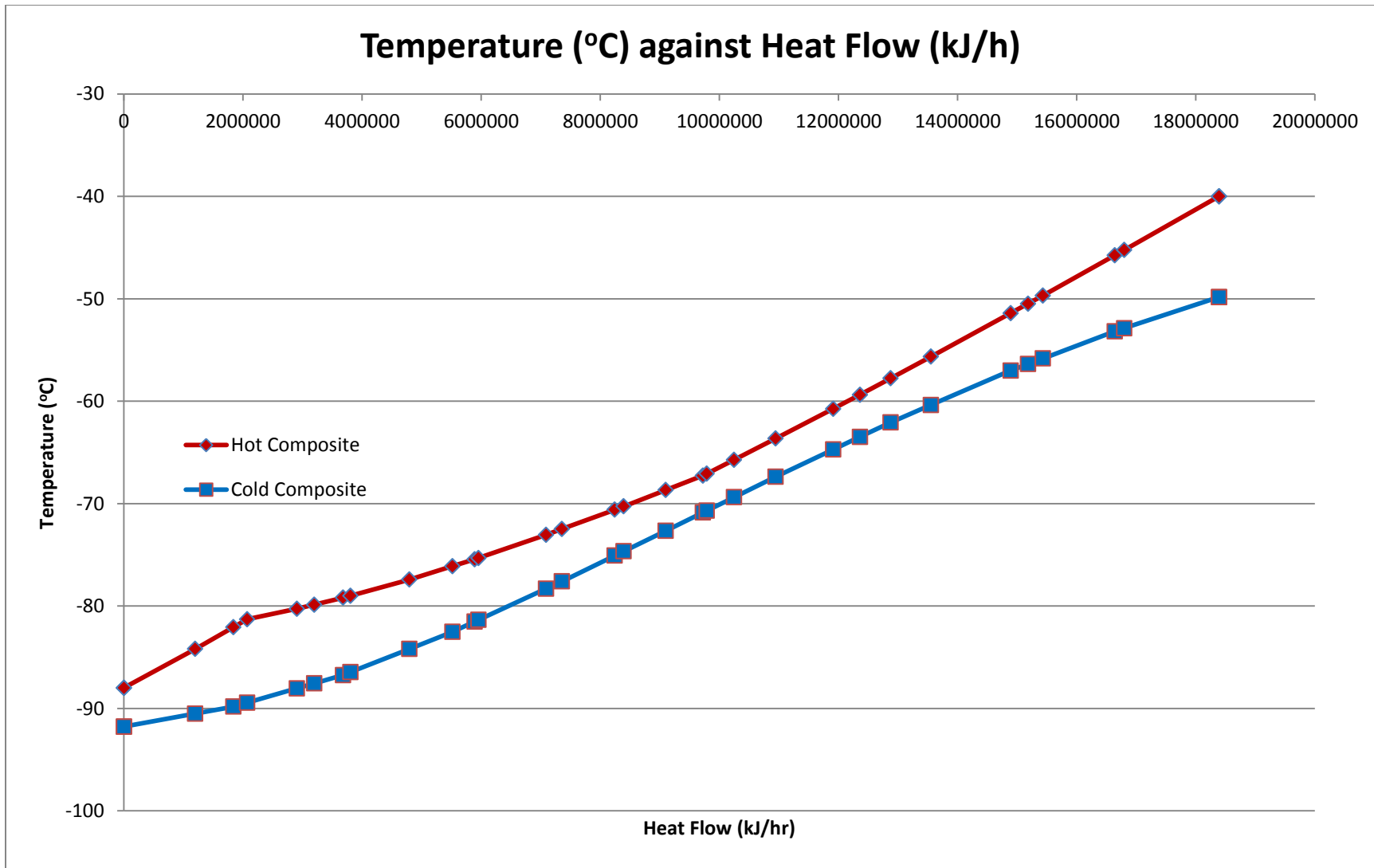**Figure 4.6**: Temperature against Heat Flow for Mixed Refrigerant Cascade in E-104
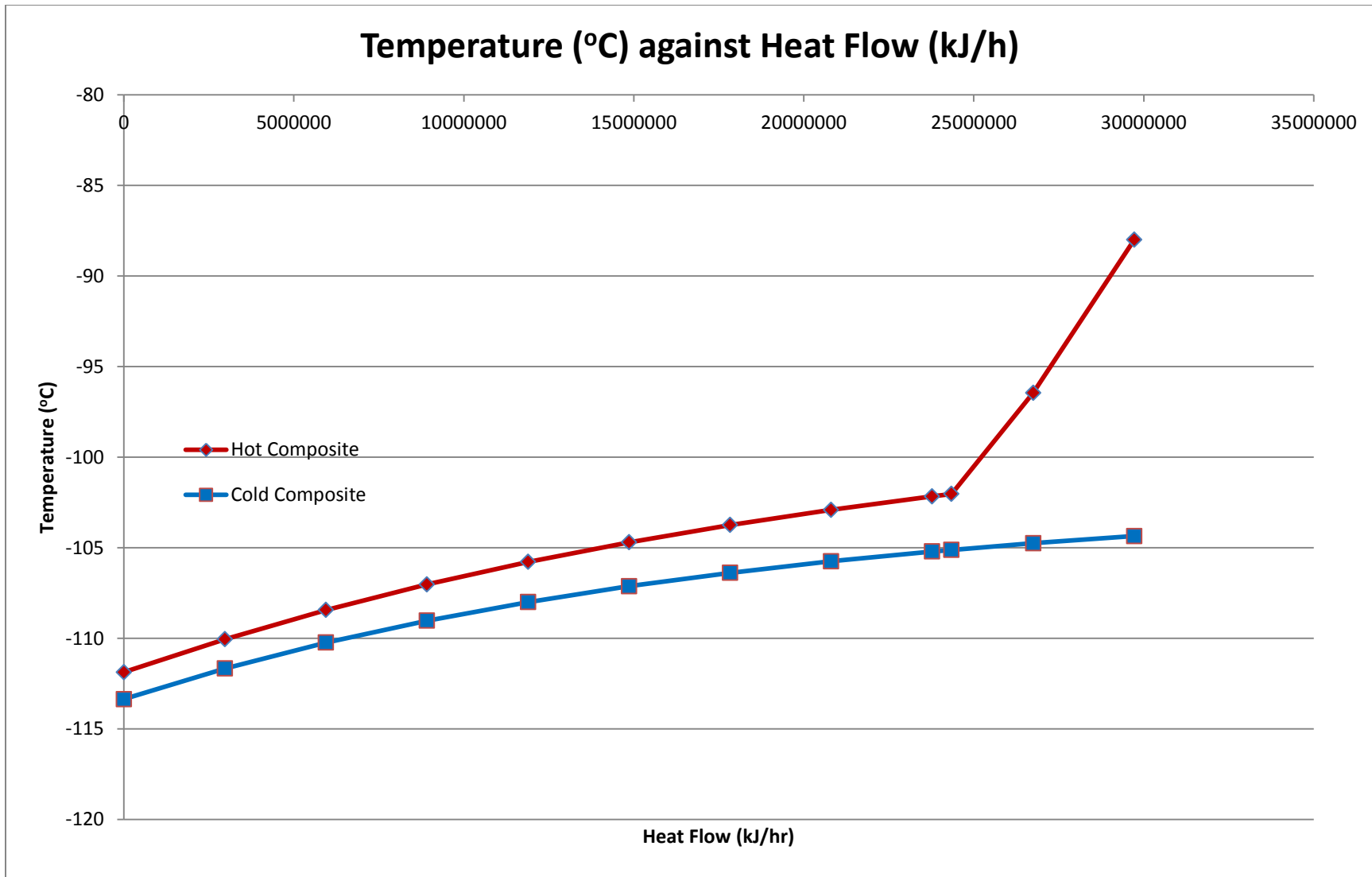
Figure 4.7: Temperature against Heat Flow for Mixed Refrigerant Cascade in E-105

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4.2 Economic Feasibility Study

4.2.1 Introduction to Economic Analysis

Economic analysis is equally important to the technical feasibility study as both of the components are integral for the success of any engineering project. According to Smith (2010) in his workbook entitle Chemical Process Design and Integration, the cost of a specific item of equipment can be formulated as shown:

$$C_E = C_B \left(\frac{Q}{Q_B}\right)^M \times \frac{Cost\ Index\ at\ Year\ of\ Interest}{Cost\ Index\ at\ Year\ of\ Design} \qquad (4.1)$$

Where C_E = equipment cost for carbon steel at moderate pressure and temperature with the capacity Q

C_B = known base cost for equipment with capacity

M = constant depending on equipment type

The cost index that are being used in this report is the Chemical Engineering Indexes, with the cost index at the year of design is set at January 2000 with a CE Composite Index of 391.1. The cost index suggested at October 2012 is 575.4.

The corrected cost of equipment will be a function of size, materials of construction, design pressure, and design temperature. Mathematically, the corrected cost of equipment can be described as follow:

$$C_E = C_B \left(\frac{Q}{Q_B}\right)^M f_M f_P f_T (1 + f_P) \qquad (4.2)$$

f_M = correction factor for material of construction

f_P = correction factor for material for design pressure

f_T = correction factor for material for design temperature

f_P = correction factor for material for piping installation

4.2.2 Equipment Costs and Annual Revenue

Mixed refrigerant cascade will be studied for economic feasibility only as there is no any improvement in terms of production or energy consumption for pure refrigerant cascade. Only the major equipment, such as the end flash drum, heat exchangers, pumps and compressors, will be considered in the economic feasibility studies. The difference of equipment costs between the systems will be treated as the additional capital investment. The plant life is taken to be 15 years and the minimum acceptable rate of return is 10%. The purpose of this economic analysis is to study the feasibility of installation of the mixed refrigerant cascade over the improved end flash system in terms of cash flow diagram, payback period, net present worth (NPW), interest rate of return (IRR) and return on investment (ROI). Only the cost of equipment and the installation costs are being considered at this stage of study, detailed economic analysis will be covered elsewhere. The equipment costs for both the improved end flash system and mixed refrigerant system are shown respectively in Table 4.2 and Table 4.3:

Table 4.2: Estimation Costs for Improved End Flash System

Tag Number	Quantity	Equipment Costs (USD)	Corrected Costs (USD)
E-101	1	822,220	5,702,919
E-102	1	660,460	4,580,953
E-103	1	382,224	2,651,106
E-104	1	968,767	6,719,371
E-105	1	752,308	5,218,005
E-106	1	1,521,367	10,552,202
P-101	1	30,232	51,395
V-101	1	2,217,779	12,667,951
K-101	1	396,315	1,616,964
K-102	1	422,677	1,724,524
K-103	1	744,567	3,037,835
K-104	1	881,343	3,595,881
K-105	1	1,234,637	5,037,319
Total Capital Investment		11,034,897	63,156,424

Table 4.3: Estimation Costs for Mixed Refrigerant Cascade System

Tag Number	Quantity	Equipment Costs (USD)	Corrected Costs (USD)
E-101	1	578,357	4,011,484
E-102	1	2,686,863	18,636,084
E-103	1	2,937,855	20,376,966
E-104	1	1,617,017	11,215,627
E-105	1	3,821,546	26,506,245
E-106	1	205,218	1,423,395
E-107	1	1,450,796	10,062,721
E-108	1	667,348	4,628,726
E-109	1	319,912	2,218,907
P-101	1	30,320	51,544
V-101	1	2,160,689	12,341,857
K-101	1	277,407	1,131,822
K-102	1	407,851	1,664,031
K-103	1	658,216	2,685,521
K-104	1	868,929	3,545,232
K-105	1	721,969	2,945,633
K-106	1	769,273	3,138,632
Total Capital Investment		20,179,567	126,584,428

The difference of total price in Table 4.2 and Table 4.3 reflects the additional costs of installation of pure refrigerant cascade unit over the improved end flash system in a typical LNG plant. The additional of USD 63.4 million will thus be considered as the capital investment needed and the profitability tools will be used to justify the feasibility of the system. For instances, the engineering project is said to be desirable if the NPW is positive at the end of project life or the IRR is higher than the MARR. The estimation of additional annual revenue is formulated as shown:

$$Additional\ Annual\ Revenue = \frac{(7.43 - 7.40) \times 10^9}{1W/3.412\ Btu} \times \frac{USD\ 16.85}{10^6\ Btu} \times \frac{7920\ hr}{year}$$

$$= USD\ 13,660,146\ /year$$
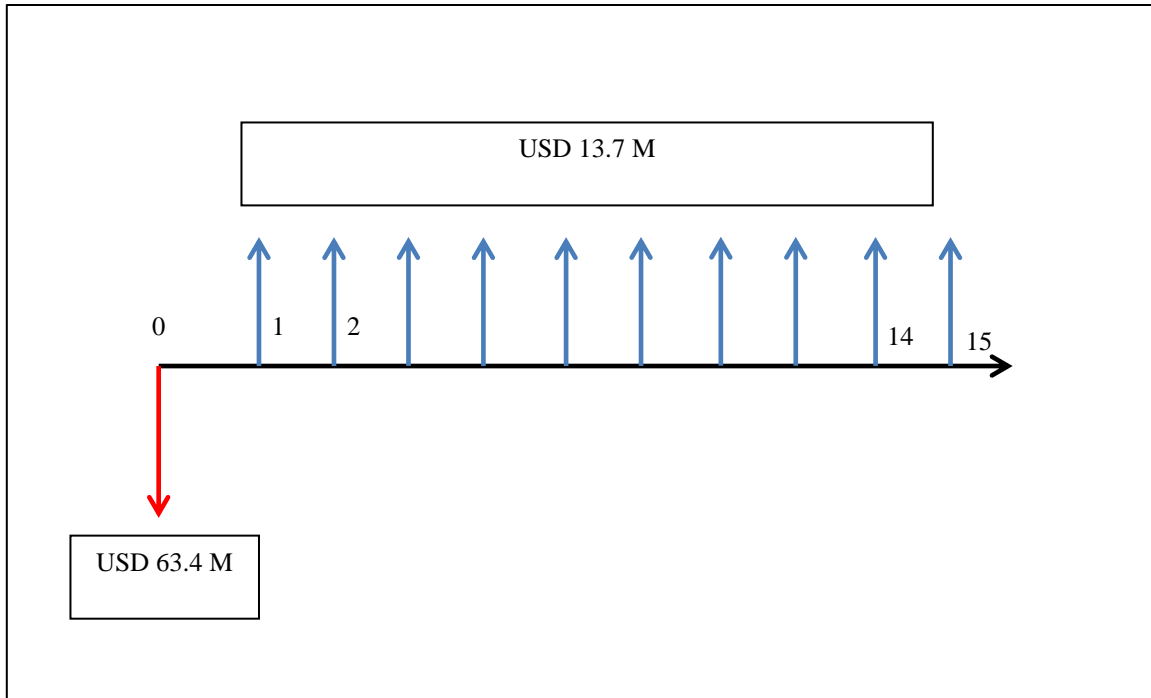
Figure 4.8: Cash Flow Diagram

Referring to the cash flow diagram, it is interesting to note that the annual production cost is assumed to be equal for both the cases considered, thus the annual expenses is taken to be zero throughout the project life. The decision of installation of mixed refrigerant cascade unit over improved end flash system will cost USD 63.4 million while have an annual revenue difference of USD 13.6 million. It is also assumed that the salvage value for both of the system are equivalent, thus the difference in salvage value will be zero as well as shown in Figure 4.8.

4.2.3 Profitability Analysis

NPW is one of the most reliable profitability tools to justify whether an engineering project is economically desirable. If an engineering project shows a positive NPW, thus the project is said to be profitable. Referring to Table 4.4, the NPW for the installation of mixed refrigerant cascade is USD 133 million. The positive value of NPW indicates installation of mixed refrigerant cascade is favorable over the installation of improved end flash system. Payback period is another profitability tools which is used to measure the amount of years needed for the project to reach the breakeven. According to Table 4.4 and Figure 4.9, the payback period for the engineering project is at the end of year 5. At the end of year 5, the project has a positive value of cumulative cash flow which stands at USD 10 million.

An engineering project is said to be favorable when the IRR is higher than the MARR. Figure 4.9 shows the graph of cumulative discounted cash flow against interest rate of return. Referring to Figure 4.10, the project of installation of mixed refrigerant cascade unit has an IRR of 17.8%, which is higher than the MARR. ROI can be formulated as shown:

$$ROI = \frac{Cummulative\ Cash\ Flow\ at\ Start\ of\ Project\ Life}{Project\ Life\ \times TCI} \times 100\%$$

$$= \frac{196,706,113}{15 \times 63,428,004} \times 100\%$$

$$= 20.68\%$$

Generally, as a performance indicator, ROI shows the efficiency of an investment or to compare the efficiency of a number of different investments. In the context of this engineering project, a positive value of 20.68% indicates that the installation of mixed refrigerant cascade is indeed economically beneficial if compared to the installation of improved end flash system. In other words, the additional investment of approximately USD 63.4 million is well justified, as the extra investment costs will yield greater profit to the LNG plant.

Table 4.4: Profitability Analysis of Installation of Mixed Refrigerant Cascade

| End of year | Sales of LNG, Btu/year | Selling price of LNG, USD/MMBtu | During year | | At year end | At commencement of project | | | |
			Revenue, USD	Investment, USD	Net cash flow, USD	Cumulative cash flow, USD	Discounted cash flow at 10%, USD	Cumulative discounted cash flow at 10%, USD	Discounted cash flow at 17.79%, USD	Cumulative discounted cash flow at 17.79%, USD
-2	0	-	0	0	0	0	0	0	0	0
-1	0	-	0	63,428,004	-63,428,004	-63,428,004	-57,661,822	-57,661,822	-53,850,183	-53,850,183
0	4.05346E+11	16.85	6,830,073	0	6,830,073	-56,597,931	5,644,689	-52,017,133	4,923,089	-48,927,095
1	7.29622E+11	16.85	12,294,132	0	12,294,132	-44,303,799	9,236,763	-42,780,370	7,523,437	-41,403,658
2	8.10691E+11	16.85	13,660,147	0	13,660,147	-30,643,652	9,330,064	-33,450,306	7,097,084	-34,306,574
3	8.10691E+11	16.85	13,660,147	0	13,660,147	-16,983,505	8,481,876	-24,968,429	6,025,403	-28,281,171
4	8.10691E+11	16.85	13,660,147	0	13,660,147	-3,323,358	7,710,797	-17,257,633	5,115,548	-23,165,623
5	8.10691E+11	16.85	13,660,147	0	13,660,147	10,336,788	7,009,815	-10,247,817	4,343,085	-18,822,538
6	8.10691E+11	16.85	13,660,147	0	13,660,147	23,996,935	6,372,559	-3,875,258	3,687,266	-15,135,272
7	8.10691E+11	16.85	13,660,147	0	13,660,147	37,657,082	5,793,236	1,917,978	3,130,478	-12,004,794
8	8.10691E+11	16.85	13,660,147	0	13,660,147	51,317,228	5,266,578	7,184,555	2,657,766	-9,347,028
9	8.10691E+11	16.85	13,660,147	0	13,660,147	64,977,375	4,787,798	11,972,353	2,256,435	-7,090,592
10	8.10691E+11	16.85	13,660,147	0	13,660,147	78,637,522	4,352,544	16,324,897	1,915,707	-5,174,885
11	8.10691E+11	16.85	13,660,147	0	13,660,147	92,297,669	3,956,858	20,281,755	1,626,429	-3,548,456
12	8.10691E+11	16.85	13,660,147	0	13,660,147	105,957,815	3,597,144	23,878,899	1,380,834	-2,167,623
13	8.10691E+11	16.85	13,660,147	0	13,660,147	119,617,962	3,270,131	27,149,029	1,172,323	-995,299
14	8.10691E+11	16.85	13,660,147	0	13,660,147	133,278,109	2,972,846	30,121,875	995,299	0
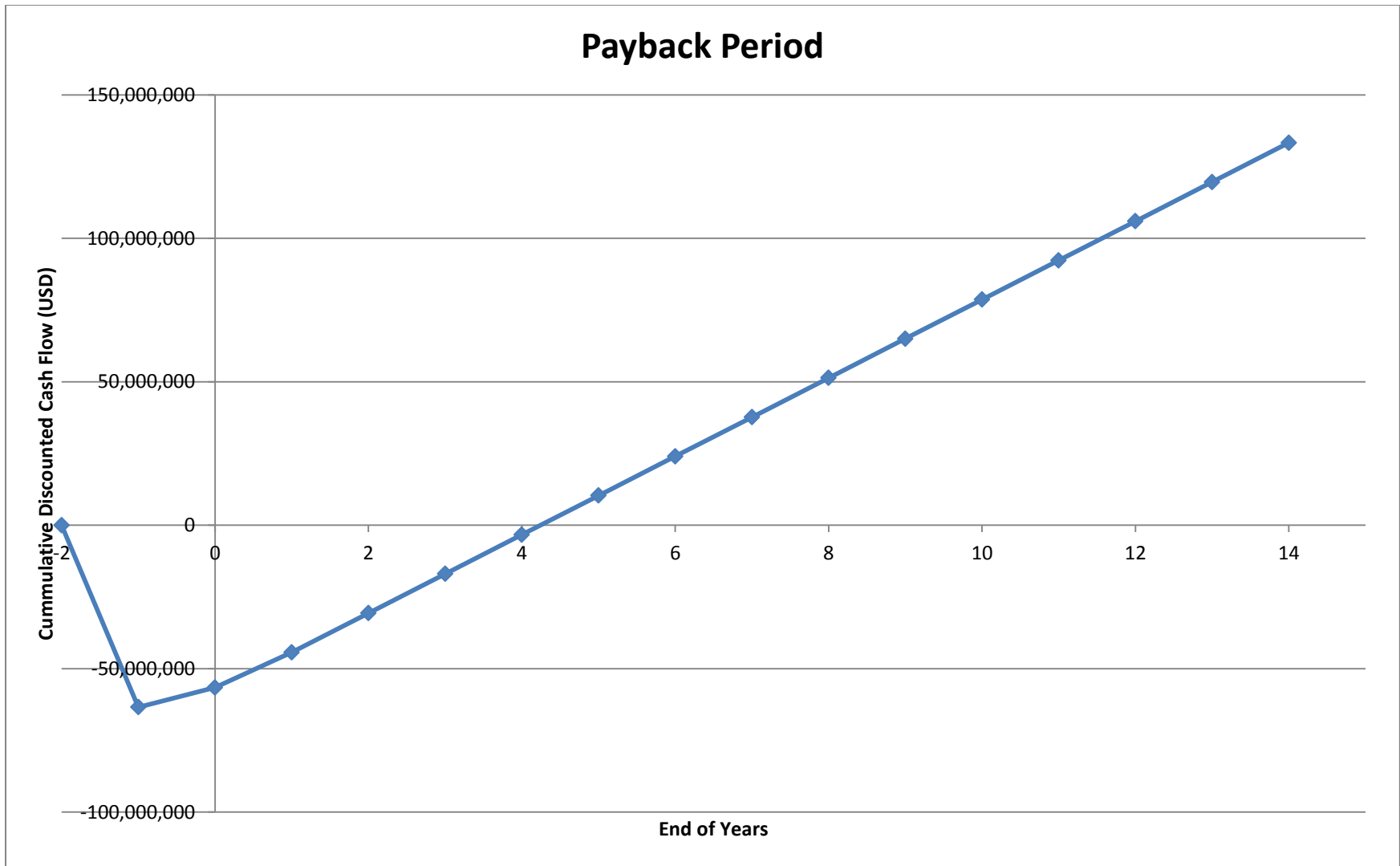
Figure 4.9: Graph of Payback Period

Figure 4.10: Graph of NPW against Interest Rates

Judging from the aspect of technical feasibility and economic analysis, the mixed refrigerant cascade will bring the greatest benefits to a typical LNG plant as mixed refrigerant cascade gives the lowest specific power of LNG production and highest ROI in all the three cases studied. Mixed refrigerant cascade serves as an upgrade to the improved end flash system as the total LNG per hour has been increase by 0.03 GW and the energy consumption is 22% lower than the improved end flash system. The difference in total LHV successfully brings additional annual revenue of approximately USD 13.6 million with acceptable investment costs. One point to note is that detailed economic analysis on either the mixed refrigerant cascade or improved end flash system is not covered in this paper. This paper compares both the cases and performs profitability analysis to investigate the additional profits brought by installation of mixed refrigerant cascade over improved end flash system. Thus, the value of payback period, NPW, IRR and ROI determined in this paper will differ from the standalone case.

It is interesting to note that the optimization is achieved as the uses of mixed refrigerant successfully reduce the heavy mechanical load imposed by the high compressor duty. This is done by closing the vertical spacing as close as possible to the region of 3℃ between the refrigerant and the cooling curve of the end flash gas as shown in Figure 4.1. The failure of pure refrigerant cascade is due to the fact that pure refrigerant offers little or no flexibility at all in different liquefaction stages of the end flash gas. For instances, referring to Figure 4.1, the use of pure refrigerant in subcooling the liquefied end flash gas is acceptable, however, the vertical spacing between the hot composite curve and cold composite curve in the precooling zone is way beyond the minimum approach temperature of 3℃, resulting in ineffectiveness in heat exchange. Such issue is addressed using the mixed refrigerant as the cold composite curve successfully assimilates the shape of hot composite curve while at the same time the temperature difference across the stream is always larger than 3℃ along the streams, as shown in Figure 4.2.

CHAPTER 5

CONCLUSIONS AND RECOMMENDATIONS

5. CONCLUSIONS AND RECOMMENDATIONS

5.1 Conclusions

Detailed studies on the base case and the alternative processes show that the mixed refrigerant cascade is indeed able to increase the LNG rundown to 542 t/h as compared to 540 t/h of LNG rundown in improved end flash system. The addition of 2 t/h results in a significant increment of total LHV produced by 0.03 GW and thus yields additional annual revenue of approximately USD 13.6 million. It is important to highlight that only 14 t/h of end flash gas is being treated as fuel gas for mixed refrigerant cascade while for the improved end flash system, it needs17 t/hr.

In line with the second objective of this project, which is to reduce the specific power of LNG production, the mixed refrigerant cascade offers an elegant solution by consuming only 273 kJ of energy to produce 1 kg of LNG while for the same parameter, improved end flash system need 351 kJ. The saving in energy consumption is due to the fact that mixed refrigerant cascade utilizes less machinery power than the improved end flash system. The economic benefit brought by mixed refrigerant cascade over the improved end flash system is justified by performing profitability analysis, such as the NPW, IRR, ROI and payback period. All the parameters show positive results which indicates that the installation of mixed refrigerant cascade is favorable over the base case. For instance, the IRR is 17.8% which is higher than MARR set at 10% and the ROI stands 20.68%. Thus, it is suffices to say that the installation of mixed refrigerant cascade is favorable from the technical aspect and economics analysis.

5.2 Recommendations

Further studies can be done to investigate the effect of different qualities of natural gas on the end flash system. For instance, the molar composition of natural gas that has been defined elsewhere in this paper is basically low nitrogen content natural gas. In the near futures, as more exploration and production of natural gas will be located in harsh conditions in which the natural gas might have high nitrogen content, the studies of different qualities of natural gas is indeed inevitable as the presence of nitrogen will affect the calorific value of the LNG throughput. Besides that, denitrogen column should be used in the place of end flash drum as end flash drum is not able to separate nitrogen from the LNG stream effectively (Paradowski, 2005). In results, slight modifications on the optimization case are indeed needed to address for high nitrogen content of natural gas.

Besides that, standalone economic analysis on the mixed refrigerant cascade should be taken into consideration to have a better understanding on the profitability of the system. It is of the author interest to compare the economics value brought by both the improved end flash system and mixed refrigerant cascade in order to determine the better solution among the cases discussed. The NPW, IRR, ROI and payback period determined in this paper do not necessary reflects the true value of the economics benefit brought by the optimization case.

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