PROCESS DEBOTTLENECKING OF SMALL SCALE LIQUEFIED NATURAL GAS (LNG) PLANT

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PROCESS DEBOTTLENECKING OF SMALL SCALE LIQUEFIED NATURAL GAS (LNG) PLANT

NURUL AIN BINTI RAMLI

A report submitted in partial fulfillment of the requirement for the award of the degree of Bachelor of Chemical Engineering (Gas Technology)

Faculty of Chemical and Natural Resources Engineering
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APRIL 2009

I declare that this thesis entitled “Process Debottlenecking of Small Scale Liquefied Natural Gas (LNG) Plant” is the result of my own research except as cited in references. This thesis has not been accepted for any degree and is not concurrently submitted in candidature of any other degree

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"To my beloved mother and father, family and someone special who gave me encouragement toward this study"

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ABSTRACT

Process debottlenecking is one of a process improvement and particularly important when the current conditions of a plant reaches maximum production rate without satisfying market demands. In the case of liquefied natural gas (LNG), transcontinental demands always show steady increment. Since LNG is a source of clean energy and a feedstock to chemical productions, process debottlenecking of existing LNG plants offers economic benefits. In this research, the first objective was to perform process debottlenecking of a published flow sheet of a small scale LNG plant by using Aspen HYSYS. “Bottlenecks” or unit operations which reach bottlenecked conditions were identified by increasing the inlet flow rate. Simulation results showed that, LNG heat exchanger 2 was the single bottlenecked unit operations identified due to the occurrence of temperature cross. The bottlenecks removal was then performed by transferring duty of this active bottleneck to an additional cooler installed. Five different modifications were designed which installation of a cooler at different stream in the flow sheet and applied. Modification 5 showed the highest percentage of LNG production with 5298.68% increment from the existing plant. Production revenue for this modification is RM 2552098.60 after takes into consideration the highest cost of its additional cooling duty which is RM 400000.00. General economic benefits for this work need to be further analyzed so that the importance of process debottlenecking of LNG plant become more comprehensive.

ABSTRAK

Proses menyingkirkan gangguan terhadap sesuatu kerja adalah salah satu proses kemajuan dan penting pada masa kini terutamanya apabila hasil pengeluaran tertinggi tidak memenuhi kehendak pasaran. Merujuk kepada kes Gas Asli Cecair, bahan ini selalunya menunjukkan permintaan yang memberangsangkan di seluruh dunia. Memandangkan Gas Asli Cecair adalah sumber tenaga yang bersih dan keperluan bagi produksi kimia, proses ini menawarkan kelebihan daripada segi ekonomi. Penyelidikan ini mempunyai dua objektif. Objektif pertama adalah untuk melakukan proses penyingkiran gangguan pada kertas kajian loji Gas Asli Cecair skala kecil menggunakan kaedah simulasi melalui pengsimulasi Aspen HYSYS. Unit operasi yang mengalami masalah akan dikenalpasti melalui kaedah menaikkan kadar pengaliran awal sesuatu proses. Hasil simulasi memunjukkan bahawa alat penukaran haba yang kedua adalah unit operasi yang mengalami gangguan melalui pengecaman ketika berlakunya perselisihan suhu. Proses penyingkiran ini akan melibatkan pemindahan duti pada unit operasi yang mengalami gangguan kepada alat penyejuk tambahan. Terdapat lima modifikasi yang telah dilakukan dan diaplikasi melalui penambahan bahan penyejuk pada aliran yang berbeza. Modifikasi kelima menunjukkan hasil pengeluaran yang tertinggi iaitu kenaikan sebanyak 5298.68% daripada loji asal. Hasil pengeluaran untuk modifikasi ini adalah RM 2552098.60 selepas melalui pertimbangan pembelian alat penyejuk tambahan iaitu RM 400000.00. Analis ekonomi bagi penyelidikan ini dicadangkan dianalis secara lebih telus lagi supaya kepentingan proses ini dapat dilihat secara lebih menyeluruh.

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CHAPTER 1

INTRODUCTION

1.1 World LNG Trading

Natural Gas demand is expected to increase nearly 40% from 22 Trillion cubic feet to 31 Trillion cubic feet between 2002 and 2025. (National Energy Technology, Future Supply and Emerging Resources Liquefied Natural Gas). According to the Energy Information Administration, world natural gas consumption and production are expected to increase by more than 50 percent from 2005 through 2030. Asia is expected to become the world's number one gas consumer, taking over that spot from North America, as China's economy grows 6.4 percent annually. Non-Organization for Economic Co-operation and Development (OECD) countries are expected to account for more than 70 percent of the world's total growth in consumption and production of natural gas over the forecast period. A significant portion of the non-OECD production growth is expected to be in the form of export projects, particularly LNG projects. World LNG trade is projected to more than double by 2030, with the center of the trade moving away from northeast Asia toward an even Atlantic/Pacific basin split. Figure 1.1 shows us the world natural gas reserves by geographic region and Figure 1.2 illustrates the world natural gas production.

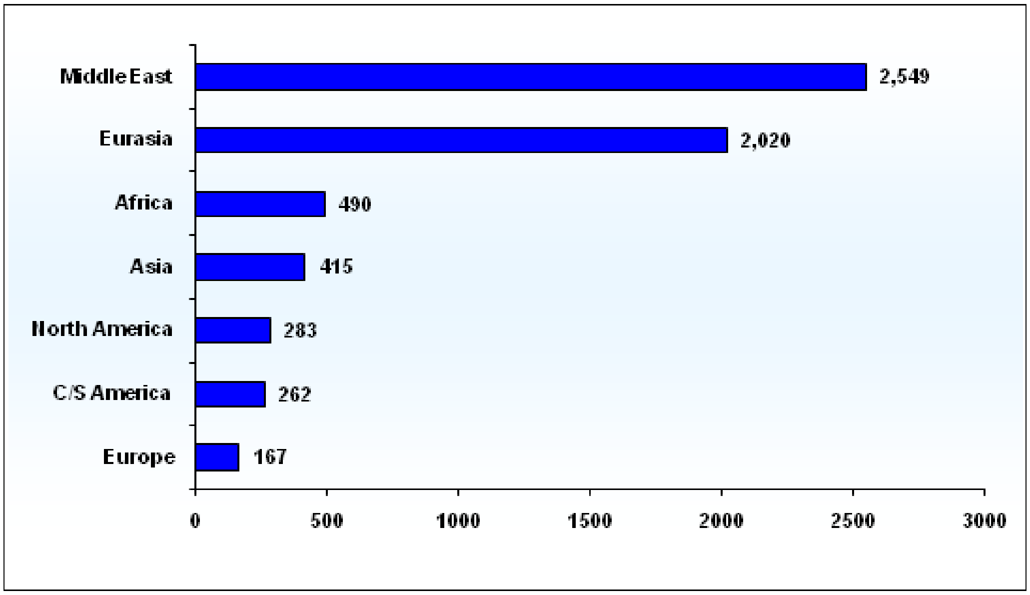

Figure 1.1: World Natural Gas Reserves by Geographic Region as of January 1, 2008, Sources: Energy Information Administration (EIA), Oil & Gas Journal, Vol. 105, No. 48 (December 24, 2007), pp. 24-25.

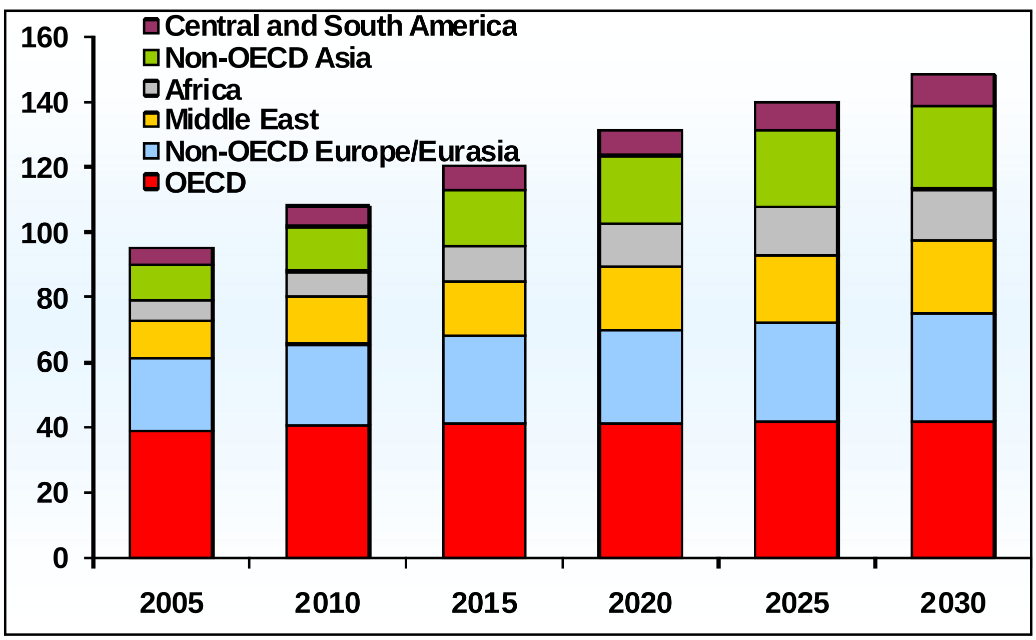

Figure 1.2: World Natural Gas Production (Trillion Cubic Feet), Sources: Energy Information Administration (EIA), International Energy Outlook 2008.

The efficient and effective movement of natural gas from producing regions to consumption regions requires an extensive and elaborate transportation system. In many instances, natural gas produced from a particular well will have to travel a great distance to reach its point of use. The transportation system for natural gas consists of a complex network of pipelines, designed to quickly and efficiently transport natural gas from its origin, to areas of high natural gas demand.

Generally, the limitations of the supply natural gas because of the above complexities can be solved by converting the phase of natural gas to become liquid, known as liquefied natural gas (LNG). By having LNG, the volume of natural gas can be reduced by about 600-fold which make it can be stored and transported in a huge amount compared to pipeline. It is also more economical to transport between continents in specially designed ocean vessel, whereas traditional pipeline transportation system would be less economically attractive and could be technically and politically infeasible. On the other hand, because transportation of natural gas is closely linked to its storage, then, liquefaction of natural gas provides the greatest opportunity to store natural gas for use during high demand periods in area where geologic conditions are not suitable for developing underground storage facilities. For example, in the northeastern part of United States, which is a region lacking in underground storage; LNG is a critical part of the region's supply during cold snaps. On the other hand, in region where pipelines capacity from supply area can be very expensive and use is highly seasonal, liquefaction and storage of LNG occurs during off-peak periods in order to reduce expensive capacity commitments during peak-periods. From the above discussion, we can conclude that LNG technology makes natural gas available throughout the world.

Liquefied natural gas is a natural gas that has been liquefied or converted to liquid form by reducing the temperature below -161 ^{0}C (260 ^{0}F) at 1 atm. Liquefied natural gas is primarily methane, nature's simplest and most abundant hydrocarbon fuel. Methane is composed of one carbon and four hydrocarbon atom (CH_4). LNG gas takes up about 1/600 th the volume of natural gas at store burner tip. It is odorless, colorless, non-toxic, non-corrosive and clear fluid which is less than half the density of water (roughly 0.41 to 0.5 kg/L, depending on temperature, pressure and composition, compared to water at 1.0 kg/L). The liquefaction process involves of certain components, such as dust, helium, and heavy hydrocarbons, which could causes difficulty downstream. Then gas is sent to a liquefaction plant where additional processing removes the remaining water vapor and carbon dioxide from the gas. A refrigeration process turns it into a liquid and further purities the stream so that LNG is predominantly methane. It also contains small amounts of ethane, propane, butane and heavier alkanes. The purification process can be designed to give almost absolutely methane (Hoegh, LNG articles).

Natural gas and its component are used as fuel for generating electricity and as raw material to manufacture a wide variety of products, from fibers for clothing, to plastic for healthcare, computing, and furnishing. Besides being used as the power generation plants, for feed to chemical plants, LNG is also a very promising fuel for aero planes, new generation rockets and ground vehicles, either as direct fuel for engines or as fuel for fuel cells (Liu and You, 1999).

1.2 Natural Liquefaction Process

Liquefaction is carried out at a pressure determined by economics factors and generally accomplished in the range of temperature between -116 ^{0}C to 161 ^{0}C at near atmospheric pressure (Fischer-Calderon, 2003). A higher pressure reduces the energy required to liquefy the natural gas, since the temperature range during the liquefaction process rises, although the final sub cooling temperature remains unchanged. Natural gas is liquefied over a temperature interval owing to the presence of hydrocarbons other than methane. The initial liquefaction temperature is higher with increasing contents of heavy hydrocarbons. For instance, it may begin at around -10 ^{0}C and continue to a temperature close to the vapor-liquid equilibrium temperature of methane under pressure, around -100 ^{0}C. The liquid phase obtained is then sub cooled to the boiling point of LNG at atmospheric pressure.

1.2.1 Types of Natural Liquefaction Process

1.2.1.1 Cascade Process

The cascade produces LNG by employing several closed-loop discrete cooling circuits or stages. Each circuits is utilizing pure refrigerant and collectively configured in order of progressively lower temperatures and generally have multistage refrigerant expansion and compression, typically operating at different evaporation temperature levels. The first cooling circuit may utilize propane, the second circuit utilizes ethane, and the third circuit utilizes methane as the refrigerant. After compression, propane is condensed with cooling water/air, ethane is condensed with evaporating propane and methane is condensed with evaporating ethane. Figure 1.3 shows The Simplified Cascade Process in LNG Production (From CPI, 2006).

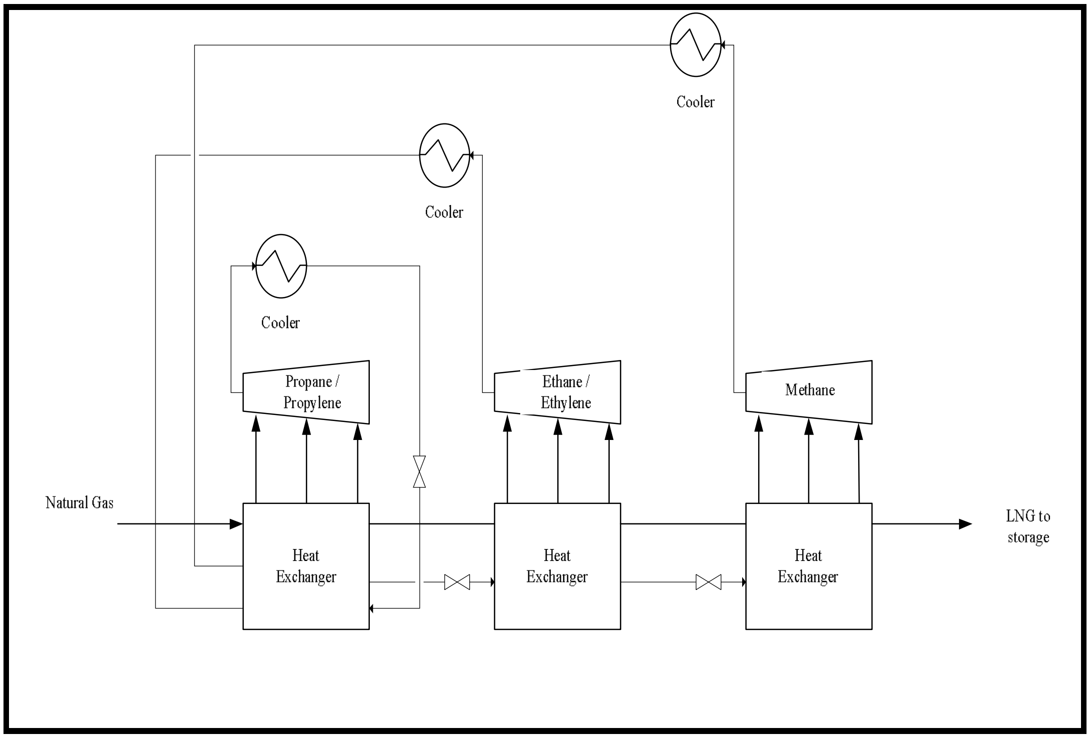

Figure 1.3: The Simplified Cascade Process in LNG Production (From CPI, 2006)

1.2.1.2 Single Mixed Refrigerant Process

A mixture of component having different volatilities, such as nitrogen, methane, ethane, propane and butane, is vaporized, by following in the enthalpy-temperature diagram a path of parallel to the one followed by the natural gas. This helps to liquefy the natural gas in a single mixed-refrigerant modified cascade cycle. In this cycle, the vaporization of a portion the liquid fractions obtained at increasingly lower temperatures serves to continue the condensation of the refrigerant mixture. The incorporation of the nitrogen makes it possible to sub cool to -160^{0}C, and thus avoid the loss of “flashed” gas by expansion, which occurs in the conventional cascade process. Figure 1.4 shows The Simplified Single Mixed Refrigerant Process (From Lee, 2000).

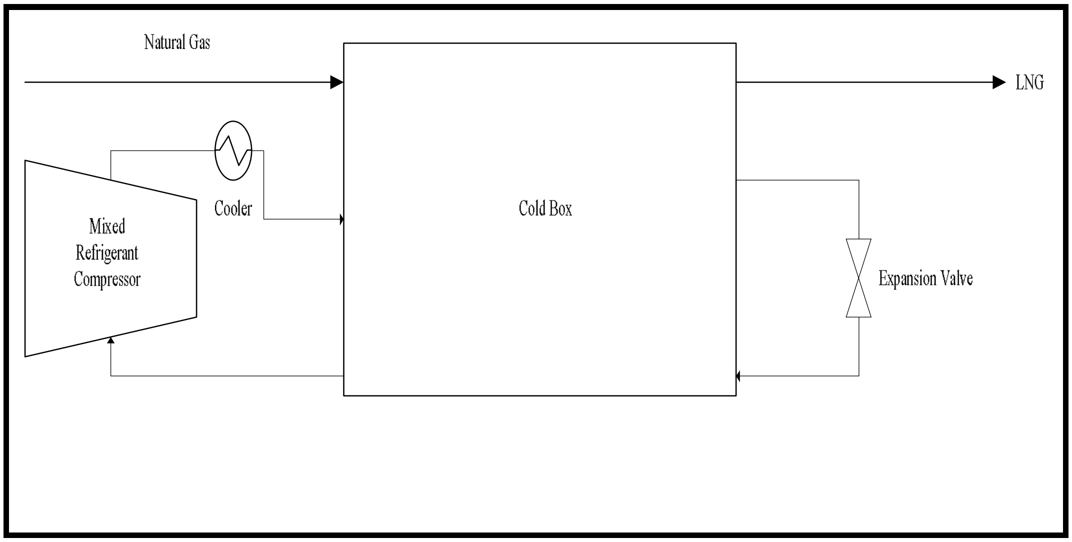

Figure 1.4: The Simplified Single Mixed Refrigerant Process (From Lee, 2000)

1.2.1.3 Propane Pre-cooled Mixed Refrigerant Process

The propane pre-cooled mixed refrigerant process, Figure 1.5, utilizes a mixed refrigerant (MR) that has a lower molecular weight and is composed of nitrogen, methane, ethane and propane. The natural gas feed is initially cooled by a separate propane chiller package to an intermediate temperature, about - 35°C (-31°F), at which the heavier components in the feed gas condense out and are sent to fractionation. The natural gas is then sent to the main heat exchanger, which is composed of a large number of small-diameter, spiral-wound tube bundles. These permit very close temperature approaches between the condensing and boiling streams. The MR refrigerant is partially condensed by the propane chiller before entering the cold box. The separate liquid and vapor streams are then chilled further before being flashed across Joule-Thompson valves that provide the cooling for the final gas liquefaction. Figure 1.5 shows The Propane Pre- cooled Mixed Refrigerant Process (From S. Mokhatab and Michael J.).

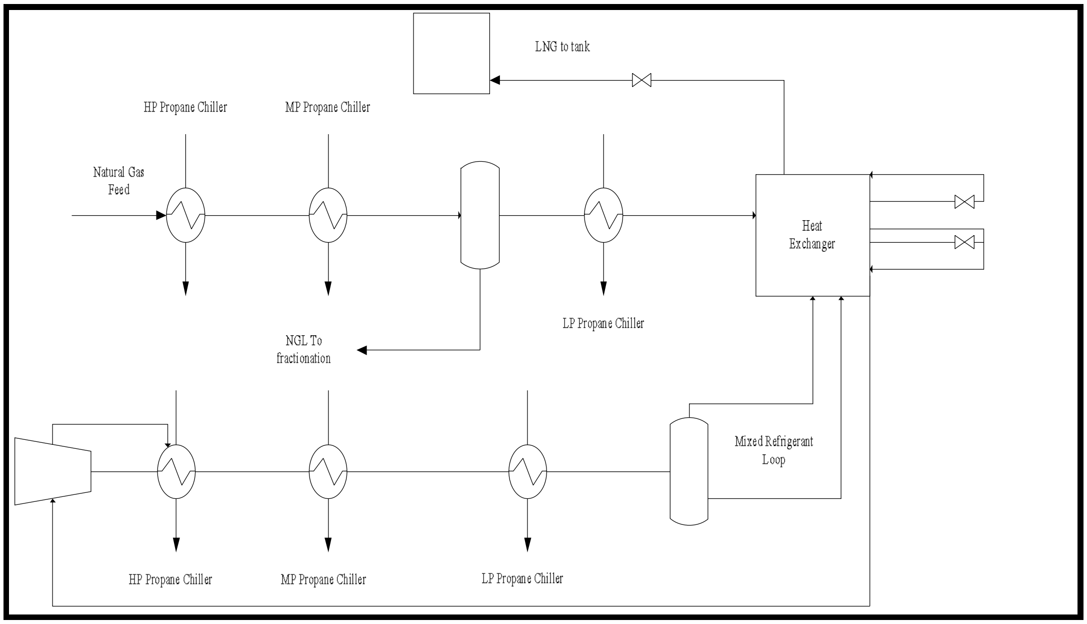

Figure 1.5: The Propane Pre-cooled Mixed Refrigerant Process (From S. Mokhatab and Michael J.)

1.2.1.4 Expansion Process

The expansions process expands natural gas from high pressure to low pressure with a corresponding reduction in temperature. As according to Joule-Thomson Effect, at which the expansion device such as turbo-expander, liquid turbine, and etc, must be adiabatic and reversible. It can be either isentropically or isenthalpically and operates on the principle that gas can be compressed to a selected pressure, cooled, and then allowed to expand. Figure 1.6 shows Simplified Expansion Process in LNG Production (From Barclay, 2005)

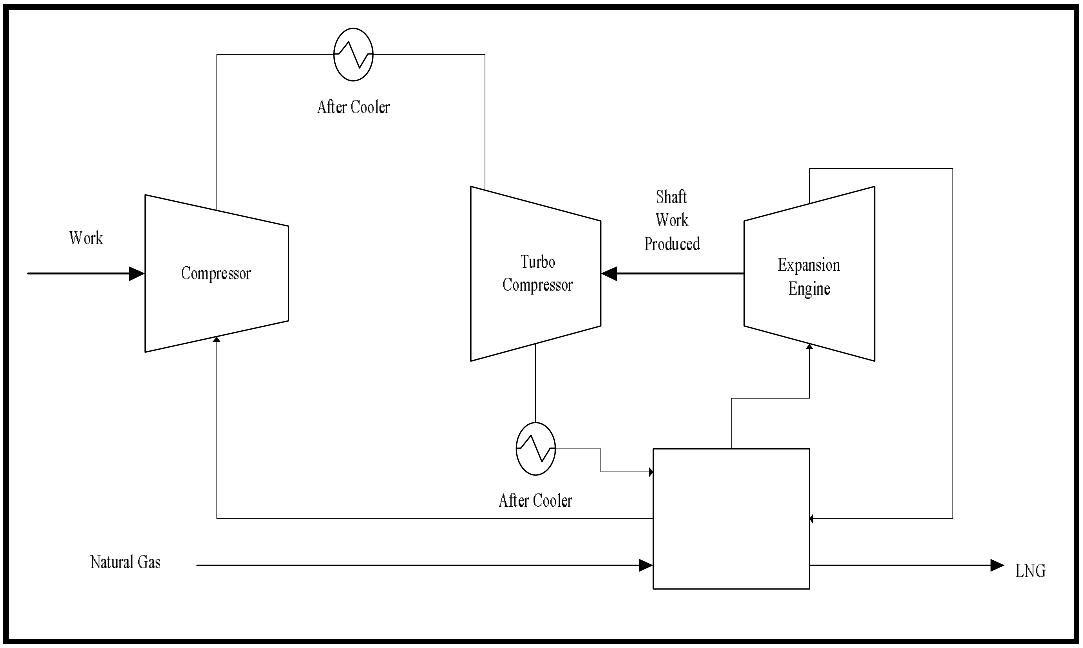

Figure 1.6: The Simplified Expansion Process in LNG Production (From Barclay, 2005)

1.3 LNG Cryogenic Plants

Natural gas liquefaction plants are generally classed as either peak-shaving or base-load plants depending on their size and role. These plants play an important role in order to deliver their annual capacity. The base load LNG plant usually use for marine (transcontinental) transportation. Nowadays, about 70 of base-load trains operating or under construction at 15 sites world wide and capable of producing from a single product line or train a capacity of up to 3.4 million tones per annum (Mtpa). In this type of plant, often two to three trains are installed to provide the required economies of scale. For the peak shaving plant, it facilities are usually small which is up to 0.9 million tones per annum (Mtpa). Peak shaving plant is also used to overcome mismatches between supply and demand. They liquefy and store excess natural gas during periods of low demand and vaporize it at times of peak demand (winter season). Besides above plants, other type of the LNG cryogenic plants is small scale plant. The opportunity of developing small scale natural gas liquefaction plants created from the continued commercial development of LNG vehicles. The markets for smaller-scale LNG liquefiers include onshore gas wells, customer sites that are remotely situated from current gas pipelines, and industrial customer peak shaving installations. Comparing with the medium-sized or large-scale liquefaction plant, the key characteristic of small-scale one are simple process, low investment, miniature size and skid-mounted package.

1.4 Techniques for Debottlenecking Process

Generally, bottleneck identification process can be classified into various types such as actual process performance and process experience. These methods involved the techniques of process simulation, hierarchical and heuristic which based on process experience, optimization which include a combination of process analysis and process synthesis, a two-stage debottlenecking process which combine the use of linear programming model at first stage followed by removal of bottlenecks, a developed algorithmic which applied to the retrofitting of an ammonia process and lastly, a combination between all these techniques. The most commonly used debottlenecking approach is the sequential method. In spite of its extensive usage, it is important to examine the ability of a sequential approach to attain the process true potential and in achieving maximum debottlenecking (Musaed, Nasser and Mahmoud, September, 2007). For this work, identification of the active bottleneck focuses on the techniques of process simulation using Aspen HYSYS simulator. Maximum debottlenecking can be achieved with the some development and alteration at the bottleneck conditions by applying the heuristic approach. As the bottleneck usually takes places in the equipments, then, evaluation in term of costing done based on the changing of the certain parameters.

1.5 Heat Transfer Equipment

In the process industries the transfer of heat between two fluids is generally done in heat exchangers. The most common type is one which the hot and cold fluids do not come into direct contact with each other but are separated by a tube wall or a flat or curved surface. The transfer of heat from the hot fluid to the wall or tube surface is accomplished by convection, through the tube wall or plate by conduction, and then by convection to the cold fluid.

Generally, shell and tube heat exchanger is used, which is the most important type of exchanger use in the process industries. In these exchangers the flows are continuous. Many tubes in parallel are used, where one fluid flows inside these tubes. The tubes, arranged in a bundle, are enclosed in a single shell and the other fluid flows outside the tubes in the shell side. The simplest shell and tube exchanger is shown in Figure 1.7(a) for one shell pass and one tube pass, or a 1-1 counter flow exchanger. The cold fluid enters and flows inside through all the tubes in parallel in one pass. The hot fluid enters at the other end and flows counter flow across the outside of the tubes. Cross baffles are used so that the fluid is forced to flow perpendicular across the tube bank rather than parallel with it. The added turbulence generated by this cross flow increases the shell side heat transfer coefficient.

In Figure 1.7(b), a 1-2 parallel counter flow exchanger is shown. The liquid on tube side flows in two passes as shown and the shell side liquid flows in one pass. In the first pass of the tube side, the cold fluid is flowing counter flow to the hot shell side fluid, in the second pass of tube side, the cold fluid flows in parallel (co-current) with the hot fluid. Another type of exchanger has two shell side passes and four tube passes. Other combinations of number of passes are also used sometimes, with the 1-2 and 2-4 types being the most common.

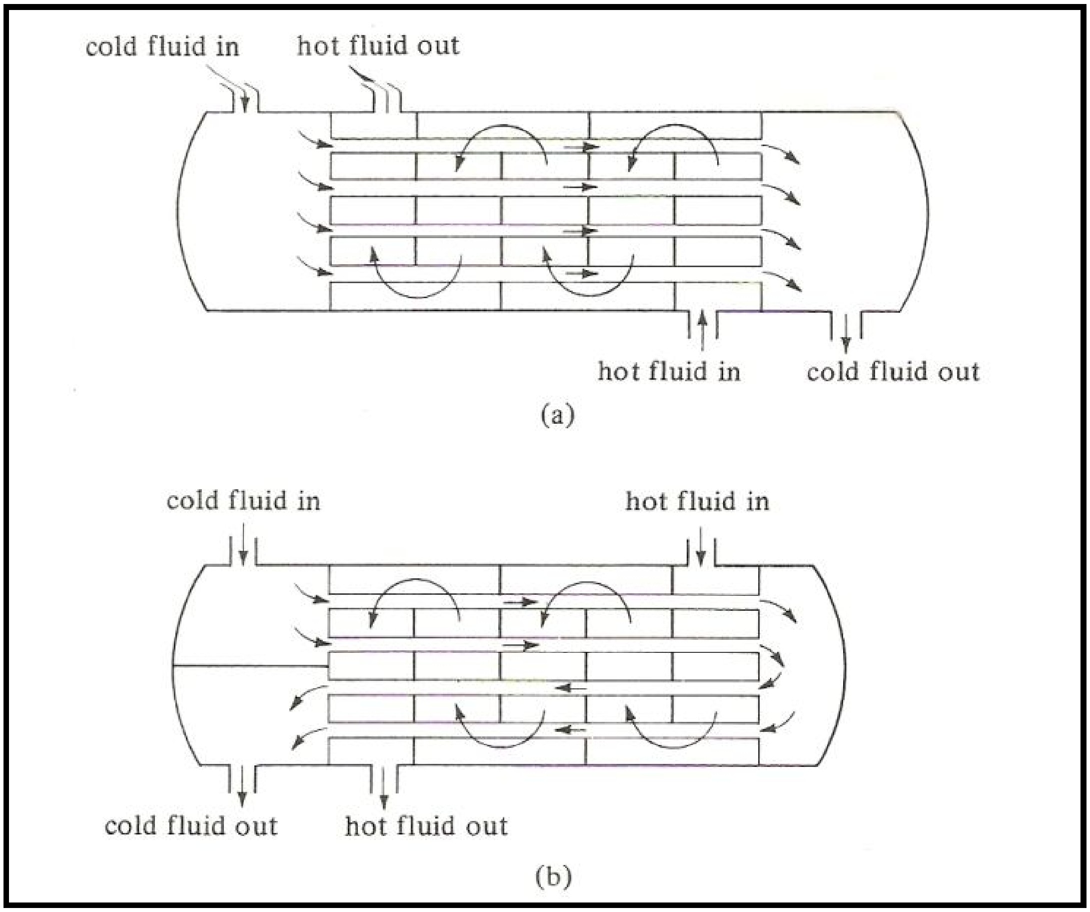

Figure 1.7: Shell and Tube Heat Exchanger; (a) 1 Shell Pass and 1 Tube Pass (1-1 Exchanger), (b) 1 Shell Pass and 2 Tube Passes (1-2 Exchanger)

1.5.2 Basic Design Procedure and Theory

The general equation for heat transfer across a surface is a $Q = UA\Delta T_m$, where :

Q = heat transfer per unit time, W,

U = the overall heat transfer coefficient, $W/m^2.^0C$,

A = heat transfer area, m^2,

ΔT_m = the mean temperature difference, the temperature driving force, ^0C.

The prime objective in the design of an exchanger is to determine the surface area required for the specified duty (rate of heat transfer) using the temperature difference available.

The overall coefficient is the reciprocal of the overall resistance to heat transfer, which is the sum of several individual resistances. For heat exchange across a typical heat exchanger tube relationship between overall coefficient and the individual coefficient, which are reciprocals of the individual resistances, is given by Equation 1.1:

$$\boxed{\frac{1}{U_o} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2k_w} + \frac{d_o}{d_i} \times \frac{1}{h_{id}} + \frac{d_o}{d_i} \times \frac{1}{h_i}} \qquad \text{(Equation 1.1)}$$

Where U_0 = the overall coefficient based on the outside area of the tube, $W/m^2.^0C$,

h_0 = outside fluid film coefficient, $W/m^2.^0C$,

h_i = inside fluid film coefficient, $W/m^2.^0C$,

h_{od} = outside dirt coefficient (fouling factor), $W/m^2.^0C$,

h_{id} = inside dirt coefficient, $W/m^2.^0C$,

k_w = thermal conductivity of the tube wall material, $W/m^2.^0C$,

d_i = tube inside diameter, m,

d_0 = tube outside diameter, m.

The magnitude of the individual coefficients will depend on the nature of the heat transfer process (conduction, convection, condensation, boiling or radiation), on the physical properties of the fluids, on the fluid flow rates, and on the physical arrangement of the heat transfer surface. As the physical layout of the exchanger cannot be determined until the area is known the design of an exchanger is of necessity a trial and error procedure.

1.5.3 Overall Heat Transfer Coefficient

Typical values of the overall heat transfer coefficient for various types of heat exchanger are given in Table 1.1. The values given in Table 1.1 can be used for the preliminary sizing of equipment for process evaluation, as trial values for starting a detailed thermal design.

Table 1.1: Overall Heat Transfer Coefficients

Shell and tube exchangers		
Hot fluid	Cold fluid	U (W/m^2 °C)
Heat exchangers		
Water	Water	800–1500
Organic solvents	Organic solvents	100–300
Light oils	Light oils	100–400
Heavy oils	Heavy oils	50–300
Gases	Gases	10–50
Coolers		
Organic solvents	Water	250–750
Light oils	Water	350–900
Heavy oils	Water	60–300
Gases	Water	20–300
Organic solvents	Brine	150–500
Water	Brine	600–1200
Gases	Brine	15–250
Heaters		
Steam	Water	1500–4000
Steam	Organic solvents	500–1000
Steam	Light oils	300–900
Steam	Heavy oils	60–450
Steam	Gases	30–300
Dowtherm	Heavy oils	50–300
Dowtherm	Gases	20–200
Flue gases	Steam	30–100
Flue	Hydrocarbon vapours	30–100
Condensers		
Aqueous vapours	Water	1000–1500
Organic vapours	Water	700–1000
Organics (some non-condensables)	Water	500–700
Vacuum condensers	Water	200–500
Vaporisers		
Steam	Aqueous solutions	1000–1500
Steam	Light organics	900–1200
Steam	Heavy organics	600–900

1.5.4 Fouling Factors (Dirt Factors)

Most process and service fluids will foul the heat transfer surfaces in an exchanger to a greater or lesser extent. The deposited material will normally have a relatively low thermal conductivity and will reduce the overall coefficient. It is therefore necessary to oversize an exchanger to allow for the reduction in performance during operation. The effect of fouling is allowed for in design by including the inside and outside fouling coefficients in Equation. Fouling factors are usually quoted as heat transfer resistances, rather than coefficients. They are difficult to predict and are usually based on past experience. Estimating fouling factors introduces a considerable uncertainty into exchanger design; the value of the other coefficients. Fouling factors are often wrongly used as factors of safety in exchanger design. Some work on the prediction of fouling factors has been done by HTRI; see Taborek *et al.* (1972). Fouling is the subject of books by Bott (1990) and Garrett-Price (1985).

Typical values for the fouling coefficients of common process and service fluids are given in Table 1.2. These values are for shell and tube exchangers with plain (not finned) tubes. More extensive data on fouling factors are given in the TEMA standards (1998), and by Ludwig (1965).

Table 1.2: Typical Value of Fouling Factor Coefficients

Fluid	Coefficient (W/m^2 °C)	Factor (resistance) (m^{2}°C/W)
River water	3000–12,000	0.0003–0.0001
Sea water	1000–3000	0.001–0.0003
Cooling water (towers)	3000–6000	0.0003–0.00017
Towns water (soft)	3000–5000	0.0003–0.0002
Towns water (hard)	1000–2000	0.001–0.0005
Steam condensate	1500–5000	0.00067–0.0002
Steam (oil free)	4000–10,000	0.0025–0.0001
Steam (oil traces)	2000–5000	0.0005–0.0002
Refrigerated brine	3000–5000	0.0003–0.0002
Air and industrial gases	5000–10,000	0.0002–0.0001
Flue gases	2000–5000	0.0005–0.0002
Organic vapours	5000	0.0002
Organic liquids	5000	0.0002
Light hydrocarbons	5000	0.0002
Heavy hydrocarbons	2000	0.0005
Boiling organics	2500	0.0004
Condensing organics	5000	0.0002
Heat transfer fluids	5000	0.0002
Aqueous salt solutions	3000–5000	0.0003–0.0002

The selection of the design fouling coefficient will often be an economic decision. The optimum design will be obtained by balancing the extra capital cost of a larger exchanger against the savings in operating cost obtained from the longer operating time between cleaning that the larger area will give. Duplicate exchangers should be considered for severely fouling systems.

1.5.5 Temperature Cross in the Heat Exchanger

Before equation 1.1 can be used to determine the heat transfer area required for a given duty, an estimate of the mean temperature difference ΔT_m must be made. This will normally be calculated from the terminal temperatures differences: the difference in the fluid temperatures at the inlet and outlet of the exchanger. The well known "logarithmic mean" temperature difference is only applicable to sensible heat transfer in true co-current flow or counter-current flow (linear temperature enthalpy curves). For counter-current flow, Figure 1.9(a), the logarithmic mean temperature is given by Equation 1.2:

$$\boxed{\Delta T_{\text{lm}} = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \dfrac{(T_1 - t_2)}{(T_2 - t_1)}}}$$

(Equation 1.2)

Where ΔTl_m = log mean temperature difference,

T_1 = inlet shell side fluid temperature,

T_2 = outlet shell side fluid temperature,

t_1 = inlet tube side temperature,

t_2 = outlet tube side temperature.

The equation is the same for co-current flow, but the terminal temperature differences will be (T_1 - t_1) and (T_2 - t_2). Strictly, Equation 1.2 will not apply when there is no change in the specific heats, the overall heat transfer coefficient is constant, and there are no heat losses. In design, these conditions can be assumed to be satisfied providing the temperature change in each fluid stream is not large.

In most shell and tube exchangers the flow will be a mixture of co-current, counter-current and cross flow. Figure 1.8(b) and (c) show typical temperatures profiles for an exchanger with one shell pass and two tube passes (a 1: 2 exchanger).

Figure 1.8(c) shows a temperature cross where the outlet temperature of cold stream is above that of the hot stream.

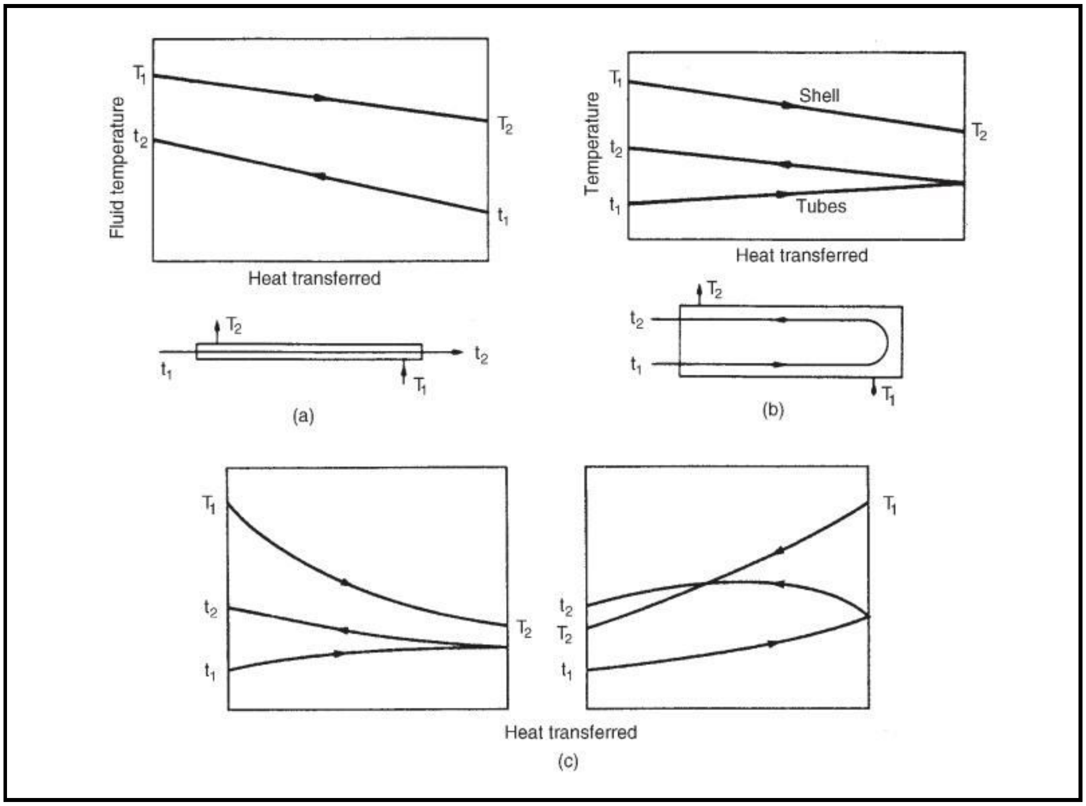

Figure 1.8: Temperature Profiles; (a) Counter-Current Flow, (b) 1:2 Exchanger, (c) Temperature Cross

The usual practice in the design of shell and tube exchanger is to estimate the "true temperature difference" from the logarithmic mean temperature by applying a correction factor to allow for the departure from true counter-current flow:

$$\boxed{\Delta T_m = F_t \Delta T_{\text{lm}}} \qquad \text{(Equation 1.3)}$$

where ΔTl_m = true temperature difference, the mean temperature difference for use in the design Equation 2.1,

F_T = the temperature correction factor.

The correction factor is a function of shell and tube fluid temperatures, and the number of tube and shell passes. It is normally correlated as a function of two dimensionless temperature ratios:

$$R = \frac{(T_1 - T_2)}{(t_2 - t_1)}$$

(Equation 1.4)

and

$$S = \frac{(t_2 - t_1)}{(T_1 - t_1)}$$

(Equation 1.5)

R is equal to the shell side fluid flow rate times the fluid mean specific heat; divided by the tube side fluid flow rate times the tube side fluid specific heat. S is a measure of the temperature efficiency of the exchanger. For a 1 shell: 2 tube pass exchanger, the correction factor is given by Equation 1.6:

$$F_t = \frac{\sqrt{(R^2+1)}\ln\left[\frac{(1-S)}{(1-RS)}\right]}{(R-1)\ln\left[\frac{2-S[R+1-\sqrt{(R^2+1)}]}{2-S[R+1+\sqrt{(R^2+1)}]}\right]}$$

(Equation 1.6)

The derivation of Equation is given by Kern (1950). The equation for a 1 shell: 2 tube pass exchanger can be used for any exchanger with an even number of tube passes, and is plotted in Figure 1.9.

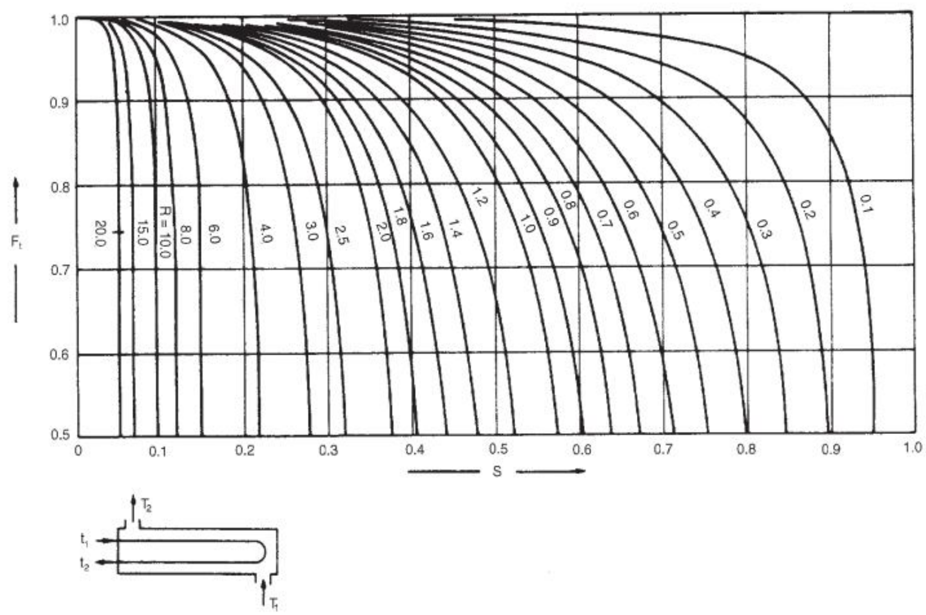

Figure 1.9: Temperature Correction Factor: One Shell Pass; Two or More Even Tube Passes

Temperature correction factor plots for other arrangement can be found in the TEMA standards and the books by Kern (1950) and Ludwig (1965). Mueller (1973) gives a comprehensive set of figures for calculating the log mean temperature correction factor, which includes figures for cross flow exchangers.

The following assumptions are made in the derivation of the temperature correction factor F_t, in addition to those made for the calculation of the log mean temperature difference:

1. Equal heat transfer areas in each pass
2. A constant overall heat transfer coefficient in each pass
3. The temperature of the shell side fluid in any pass is constant across any cross section
4. There is no leakage of fluid between shell passes

Though these conditions will not be strictly satisfied in practical heat exchanger, the F_T values obtained from the curves will give an estimate of the “true mean temperature difference” that is sufficiently accurate for most designs. Mueller (1973) discusses these assumptions, and gives F_t curves for conditions when all the assumptions are not met; see also Butterworth (1973) and Emerson (1973).

The shell side leakage and bypass streams will affect the mean temperature difference, but are not normally taken into account when estimating the correction factor F_t. Fisher and Parker (1969) gives curves which show the effect of leakage on the correction factor for a 1 shell pass: 2 tubes pass exchanger.

The value of F_t will be close to one when the terminal temperature differences are large, but will appreciably reduce the logarithmic mean temperature difference when the temperature of shell and tube fluids approach each other; it will fall drastically when there is a temperature cross. A temperature cross will occur if the outlet temperature of the cold stream is greater than the inlet temperature of the hot stream

Where the F_t curve is near vertical values cannot be read accurately, and this will introduce a considerable uncertainty into the design.

An economic exchanger design cannot normally be achieved if the correction factor F_t falls below about 0.75. In these circumstances an alternative type of exchanger should be considered which gives a closer approach to true counter-current flow. The use of two or more shells in series, or multiple shell-side passes, will give a closer approach to true counter-current flow, and should be considered where a temperature cross is likely to occur.

Where both sensible and latent heat is transferred, it will be necessary to divide the temperature profile into sections and calculate the mean temperatures difference for each section.

1.6 Problem Statement and Objectives

When the process reaches maximum production rates without satisfying the market demand, it is referred to as being “bottleneck”. Hence, it will result in static profit because of this maximum throughput. Throughput is an output or product that produced after the input passing through all the process and reactions involved. When there is bottlenecked condition in LNG plant, debottlenecking process should be implemented to identify and remove the obstacles that limit the production. On the other hand, this process plays an important role in yield enhancement, especially, for enhancing the company’s revenue and profits.

In line with the statements of the problems, this work involves two objectives. The first objective is to debottleneck the small scale LNG plant in order to increase the throughput or production capacity. In this work, a flow sheet of liquefaction process for LNG production with a complete available process data by H.A.Razik (2007) has been selected as a case study. Then, bottleneck condition was identified by simulated this flow sheet in the Aspen HYSYS software. After bottleneck condition have been validated by this process simulation tool, modification then should be done to satisfy the second objective which is to design the schemes for the debottlenecking process in order to increase the production. Typically, the designs of the required schemes need a details analysis of the current plant performance. By taking the current plant performance as a reference, designing of the modifications was implemented regarding to the desire to prolong or avoid the bottlenecks situation. This further process improvement is analyzed by doing an economic evaluation for every modification after debottlenecking process has been implemented to see their performance compared to the current plant.

CHAPTER 2

LITERATURE REVIEW

H.A. Razik (2007) reconciled and validated of Mixed Refrigeration Cycle process for LNG production. Process data of a single mixed refrigerant cycle flow sheet for the production of liquefied natural gas (LNG) has been reconciled to establish complete mass and energy balance from the previous work (Cao, et. al., 2006). The reconciliation approach involved the combination of structural decomposition of complex units and systematic iterative estimations of the unknown process data. Complex LNG exchanger units were decomposed into a series of simple heat exchanger under strict energy balance constraints. Estimation of unknown process data were achieved through simulations and validated against a number of reported process performance data. The best reconciled flow sheet was based on the smallest of Sum of Square Error (SSE).

Then, in this research, the reconciled and validated flow sheet process data of this single mixed refrigerant cycle will be regarded as a base case. Base case flow sheet simulated in the Aspen HYSYS software with the use of Peng-Robinson and Lee-Kesler-Plocker as an equation of states.

2.1 Design of Increasing Production Capacity

Process Engineering Associates, LLC (PROCESS) (2009) implemented a preliminary study of the crude distillation unit for potential revamp opportunities and then provided the process design for the identified revamp items. The crude distillation system consists of an 11-exchanger pre-heat train, desalter, pre-fractionation tower, two fired heaters, atmospheric crude distillation tower, and vacuum distillation tower. Project tasks included the process simulation of the overall crude distillation unit: current configuration and proposed modifications, rigorous simulation of the crude preheat train exchangers, modeling of hydraulics and heat transfer for the trays and packing and hydraulic calculations for piping circuits, control valves, and pumping requirements. As a result of this project, the preheat train was modified to achieve a significant reduction in pressure drop while achieving much greater heat recovery with minimum exchanger modifications. The desalter operating temperature was increased to handle heavier crude slates. Piping and control valve size modifications greatly improved unit hydraulics. The existing atmospheric and vacuum distillation columns were evaluated and determined to have sufficient capacities for the increased rates. Furnace firing increases were minimal or neutral because of the improved modified unit heat recovery and improved pump-around capacity. The resulting unit modifications resulted in a 20% capacity increase of the crude distillation unit with minimal operating cost increase.

Tata Chemicals Ltd. (TCL) (2009), a leading manufacturer of chemicals, fertilizers and food additives, successfully implemented the debottlenecking process of their facility to boost capacity for both ammonia and urea production. The project was completed in a record time of 18 months with the aim to fine tuning and enhancing production capabilities thereby expanding the existing production levels. The increased growth of production draws impetus to the company commitment of providing high quality products for its discerning customers. With this the production capability of the fertilizer increases from 8, 64,600 million tonnes (mt) to 11, 55,000 mt per year. Post the debottlenecking process, the new production capacity at the plant result the ammonia plant produced 2000 million tonnes per day (mtpd) as compared to 1520 mtpd and urea plant will produce 3500 mtpd as compared to 2620 mtpd.

Fertil (2009) is planning an expansion of its existing urea production unit through a debottlenecking process. The plant is currently producing 1,830 million tonnes per day of urea with surplus ammonia of 90,000 million tonnes per year, which Fertil intends to convert into urea and deliver the feedstock to the proposed Melamine plant coming up in Ruwais. The scopes of work involve the installation of a 400 million tonnes per day of carbon dioxide recovery unit, modification of the existing urea plant to convert surplus ammonia, installation of new 2500 million tonnes per day of granulation unit. It also includes installation of all required utilities like power, cooling water, sea water supply, steam, potable water and firefighting system. On the other hand, the scope of work also involved the supply of utilities to the new melamine plant and demolition / removal of equipment mounted on and around the prilling tower including conveyors up to the bulk storage facilities, the screen house and the final cooler building.

Abdelkader Haouari (2005) has proposed a systematic approach for successfully increase production. An LNG plant was originally designed to produce 6 MTPA of LNG. With the aim of increasing production, an optimization study was initiated. Drawing on his experience, the author identified minor modifications to the plant and changes to operations and maintenance practices which resulted in 17% increase in LNG production (7.0 MTPA). On the other hand, by implementing the sulphur recovery expansion project which improved sulphur recovery to more than 97%, reduced sulphur emission and allowed LNG production to increase by 10% (7.7 MTPA) and to further increase the production capacity of the plant to 9.5 MTPA, the debottlenecking project was implemented, which involved various modifications in all the three trains over a three year period during planned major overhauls. Finally, the author initiated an advanced process control project which optimized LNG production and plant efficiency as well as offloading console operator by automation. This project further increased LNG production by 1.5%.

Chen *et al* (2005) have proposed a process for debottlenecking a system for the separation of a conjugated diolefin the system including a first extraction section having an extractive distillation column, a stripping column and a second extraction section. The process includes the steps of withdrawing a first portion of an extract from the extractive distillation column, the extract having at least the first portion and a second portion, and transferring the first portion of the extract to a flash or separation vessel; separating the first portion of the extract into a vapor phase and a liquid phase by flashing in a flash or separation vessel; and combining the liquid phase of the separated first portion of the extract with the second portion of the extract to produce an extract feed for further processing. A system and process for the separation of a conjugated diolefin from a C4- or C5-hydrocarbon mixture containing the conjugated diolefin and higher acetylenes are also provided.

MLNG Dua is being debottlenecked to meet a higher export demand of LNG. Increased production is achieved by the up rating of power available to the main refrigeration C3 and MCR systems) and the addition of a new (extended) end flash unit. Existing process and utilities systems are checked for new operation and modified as required, including equipment, valves, relief valves, piping, etc. Where possible, replacement of existing equipment is avoided and changes are limited to simple modifications. Additional equipment is added to existing units only where absolutely necessary.

DSM's special products division is to continue a process of debottlenecking at its benzaldehyde facilities in Rotterdam and Geleen to add another 10 000 tonne of capacity by stages during 1996. This adds to a 3000 tonne increase this year and results from increasing demand for benzyl alcohol, particularly for coatings applications. The company also plans to up the capacity of its benzoic acid and sodium benzoate acid facilities, by some 4000 tonne apiece, to come on-stream in the last of 1996.

Australian Gold Reagents Pty Ltd (1988) recommended two major techniques to expand existing sodium cyanide solution plant and increase its annual production from 15,000 tonnes per annum to 40,000 tonnes per annum. This would be achieved by doubling the number of reactors and by removing engineering obstacles to utilize the major items of plant at their optimum capacity, otherwise known as debottlenecking. This would permit an increase in the flow rates of the feed gases (natural gas and ammonia), product and waste gas streams through the plant. The storage capacity would also be increased to 8833 tonnes of sodium cyanide solution.

Jully and Dominic (2006) have has done a meaningful work in increasing the throughput. With the field of batch pharmaceutical cream production, they categorized two types of bottlenecks. These are then equipment capacity-related size bottleneck equipment that is limited in size as well as the scheduling bottleneck (due to the long occupancy of a piece of equipment size). The ability to identify and remove process bottlenecks that create obstacles in a manufacturing process will increase plant throughput and fulfill customer needs. A good tool to identify batch process is via a throughput analysis study by measure the equipment utilization and equipment uptime. As a result, the annual process throughput is increased significantly with the reduction of equipment uptime of the process time bottleneck by using SuperPro Designer.

From these previous works, it should be concluded that by doing a debottlenecking process whether to remove the bottlenecks equipment, placing the new unit operation or modify the current unit operation, the plant performance will increased and the demand of certain production will be satisfied. Besides that, this work is meaningful for the further analysis in line with the valuable discovery of performance of small scale LNG processing plant.

CHAPTER 3

METHODOLOGY

3.1 Model Development

Figure 3.1 in the Appendix C showed the base case simulation flow sheet for the production of Liquefied Natural Gas (LNG) modeled in Aspen HYSYS simulator. Due to the nature of the process that is operated in steady state mode, the base case simulation model was developed to reflect the actual operating condition of the current production in the existing small scale LNG plant manufacturing facility.

In the base case process, there are two major processing steps which includes the compression with inter cooling (due to decreasing the irreversible degree and the power consumption of compression process for Mixed Refrigerant Cycle) and the production of the LNG.

3.2 Analyze Existing Production Capacity and Performance of the Small Scale LNG Plant

For this stage, the results and process data of the flow sheet by H.A. Razik (2007) are tabulated in Table 3.1 which summarizes the simulation results including the power consumption of compressor, load of water cooling, liquefaction rate and power per unit LNG produced. All of these values are useful for results validation as well as error detection. The flow rate of the natural gas and the mixed refrigerant are fixed to be 4.00 kg mol/h and 60.25 kg mol/h r respectively since the liquefaction process operates in a steady-state mode.

Table 3.1: Summary of Simulation Results for Figure 3.1

Parameter	Value
Power consumption of compressors (kW)	129.51
Load of water-cooling (kW)	146.50
Liquefaction rate	0.952
Power per unit LNG (kW/mol/s)	122.47

3.3 Identification of Bottleneck Unit Operation

3.3.1 Process Debottlenecking of Small Scale LNG Plant by Hierarchical and Heuristic Techniques

Hierarchical and Heuristic techniques have been used for this debottlenecking process. According to Musaed, Nasser and Mahmood, September, 2007, this approach relies on intuition engineering knowledge and physical principles to decompose the problem into sequential stages. A hierarchical approach was introduced in 1985 that proposed a method for screening alternatives and modifying equipment sizes, replacing units and adding new equipment. A heuristic approach was developed in 1993 that includes procedures for equipment design, capital costs and economic evaluation. This was applied to the addition of a new unit into an aromatic plant. While heuristic approaches utilize engineering insights, they are not identifying optimum solutions for general cases.

3.3.2 Process Debottlenecking of Small Scale LNG Plant by Simulation Tool Techniques

In order to increase the process throughout, the process bottleneck that limits the current production need to be identified. Bottlenecks are process limitations that are related to either equipment or resources such as demand for various utilities, labor, raw material, etc. Hence, process debottlenecking can be defined as the identification and removal of obstacles in the attempt to increase the plant throughput (Koulouris, et al, 2000). The ability to identify and remove process bottlenecks that create obstacles in a manufacturing process will increase plant throughput and fulfill customer needs. In this work, two types' process bottlenecks can be categorized. These are capacity utilization (related to bottleneck production) and the power consumption of related equipment (due to Mixed Refrigerant Cycle process).

A good tool to identify the active bottlenecks is via a throughput analysis study. Throughput analysis measures the equipment utilization in the processing plant with two variables, the capacity utilization and equipment effectiveness. Capacity utilization is defined as the percentage of the current operating load of equipment relative to its maximum load. On the other hand, equipment effectiveness measures the performance of a piece of equipment in line with the power consumption.

Simulation tools that are capable of tracking capacity utilization and equipment effectiveness can facilitate the identification of process bottlenecks and the development of the scenario for process debottlenecking. By using the “what if” scenario, process alternatives can be simulated via the use of simulation tools to reveal potential for the debottlenecking study.

3.4 Process Debottlenecking Schemes Design Using Aspen HYSYS Software

HYSYS contains a multi variable steady state Optimizer. Once the flow sheet has been built and a converged solution has been obtained, the Optimizer can be used to find the operating conditions which minimize (or maximize) the objective function. The Optimizer owns its own spreadsheet for defining the objective function, as well as any constraints expressions to be used.

Modifications in terms of production increment schemes design are applied to the base of MRC liquefaction process flow sheet. The main objective of such modifications is intently to avoid the temperature cross in the bottleneck unit operation in conjunction to increase the production of the LNG. Therefore, several schemes can be incorporated to the base MRC flow sheet with the aim to attain additional heat recovery to ensure that no temperatures cross occurs in any units by place a stage heat exchanger in series.

3.5 Final Evaluation with Economic Analysis

In this step, after the debottlenecking process has been implemented, the performance of the existing plant and the plant after modifications has been takes places are compared. Finally, we will gain production revenue from these plants which based on the current prices of the LNG with the some consideration of the changing parameters and cost of additional equipment.

3.6 Summary of Methodology

Figure 3.2 below summarizes procedures for analysis of MRC process and designing modification schemes for process debottlenecking of small scale LNG plant.

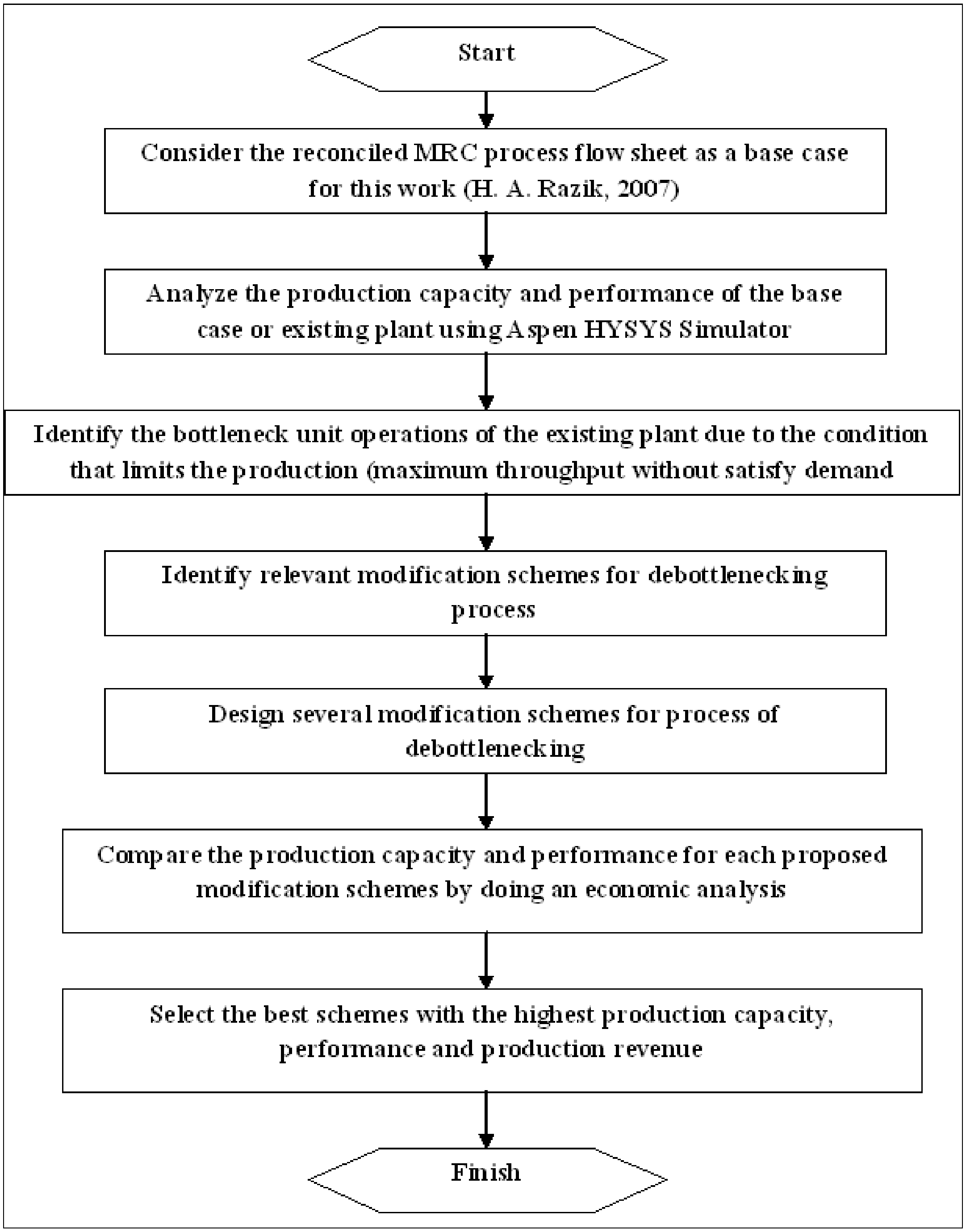

Figure 3.2: Procedures for Analysis of MRC Process and Designing Modification Schemes for Process Debottlenecking of Small Scale LNG Plant

CHAPTER 4

RESULTS AND DISCUSSION

4.1 Existing Plant (Base Case)

From the base case simulation, a complete LNG plant has a feed flow rate of NG which is 4 kg mole/hr, the process cooling duty of 1518500 kJ/hr, power consumption of compressors which is 131.23 kilowatt and 3.866 kg mole/hr of LNG production.

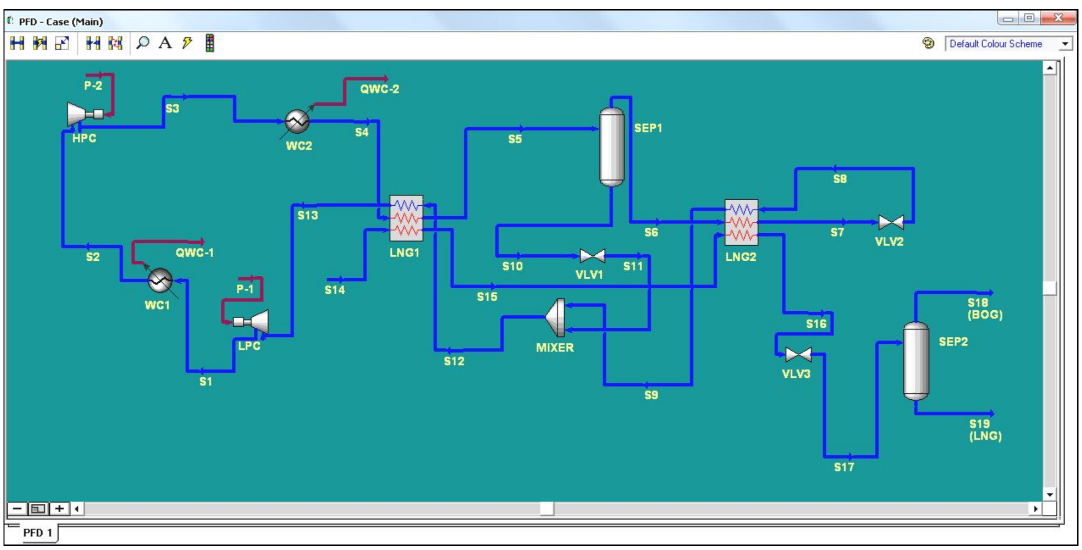

Figure 4.1: Base Case Simulation (Existing Plant)

By doing a debottlenecking process, it is validated that, this base case can achieve a higher LNG production with the constant value of power consumption. The highest LNG production that can be achieved for this existing plant is 3.938 kg mole/ hr in line with the maximum flow rate of NG of 4.074 kg mole/hr. From the observation, it has been recognized that, LNG 2 heat exchanger is the active bottleneck for this LNG processing plant. This is because when the feed flow rate is increase up to 4.075 kg mole/hr, the LNG 2 heat exchanger will face a temperature cross condition.

Temperature cross condition occur because of limitations of multi pass arrangements. Since the 1-2 heat exchanger uses one parallel pass and one counter current, it follows that the maximum heat recovery for these units should be between that of parallel and counter flow. As a practical limit, it is important that nowhere in the unit should the cold fluid temperature exceed that of the hot fluid. If so, then the heat transfer is obviously in the wrong direction. Such a situation can arise in a multi pass heat exchanger as seen in Figure 1.9. This unit represents a cold fluid, located on the tube side of the heat exchanger, making two passes through the unit, the hot fluid, on the shell side, travelling across the unit only once. Here the cold fluid is heated to a temperature slightly above that of the hot fluid near the exit for the two streams. At this axial location, near the left end of the unit, the temperature of the cold fluid in the first pass remains well below that of the hot fluid so that considerable heat transfer occurs. The cold fluid in the second pass is slightly above that of the hot fluid at the same location. The small temperature difference between the second pass cold fluid and the hot stream indicates that only a small amount of heat will be transferred between these streams. Overall heat will flow from hot to cold fluid, but a portion of the heat transfer surface is being used in a counter productive way. This condition is called temperature cross. In the unit the hot fluid exit temperature could be cooled to the average of the cold fluid inlet and exit temperature. This would, however, be highly inefficient and would require an excessively large surface area. Some engineers advocate that good design should not permit a temperature cross, indicating that the 1-2 should operate with the same heat recovery limit as a true parallel flow.

The preferred method of attaining additional heat recovery is to stage heat exchanger in series so that no temperatures cross in any unit. An equivalent solution is to put multiple 1-n arrangements within a single shell. A 2-4 unit is the equivalent of 2 1-2 units provided that the total heat transfer area is equal. Similarly a 3-6 unit is the equivalent of 3 1-2 units with equal overall area.

Other engineers suggest that a small temperatures cross may be acceptable and may provide a less expensive design than the more complex alternatives. If one were to plot the locus of points where the temperature cross occurs for the 1-2 heat exchanger on the temperature correction chart, it would be found to correspond to a relatively narrow range F_t values ranging from about 0.78 to 0.82. Lower values of F_t may be taken as an indication that a temperature cross will occur. An economic exchanger design cannot normally be achieved if the correction factor F_t falls below about 0.75. In these circumstances an alternative type of exchanger should be considered which gives a closer approach to true counter current flow. The use of two or more shells in series, or multiple shell side passes, will give a closer approach to true counter current flow, and should be considered where a temperature cross is likely to occur. Where both sensible and latent heat is transferred, it will be necessary to divide the temperature profile into sections and calculate the mean temperature difference for each section.

Because of the active bottlenecks occur at LNG 2 heat exchanger, then, figures below showed us the parameter of LNG 2 heat exchanger for base case (existing plant) that will be considered in order to analyze the performance of this bottlenecks equipment.

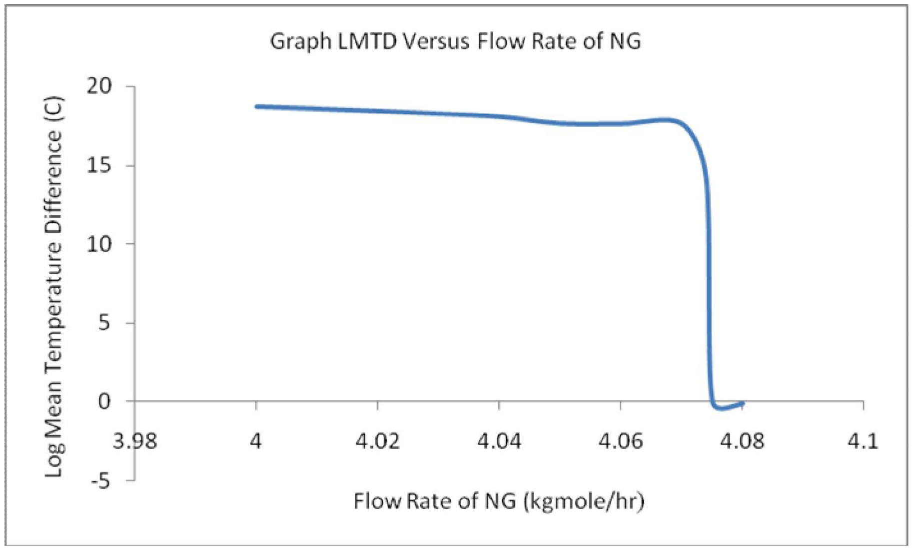

Figure 4.2(a): The Effect on Log Mean Temperature Difference to the Increment of NG Flow Rate

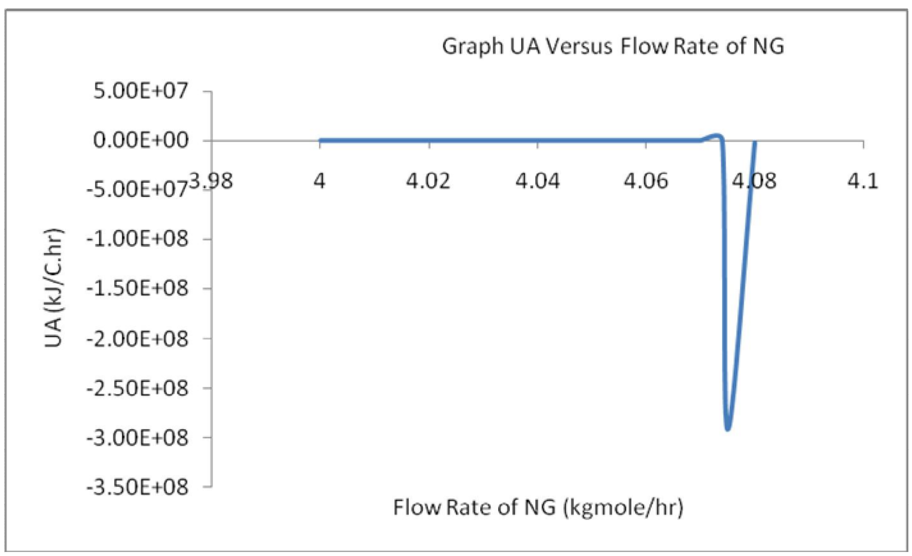

Figure 4.2(b): The Effect on UA to the Increment of NG Flow Rate

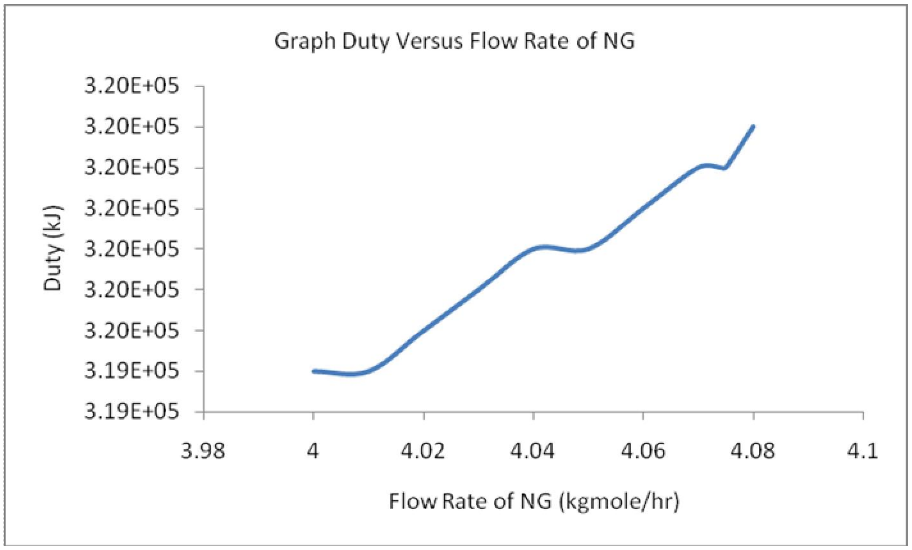

Figure 4.2(c): The Effect on Duty to the Increment of NG Flow Rate

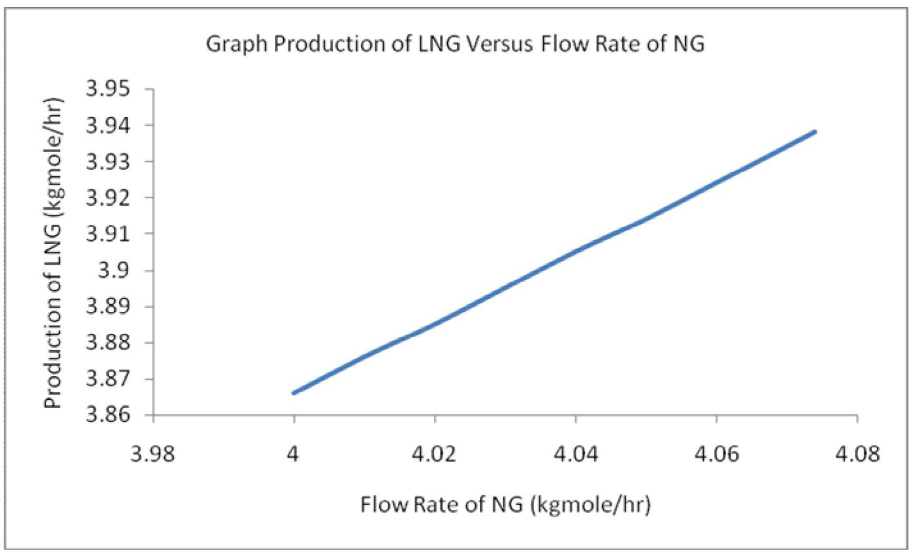

Figure 4.2 (d): The Effect on Production of LNG to the Increment of NG Flow Rate

HYSYS is capable of calculating Q/DT_{avg} (which they designate as "UA", even though it is actually UAF. They can also provide a plot of T for both streams versus the amount of heat transferred. Unfortunately, the academic portion of HYSYS in Weighted or End point Engineering design give a heat transfer area, A of 60.32 m^2 for all heat exchanger, regardless of what is specified. It calculates U by dividing UA by this value. Such U and A are meaningless and therefore, must not be used. Then, because of this limitation will result in difficulty to estimate the costing of the heat exchanger in economic evaluation, a few scheme have been develop to overcome this issues.

4.2 Modification 1 (Addition of Cooler at Stream 6)

Modification 1 focuses on the addition of the cooler at stream 6. According to R.A. Crane (2004), the preferred method of attaining additional heat recovery to ensure that no temperatures cross occurs in any units is to stage heat exchanger in series. Then, by using this approach, modification 1 is simulated into Aspen HYSYS simulator and the performance of new plant is analyzed.

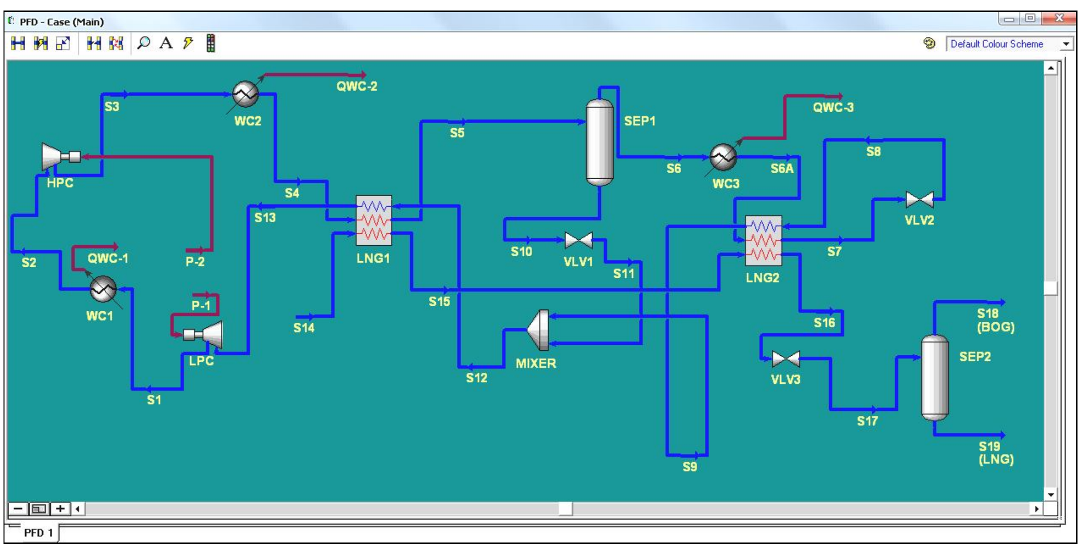

Figure 4.3: Addition of Cooler at Stream 6 (Modification 1)

Simulation result showed that the production of LNG decrease to 3.934 kg mole/hr by increasing the feed flow rate of NG by 0.066 kg mole/hr from the actual feed flow rate. The temperature cross condition occur when the feed flow rate of the NG is 4.067 kg mole/hr. On the other hand, the power consumption of compressors is 131.23 kilowatt and process cooling duty is decreased to 1518045.64 kJ/hr. Figures below showed the effects of the increment of NG flow rate to parameters of the LNG 2 heat exchanger.

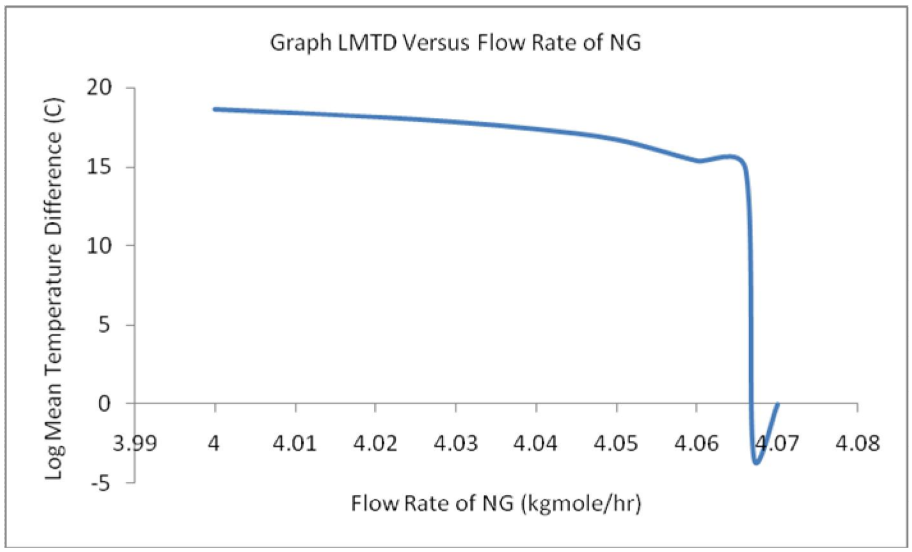

Figure 4.4(a): The Effect on Log Mean Temperature Difference to the Increment of NG Flow Rate

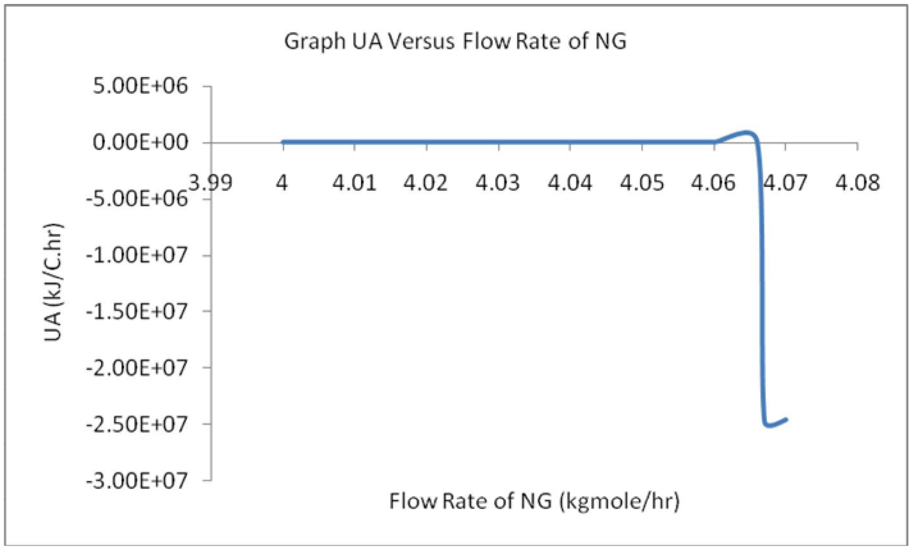

Figure 4.4(b): The Effect on UA to the Increment of NG Flow Rate

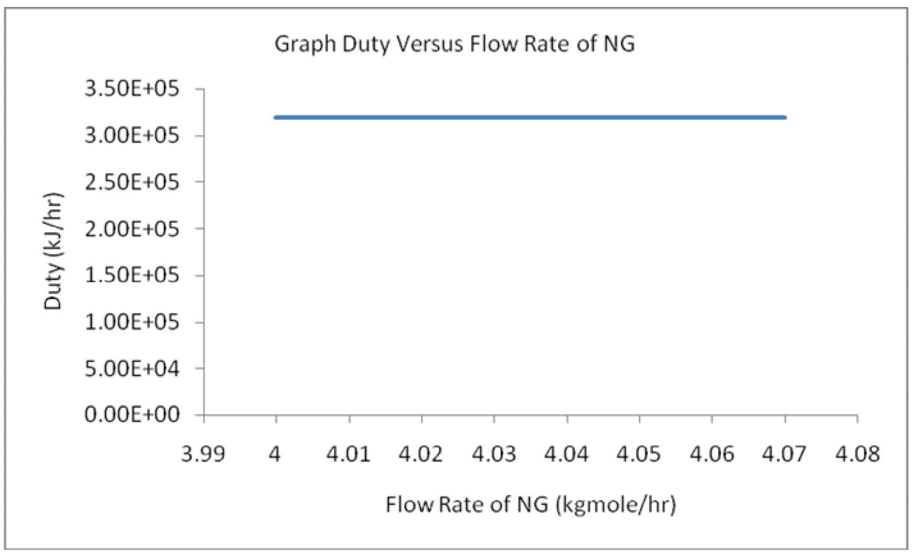

Figure 4.4(c): The Effect on Duty to the Increment of NG Flow Rate

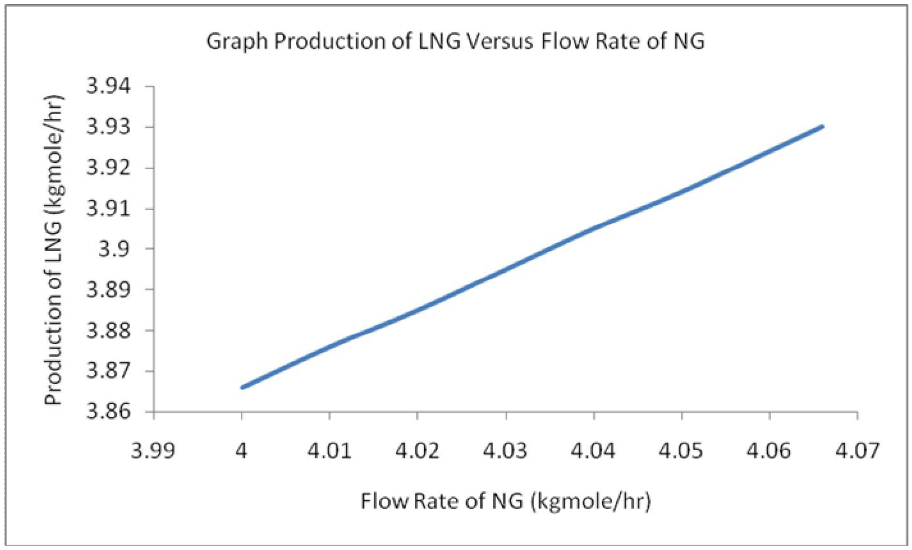

Figure 4.4(d): The Effect on Production of LNG to the Increment of NG Flow Rate

4.3 Modification 2 (Addition of the Cooler at Stream 7)

By use the same approach as before, modification 2 plant performances is analyzed to study the effect of adding cooler at the stream 7.

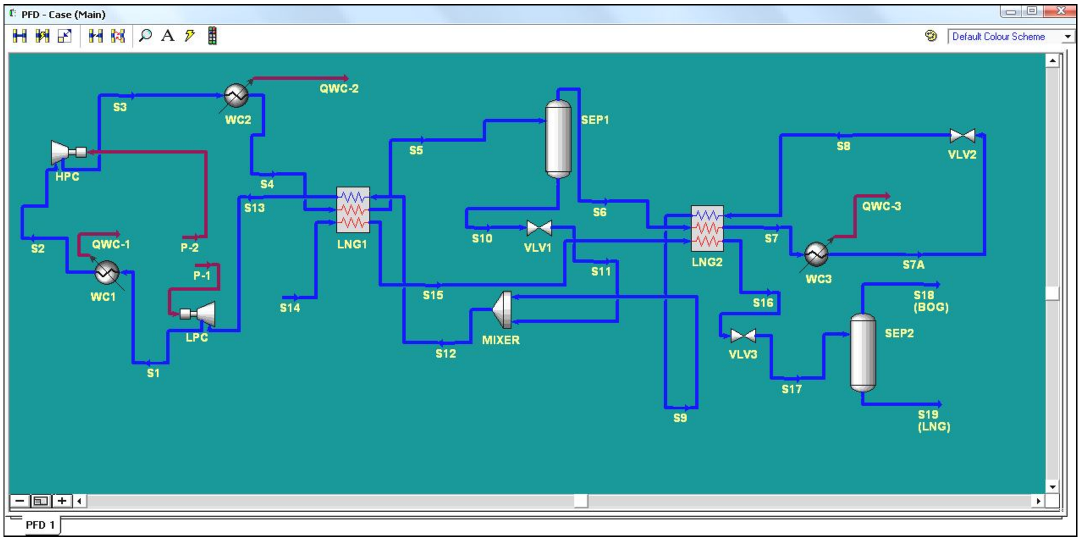

Figure 4.5: Addition of Cooler at Stream 7 (Modification 2)

Simulation result showed that the production of LNG for modification 2 is increase up to 3.944 kg mole/hr compared to the 3.938 kg mole/hr of the base simulation case. The temperature cross condition only occur when the feed flow rate is increase over 4.081 kg mole/hr. The production of the LNG is increase by 0.15 % from the actual production which is base case simulation. On the other hand, the power consumption of the compressors is 131.23 kilowatt and the process cooling duty is 1521614.30 kJ/hr. Figures below showed the effects of the increment of NG flow rate to parameters of the LNG 2 heat exchanger.

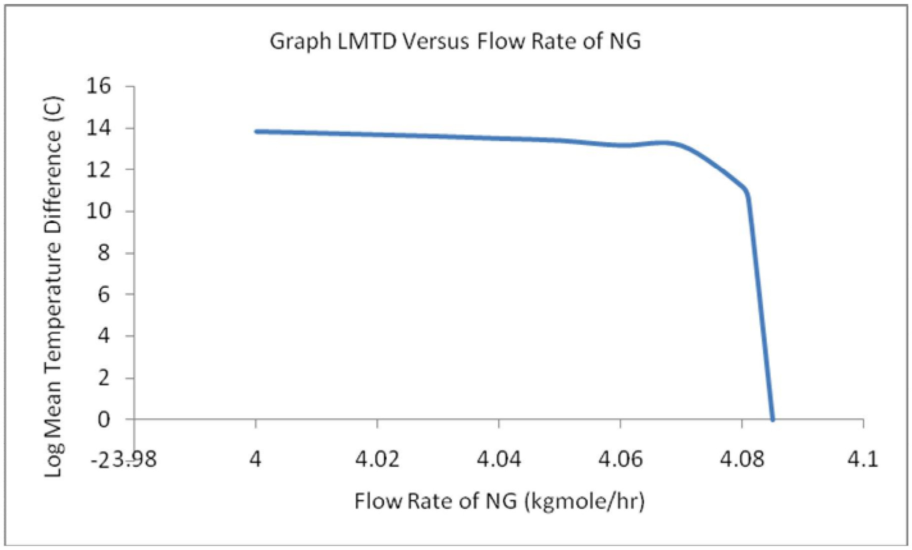

Figure 4.6(a): The Effect on Log Mean Temperature Difference to the Increment of NG Flow Rate

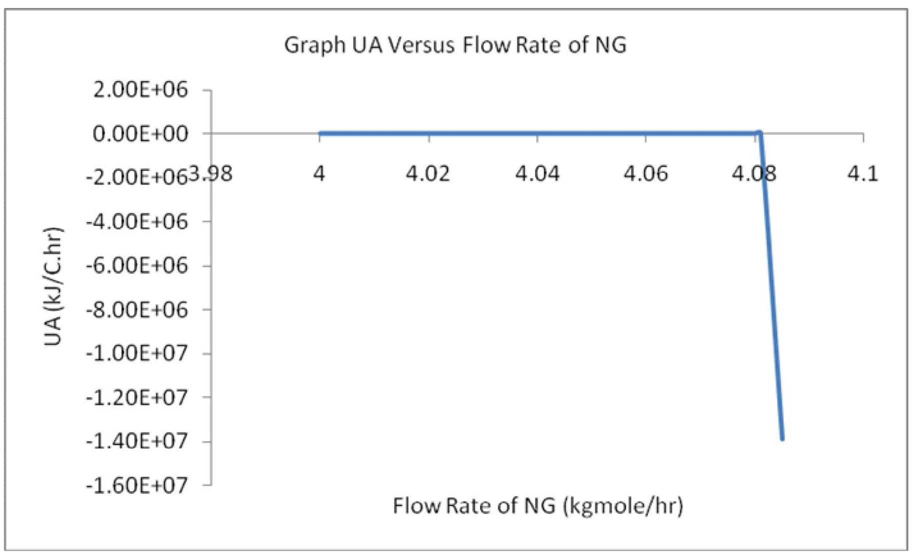

Figure 4.6(b): The Effect on UA to the Increment of NG Flow Rate

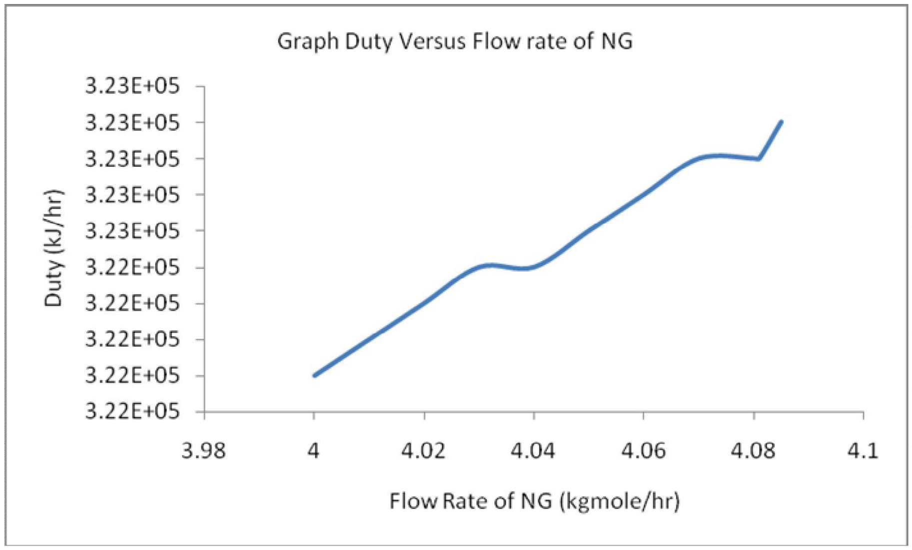

Figure 4.6(c): The Effect on Duty to the Increment of NG Flow Rate

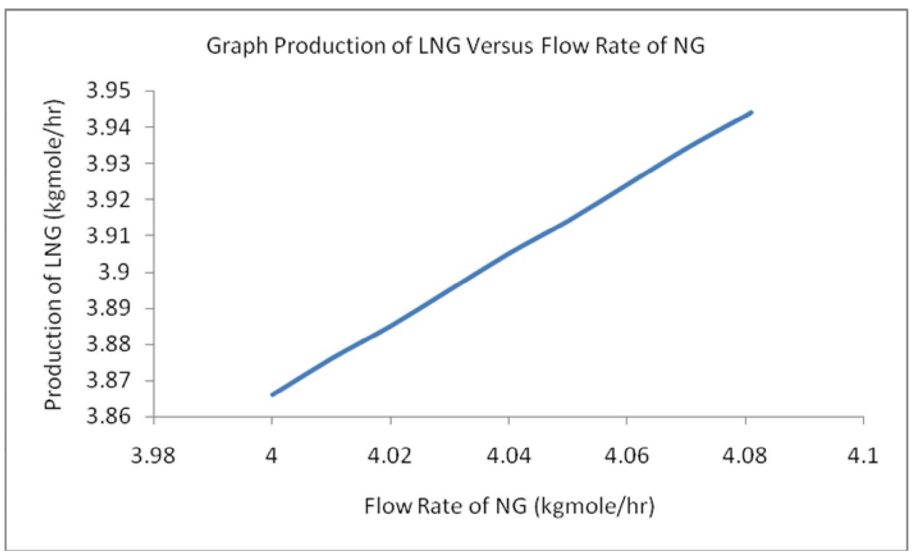

Figure 4.6(d): The Effect on Production of LNG to the Increment of NG Flow Rate

4.4 Modification 3 (Addition of Cooler at Stream 8)

Modification 3 also developed by using the same approach with the others modification before. The plant performance is analyzed to study the effect of adding cooler at the stream 8.

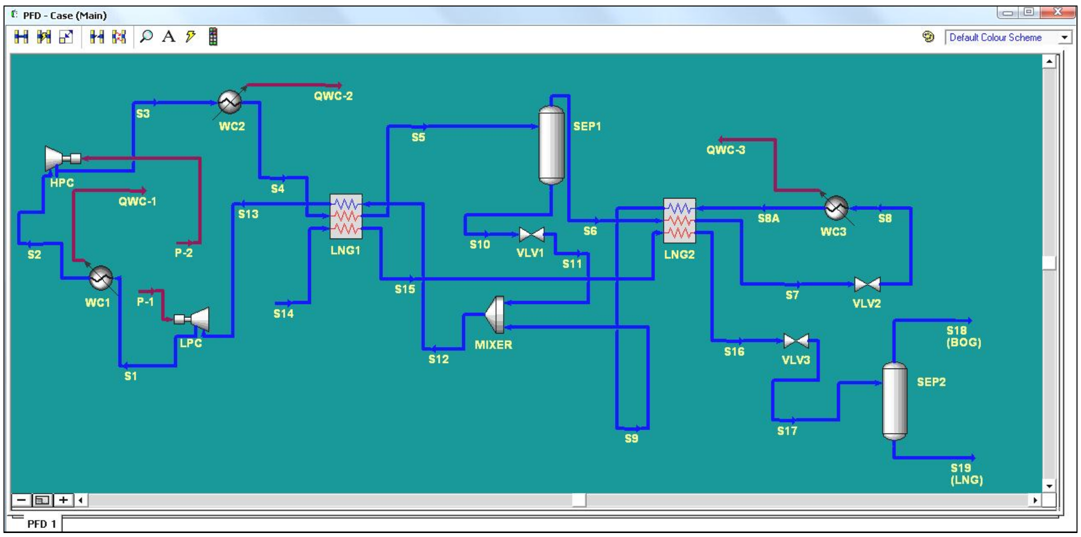

Figure 4.7: Addition of Cooler at Stream 8 (Modification 3)

Simulation result showed that the production of LNG is increase up to 11.690 kg mole/hr compared to the 3.938 kg mole/hr of the base simulation case. The temperature cross condition only occur when the feed flow rate is increase over 12.100 kg mole/hr. The production of the LNG is increase by 196.9% from the actual production which is base case simulation. On the other hand, the power consumption of the compressors is 131.230 kilowatt and the process cooling duty of 1760400 kJ/hr. Figures below showed the effects of the increment of NG flow rate to parameters of the LNG 2 heat exchanger.

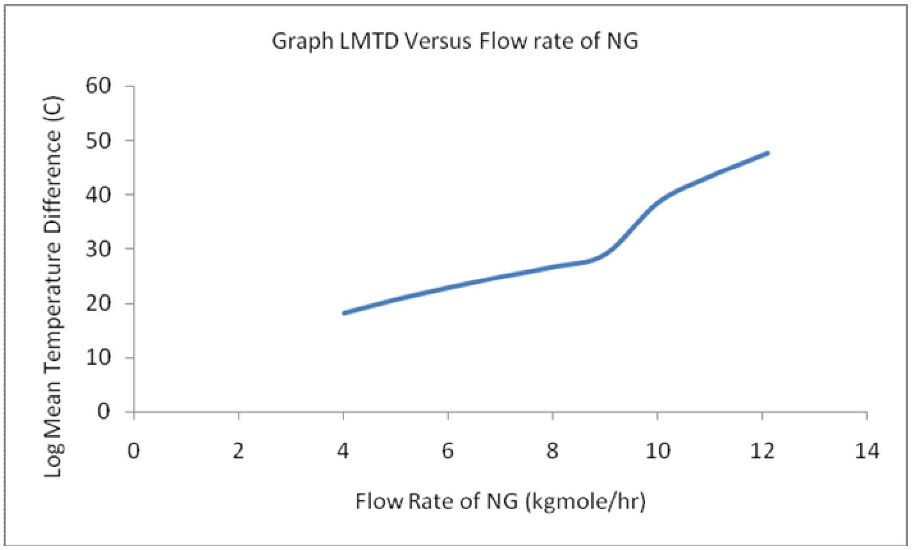

Figure 4.8(a): The Effect on Log Mean Temperature Difference to the Increment of NG Flow Rate

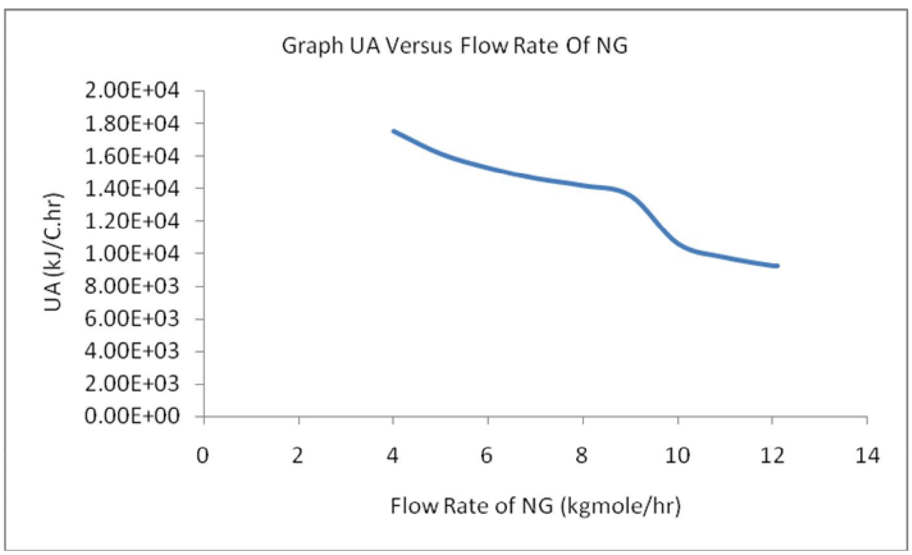

Figure 4.8(b): The Effect on UA to the Increment of NG Flow Rate

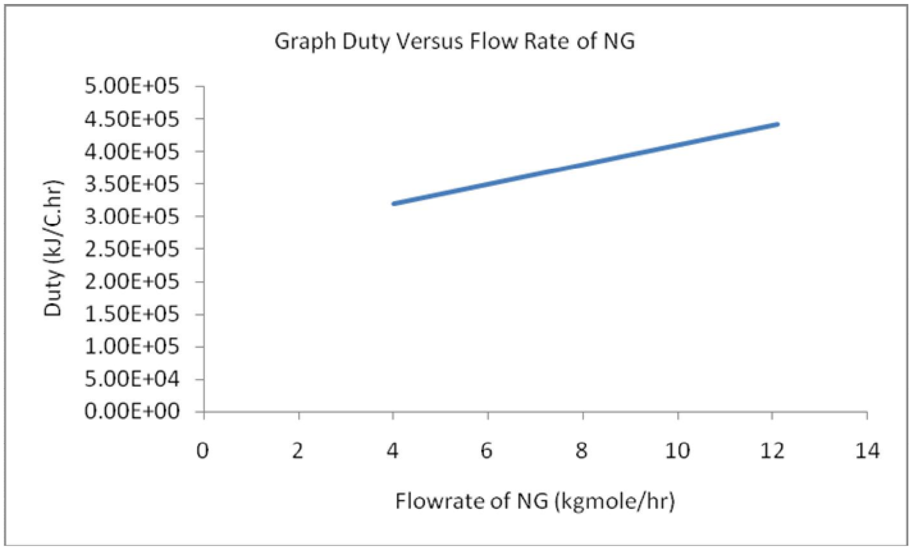

Figure 4.8(c): The Effect on Duty to the Increment of NG Flow Rate

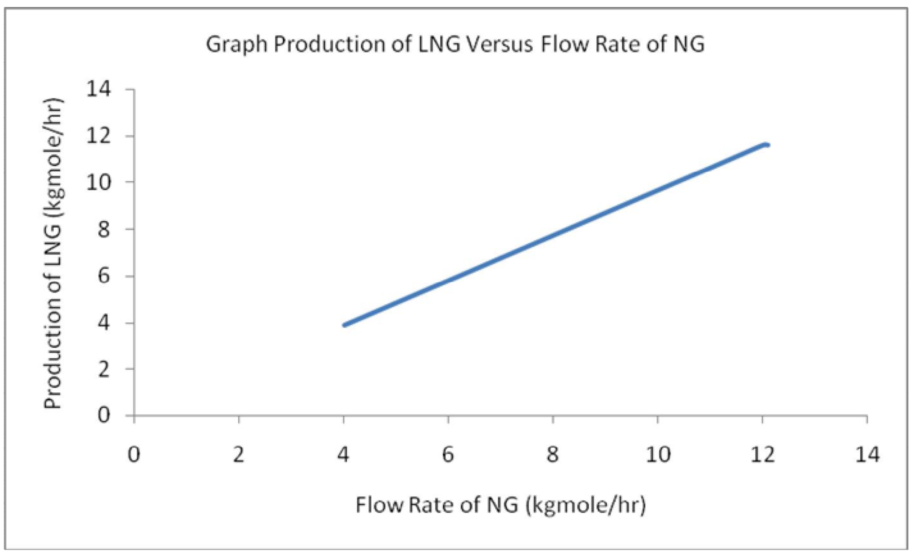

Figure 4.8(d): The Effect on Production of LNG to the Increment of NG Flow Rate

4.5 Modification 4 (Addition of Cooler at Stream 9)

For modification 4, the plant performance is analyzed to study the effect of adding cooler at the stream 9.

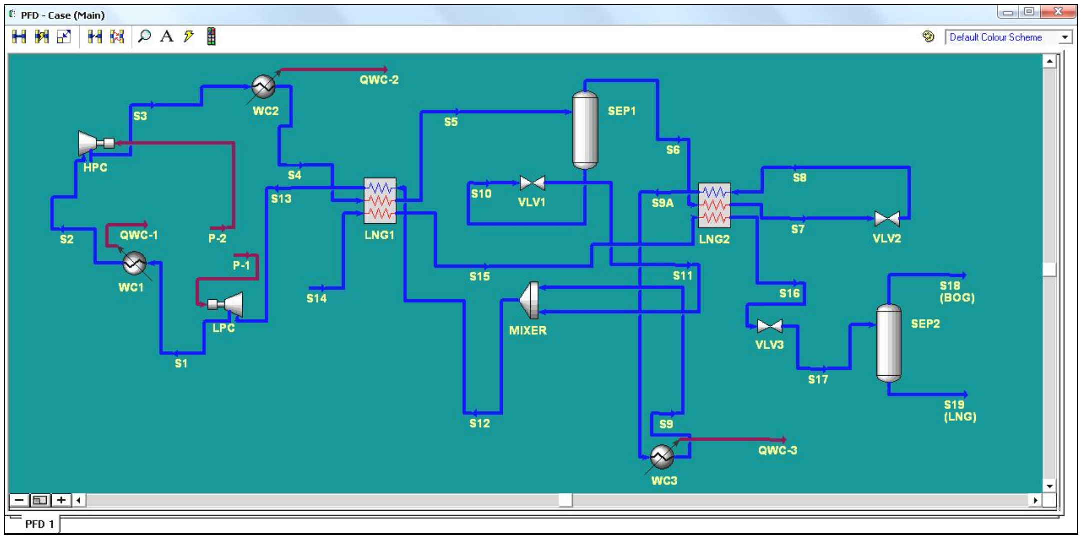

Figure 4.9: Addition of Cooler at Stream 9 (Modification 4)

Simulation result showed that the production of LNG is increase up to 3.943 kg mole/hr compared to the 3.938 kg mole/hr of the base simulation case. The temperature cross condition only occur when the feed flow rate is increase over 4.080 kg mole/hr. The production of the LNG is increase by 0.127 % from the actual production which is base case simulation. On the other hand, the power consumption of the compressors is 131.23 kilowatt and the process cooling duty is 1519008 kJ/hr. Figures below showed the effects of the increment of NG flow rate to parameters of the LNG 2 heat exchanger.

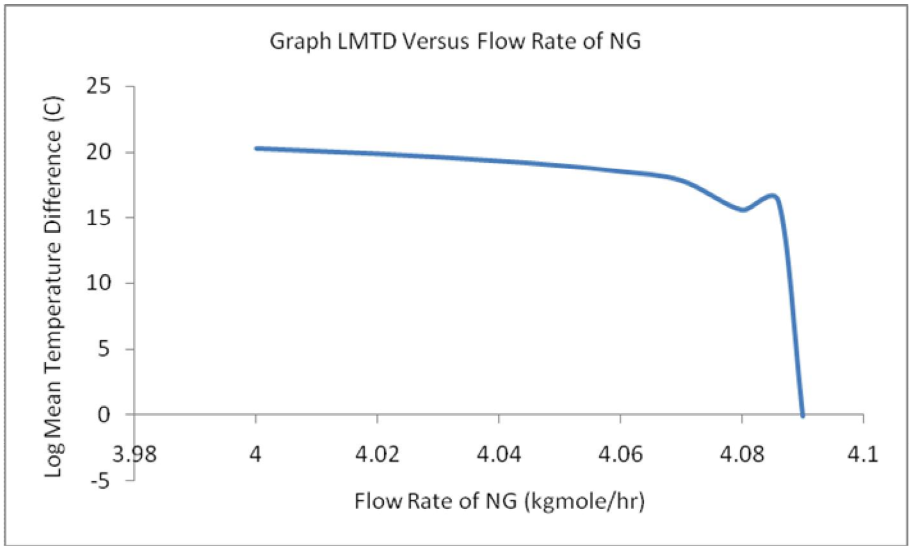

Figure 4.10(a): The Effect on Log Mean Temperature Difference to the Increment of NG Flow Rate

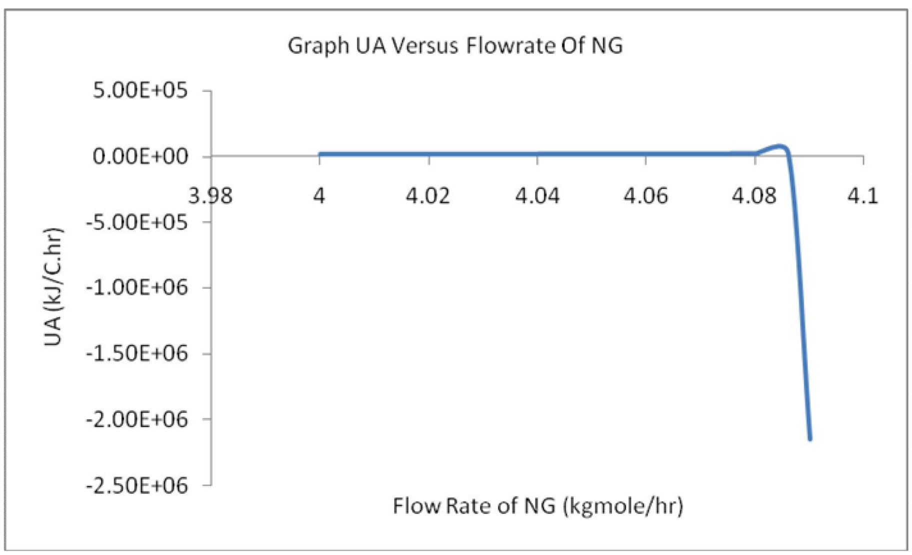

Figure 4.10(b): The Effect on UA to the Increment of NG Flow Rate

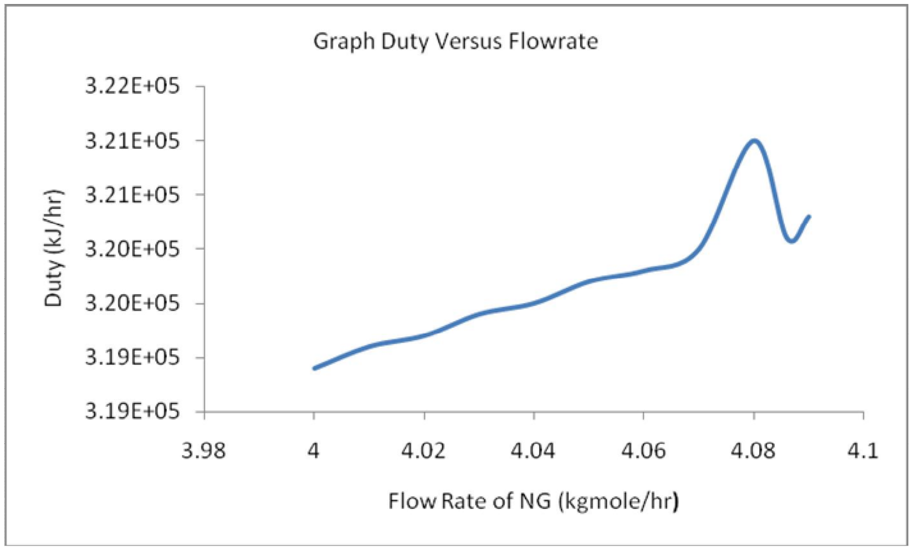

Figure 4.10(c): The Effect on Duty to the Increment of NG Flow Rate

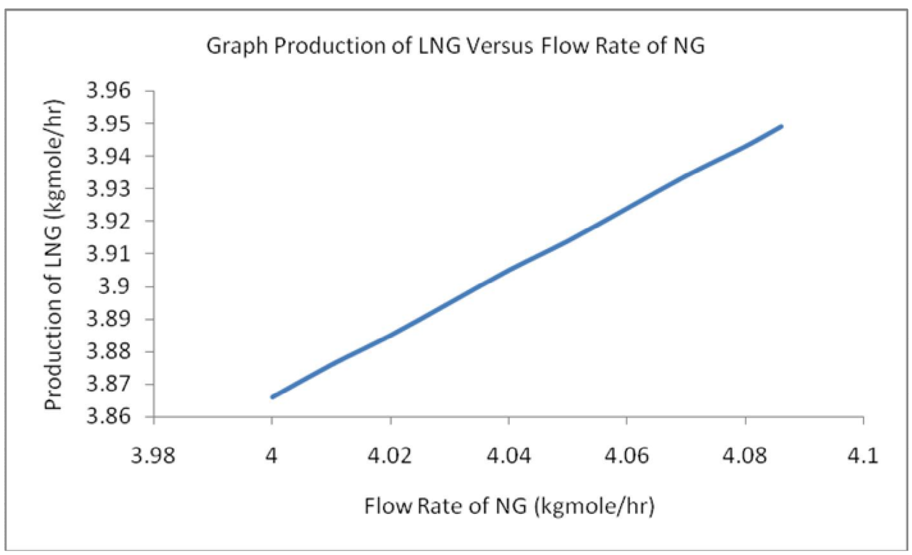

Figure 4.10(d): The Effect on Production of LNG to Increment of NG Flow Rate

4.6 Modification 5 (Addition of Cooler at Stream 15)

The plant performance of modification 5 is analyzed to study the effect of adding cooler at the stream 15.

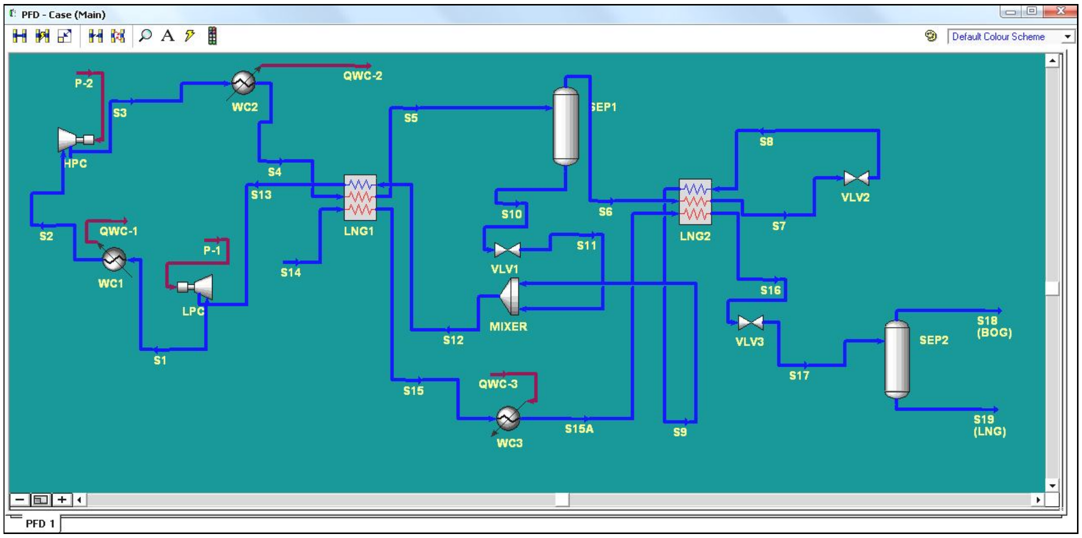

Figure 4.11: Addition of Cooler at Stream 15 (Modification 5)

Simulation result showed that the production of LNG is increase up to 212.600 kg mole/hr compared to the 3.938 kg mole/hr of the base simulation case. The temperature cross condition only occur when the feed flow rate is increase over 220 kg mole/hr. The production of the LNG is increase by 5298.68 % from the actual production which is base case simulation. On the other hand, the power consumption of the compressors is 131.23 kilowatt and the process cooling duty is 4685400 kJ/hr. Figures below showed the effects of the increment of NG flow rate to parameters of the LNG 2 heat exchanger.

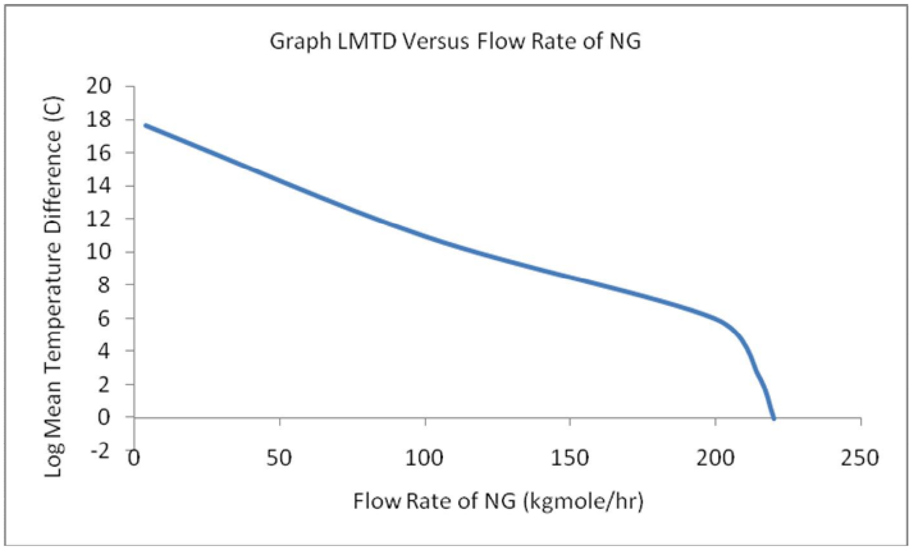

Figure 4.12(a): The Effect on Log Mean Temperature Difference to the Increment of NG Flow Rate

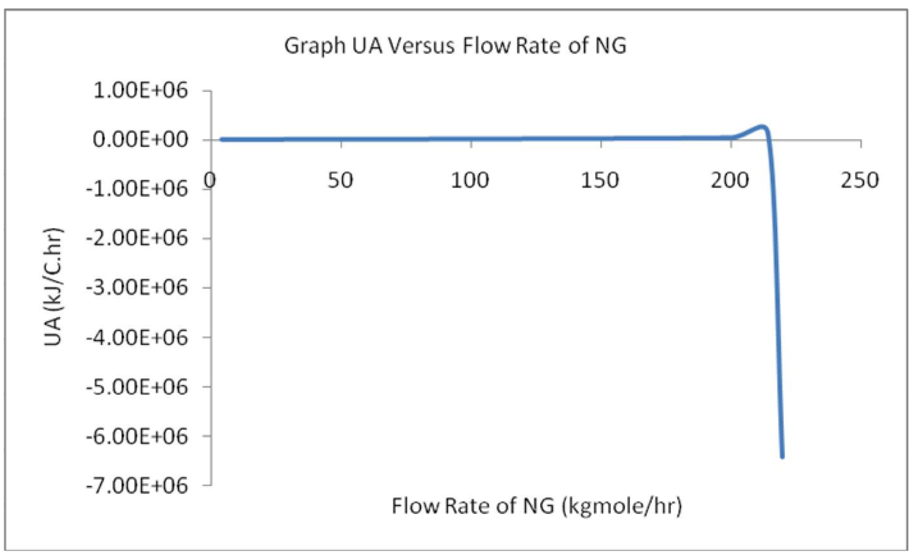

Figure 4.12(b): The Effect on UA to the Increment of NG Flow Rate

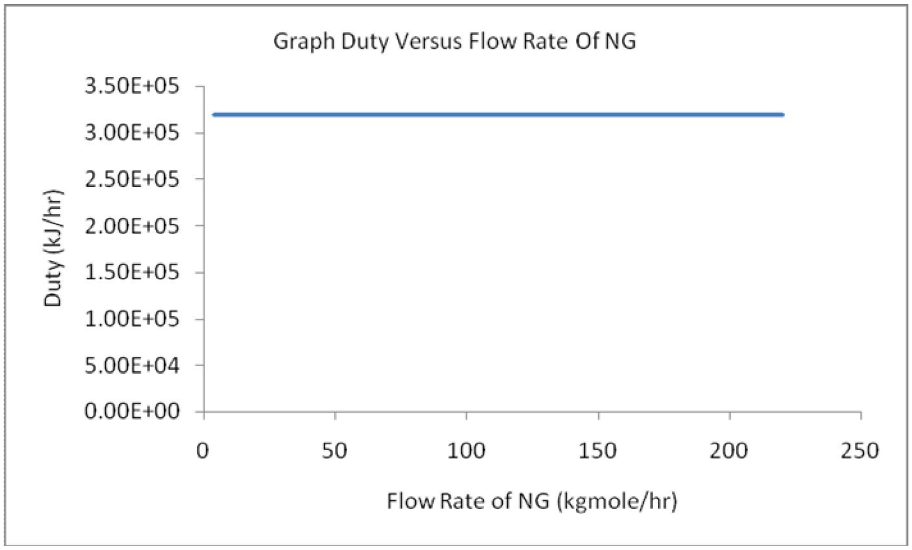

Figure 4.12(c): The Effect on Duty to the Increment of NG Flow Rate

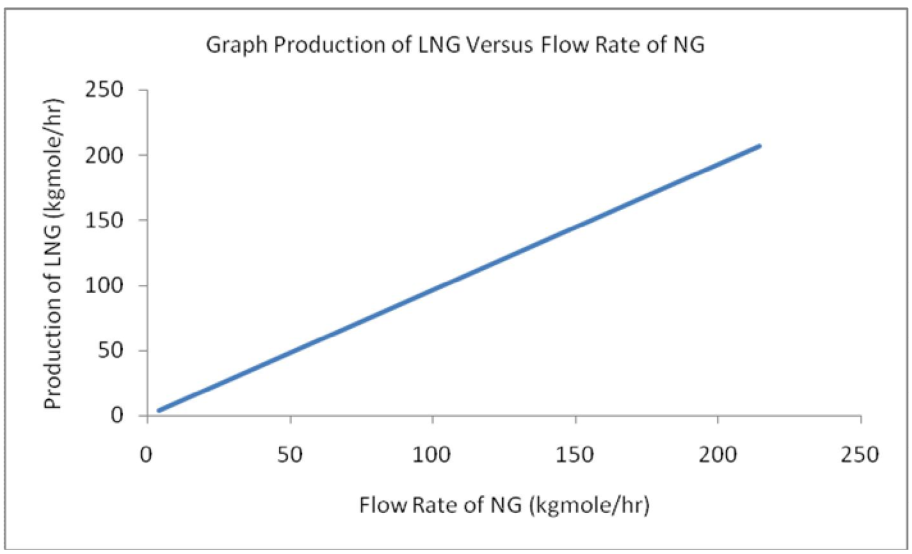

Figure 4.12(d): The Effect on Production of LNG to the Increment of NG Flow Rate

Generally, according to the simulation results, it showed that the duty of the heat exchanger in the bottleneck unit operation was proportionally to the increment of the inlet flow rate. For the log mean temperature difference and heat transfer area, the result showed that it was inversely proportional to the increment of the inlet flow rate and immediately turns to the negative value due to the occurrence of temperature cross in the active bottleneck unit operation. Chapter 1 for this work gives more details on the occurrence of temperature cross condition and heat transfer equipment.

4.7 Economic Evaluation

4.7.1 Existing Plant (Base Case)

1.	Production of LNG (kg mole/hr)	= 3.938 kg mole/hr
2.	Production of LNG (kg/year)	= 693456.28 kg/year
3.	Production of LNG (MMbtu/year)	= 32502.30 MMbtu/year
4.	Revenue of LNG sold per year	= RM 505735.72
5.	Power Consumption of Compressors	= 131.23 kilowatt
6.	Cost of Power Consumption	= RM 28503.16 / year
7.	Cooling Duty	= 1518500 kJ/hr
8.	Cost of Cooling Duty	= RM 26458.34/year
9.	Production Revenue	= RM 450774.22

4.7.2 Modification 1 (Addition of Cooler at Stream 6)

1. Production of LNG (kg mole/hr) = 3.930 kg mole/hr
2. Production of LNG (kg/year) = 692047.53 kg/year
3. Production of LNG (MMbtu/year) = 32436.27 MMbtu/year
4. Revenue of LNG sold per year = RM 504708.33
5. Power Consumption of Compressors = 131.23 kilowatt
6. Cost of Power Consumption = RM 28503.16
7. Cooling Duty = 1518405.64 kJ/hr
8. Price of Cooling Duty = RM 26456.70/year
9. Cost of Additional Cooler = RM 24122
10. Production Revenue = RM 425626.47

Then, the production revenue after one year debottlenecking takes place is,

= RM 449748.47

4.7.3 Modification 2 (Addition of Cooler at Stream 7)

1. Production of LNG (kg mole/hr) = 3.944 kg mole/hr
2. Production of LNG (kg/year) = 694512.84 kg/year
3. Production of LNG (MMbtu/year) = 32551.81695 MMbtu/year
4. Revenue of LNG sold per year = RM 506506.27
5. Power Consumption of Compressors = 131.23 kilowatt
6. Cost of Power Consumption = RM 28503.16/ year
7. Cooling Duty = 1521614.30 kJ/hr
8. Price of Cooling Duty = RM 26512.60
9. Cost of Additional Cooler = RM 120000
10. Production Revenue = RM 357255.00

Then, the production revenue after one year debottlenecking takes place is,

= RM 451490.50

4.7.4 Modification 3 (Addition of Cooler at Stream 8)

1.	Production of LNG (kg mole/hr)	=11.69 kg mole/hr
2.	Production of LNG (kg/year)	= 2044445.77 kg/year
3.	Production of LNG (MMbtu/year)	= 95823.17 MMbtu/year
4.	Revenue of LNG sold per year	= RM 1491008.57
5.	Power Consumption of Compressors	= 131.23 kilowatt
6.	Cost of Power Consumption	= RM 28503.16/year
7.	Cooling Duty	= 1760400 kJ/hr
8.	Cost of Cooling Duty	= RM 30673.21
9.	Cost of Additional Cooler	= RM 132774.00
10.	Production Revenue	= RM 1299058.20

Then, the production revenue after one year debottlenecking takes place is,

= RM 1431832.20

4.7.5 Modification 4 (Additional Cooler at Stream 9)

1. Production of LNG (kg mole/hr)	=3.949 kg mole/hr
2. Production of LNG (kg/year)	= 695393.31 kg/year
3. Production of LNG (MMbtu/year)	= 32593.08 MMbtu/year
4. Revenue of LNG sold per year	= RM 507148.40
5. Power Consumption of Compressors	= 131.23 kilowatt
6. Price of Power Consumption	= RM 28503.16
7. Cooling Duty	= 1519008 kJ/hr
8. Cost of Cooling Duty	= RM 26467.20
9. Cost of Additional Cooler	= RM 160938.00
10. Production Revenue	= RM 291240.00

Then, the production revenue after one year debottlenecking takes place is,

= RM 452178.00

4.7.6 Modification 5 (Addition of Cooler at Stream 15)

1. Production of LNG (kg mole/hr) = 207.3 kg mole/hr
2. Production of LNG (kg/year) = 36504186.70 kg/year
3. Production of LNG (MMbtu/year) = 1710951.23 MMbtu/year
4. Revenue of LNG sold per year = RM 2662240.15
5. Power Consumption of Compressors = 131.23 kilowatt
6. Cost of Power Consumption = RM 28503.16
7. Cooling Duty = 4685400 kJ/hr
8. Cost of Cooling Duty = RM 81638.40
9. Cost of Additional Cooler = RM 400000
10. Production Revenue = RM 2152098.60

Then, the production revenue after one year debottlenecking takes place is,

= RM 2552098.60

From the evaluation on the plant performance that has been done, some assumption and consideration has been made. In order to calculate the cost of the power consumption for compressors, the data is taken from the Tenaga Nasional Berhad (TNB) tariff rate (1 March 2009) and considered as Special Industrial Tariff- for Consumer Who Qualify Only (E1s). On the other hand, by referring to Table 3.4 (Engineering Economic Analysis), cost of the process cooling duty is calculated. The price of the LNG is RM 15.56 per MMbtu.

As the bottleneck take places in this processing plant, debottlenecking process should be done. For this work, five modifications have been proposed excluding the base case plant. Debottlenecking process in this work focuses on the addition of the cooler at the specified stream to analyze the plant performance. Then, in economic evaluation process, the purchase cost of this addition cooler play an important role in order to estimate the production revenue for every scheme proposed. By using the Kern's method (Coulson & Richardson's Chemical Engineering), the heat transfer area of the cooler is calculated. Overall coefficient, U, is specified into 500 $W/m^{2\cdot}C$ with the correction factor, F_t, of 0.75. This value is specified into 0.75 because an economic exchanger design cannot normally be achieved if the correction factor, F_t, falls below about 0.75. Table 4.1 in Appendix D showed the summary of the economic analysis of this work.

CHAPTER 5

CONCLUSION AND FUTURE WORK

Process Debottlenecking of Small Scale LNG plant has achieved two main objectives. The first objective is to debottleneck the small scale LNG plant in order to increase the throughput or production capacity. The second objective is to design the schemes for the debottlenecking process (increment of the production) with the analyzing of the performance (economic evaluation) of the modification schemes after debottlenecking process has been implemented.

The existing small scale LNG processing plant has been simulated and converged successfully. The simulation works were done via Aspen HYSYS and the results were validated against the data published in the literature (H.A. Razik, 2007). After debottlenecking process has been implemented, the results showed that for this existing plant, the maximum production of LNG can be achieved before bottlenecks condition takes place was 3.938 kg mole/ hr with the 131.23 kilowatt for the power consumption of the compressor and 1518500 kJ/hr of process cooling duty.

With the complete process data, the analysis and design of the modification schemes of the MRC liquefaction process then could be performed. Modification 5 showed the highest percentage of LNG production with 5298. 68% increment from the existing plant. Production revenue for this modification is RM 2552098.60 after takes into consideration the highest cost of its additional cooling duty which is RM 400000.00.

For future work, economic benefits for this work need to be further analyzed so that the importance of process debottlenecking of LNG plant become more comprehensive. It also strong recommended that to analyze the thermodynamic performance of the Mixed Refrigeration Cycle for the refrigeration system in this LNG processing plant in this case of using optimal mixture composition. On the other hand, because of the limitations of the Aspen HYSYS Software to predict the heat transfer area of the heat exchanger, it also recommends that, the use of others software such as UNISIM to overcome the bottleneck condition without do some modifications or alterations of the existing plant. Further analysis can be done after the prediction of the heat transfer area of that heat exchanger such as retrofitting process in order to increase their performance especially with strengthen their heat transfer.

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APPENDIX A

NOTATION

BOG	Boiled Off Gas
C_2H_6	Ethane
C_3H_8	Propane
CH_4	Methane
HPC	High Pressure Compressor
HYSYS	HYSYS Software
iC_4H_{10}	i- Butane
LNG	Liquefied Natural Gas
LNG1	Heat Exchanger 1
LNG2	Heat Exchanger 2
LPC	Low Pressure Compressor
LKP	Lee-Kesler-Plocker
MR	Mixed Refrigerant
MRC	Mixed Refrigerant Cycle
MTPA	Million Tonnes per Annum
N_2	Nitrogen
nC_4H_{10}	n- Butane

NG	Natural Gas
NGL_S	Natural Gas Liquids
PR	Peng- Robinson
S	Stream
SEP	Separator
TNB	Tenaga Nasional Berhad
VLV	Valve
WC	Water Cooler
WC3	Additional Cooler

APPENDIX B

This appendix gives the Aspen HYSYS simulation reports for the five modifications of the original flow sheet of the small scale LNG plant (existing plant) due to the debottlenecking process for boost the production of the LNG. It also includes the existing plant Aspen HYSYS simulation report as a guideline and the reference.

B.1 Existing Plant

B.2 Modification 1 (Addition of Cooler at Stream 6)

B.3 Modification 2 (Addition of Cooler at Stream 7)

B.4 Modification 3 (Addition of Cooler at Stream 8)

B.5 Modification 4 (Addition of Cooler at Stream 9)

B.6 Modification 5 (Addition of Cooler at Stream 15)

B.1 Existing Plant

Name	S15	S13	S1	S2	S3	S4	S5	S12	S6	S16	S17	S18 (BOG)
Vapour Fraction	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.3708	0.6243	1.0000	0.0000	0.0250	1.0000
Temperature [C]	-70.00	29.00	113.1	29.00	98.45	31.00	-35.00	-74.49	-35.18	-147.0	-148.3	-149.4
Pressure [kPa]	4980	290.0	990.0	970.0	2620	2600	2580	310.0	2560	4960	250.0	230.0
Molar Flow [kgmole/h]	4.074	60.25	60.25	60.25	60.25	60.25	60.25	60.25	22.44	4.074	4.074	0.1364
Mass Flow [kg/h]	80.27	1633	1633	1633	1633	1633	1633	1633	457.0	80.27	80.27	2.340
Std Ideal Liq Vol Flow [m3/h]	0.2415	4.346	4.346	4.346	4.346	4.346	4.346	4.346	1.373	0.2415	0.2415	7.075e-003
Heat Flow [kJ/h]	-3.611e+005	-5.032e+006	-4.764e+006	-5.051e+006	-4.846e+006	-5.093e+006	-5.718e+006	-5.697e+006	-1.787e+006	-3.824e+005	-3.824e+005	-1.009e+004
Molar Enthalpy [kJ/kgmole]	-8.863e+004	-8.352e+004	-7.906e+004	-8.383e+004	-8.044e+004	-8.453e+004	-9.491e+004	-9.456e+004	-7.966e+004	-9.387e+004	-9.387e+004	-7.398e+004
Name	S19 (LNG)	S10	S11	S14	S9	S8	S7	P-1	QWC-1	P-2	QWC-2	** New **
Vapour Fraction	0.0000	0.0000	0.0024	1.0000	1.0000	0.0277	0.0000	<empty>	<empty>	<empty>	<empty>	
Temperature [C]	-149.4	-35.18	-35.34	32.00	-35.40	-150.0	-148.0	<empty>	<empty>	<empty>	<empty>	
Pressure [kPa]	230.0	2560	2540	5000	310.0	261.5	2510	<empty>	<empty>	<empty>	<empty>	
Molar Flow [kgmole/h]	3.938	37.81	37.81	4.074	22.44	22.44	22.44	<empty>	<empty>	<empty>	<empty>	
Mass Flow [kg/h]	77.93	1176	1176	80.27	457.0	457.0	457.0	<empty>	<empty>	<empty>	<empty>	
Std Ideal Liq Vol Flow [m3/h]	0.2344	2.973	2.973	0.2415	1.373	1.373	1.373	<empty>	<empty>	<empty>	<empty>	
Heat Flow [kJ/h]	-3.723e+005	-3.931e+006	-3.931e+006	-3.211e+005	-1.766e+006	-2.086e+006	-2.086e+006	2.684e+005	2.869e+005	2.040e+005	2.466e+005	
Molar Enthalpy [kJ/kgmole]	-9.455e+004	-1.040e+005	-1.040e+005	-7.882e+004	-7.871e+004	-9.297e+004	-9.297e+004	<empty>	<empty>	<empty>	<empty>	

B.2 Modification 1 (Addition of Cooler at Stream 6)

Name	S15	S13	S1	S2	S3	S4	S5	S12	S16	S17	S18 (BOG)	S19 (LNG)
Vapour Fraction	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.3708	0.6243	0.0000	0.0250	1.0000	0.0000
Temperature [C]	-70.25	29.00	113.1	29.00	98.45	31.00	-35.00	-74.49	-147.0	-148.3	-149.4	-149.4
Pressure [kPa]	4980	290.0	990.0	970.0	2620	2600	2580	310.0	4960	250.0	230.0	230.0
Molar Flow [kgmole/h]	4.066	60.25	60.25	60.25	60.25	60.25	60.25	60.25	4.066	4.066	0.1361	3.930
Mass Flow [kg/h]	80.11	1633	1633	1633	1633	1633	1633	1633	80.11	80.11	2.335	77.78
Std Ideal Liq Vol Flow [m3/h]	0.2410	4.346	4.346	4.346	4.346	4.346	4.346	4.346	0.2410	0.2410	7.061e-003	0.2339
Heat Flow [kJ/h]	-3.605e+005	-5.032e+006	-4.764e+006	-5.051e+006	-4.846e+006	-5.093e+006	-5.718e+006	-5.697e+006	-3.817e+005	-3.817e+005	-1.007e+004	-3.716e+005
Molar Enthalpy [kJ/kgmole]	-8.865e+004	-8.352e+004	-7.906e+004	-8.383e+004	-8.044e+004	-8.453e+004	-9.491e+004	-9.456e+004	-9.387e+004	-9.387e+004	-7.398e+004	-9.455e+004

Name	S10	S11	S14	S9	S8	S7	S6A	S6	P-1	QWC-1	P-2	QWC-2
Vapour Fraction	0.0000	0.0024	1.0000	1.0000	0.0277	0.0000	0.9999	1.0000	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-35.18	-35.34	32.00	-35.40	-150.0	-148.0	-35.37	-35.18	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	2560	2540	5000	310.0	261.5	2510	2540	2560	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	37.81	37.81	4.066	22.44	22.44	22.44	22.44	22.44	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	1176	1176	80.11	457.0	457.0	457.0	457.0	457.0	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	2.973	2.973	0.2410	1.373	1.373	1.373	1.373	1.373	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-3.931e+006	-3.931e+006	-3.205e+005	-1.766e+006	-2.086e+006	-2.086e+006	-1.787e+006	-1.787e+006	2.684e+005	2.869e+005	2.040e+005	2.466e+005
Molar Enthalpy [kJ/kgmole]	-1.040e+005	-1.040e+005	-7.882e+004	-7.871e+004	-9.297e+004	-9.297e+004	-7.966e+004	-7.966e+004	<empty>	<empty>	<empty>	<empty>

Name	QAC-3	** New **										
Vapour Fraction	<empty>											
Temperature [C]	<empty>											
Pressure [kPa]	<empty>											
Molar Flow [kgmole/h]	<empty>											
Mass Flow [kg/h]	<empty>											
Std Ideal Liq Vol Flow [m3/h]	<empty>											
Heat Flow [kJ/h]	5.638											
Molar Enthalpy [kJ/kgmole]	<empty>											

B.3 Modification 2 (Addition of Cooler at Stream 7)

Name	S15	S13	S1	S2	S3	S4	S5	S12	S6	S16	S17	S18 (BOG)
Vapour Fraction	0.0056	1.0000	1.0000	1.0000	1.0000	1.0000	0.3708	0.6243	1.0000	0.0000	0.0250	1.0000
Temperature [C]	-69.90	29.00	113.1	29.00	98.45	31.00	-35.00	-74.49	-35.18	-147.0	-148.3	-149.4
Pressure [kPa]	4980	290.0	990.0	970.0	2620	2600	2580	310.0	2560	4960	250.0	230.0
Molar Flow [kgmole/h]	4.082	60.25	60.25	60.25	60.25	60.25	60.25	60.25	22.44	4.082	4.082	0.1367
Mass Flow [kg/h]	80.43	1633	1633	1633	1633	1633	1633	1633	457.0	80.43	80.43	2.344
Std Ideal Liq Vol Flow [m3/h]	0.2419	4.346	4.346	4.346	4.346	4.346	4.346	4.346	1.373	0.2419	0.2419	7.089e-003
Heat Flow [kJ/h]	-3.617e+005	-5.032e+006	-4.764e+006	-5.051e+006	-4.846e+006	-5.093e+006	-5.718e+006	-5.697e+006	-1.787e+006	-3.832e+005	-3.832e+005	-1.011e+004
Molar Enthalpy [kJ/kgmole]	-8.860e+004	-8.352e+004	-7.906e+004	-8.383e+004	-8.044e+004	-8.453e+004	-9.491e+004	-9.456e+004	-7.966e+004	-9.387e+004	-9.387e+004	-7.398e+004
Name	S19 (LNG)	S10	S11	S14	S9	S8	S7	S7a	P-1	QWC-1	P-2	QWC-2
Vapour Fraction	0.0000	0.0000	0.0024	1.0000	1.0000	0.0086	0.0000	0.0000	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-149.4	-35.18	-35.34	32.00	-35.40	-150.0	-149.8	-150.0	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	230.0	2560	2540	5000	310.0	290.1	2510	2490	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	3.945	37.81	37.81	4.082	22.44	22.44	22.44	22.44	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	78.09	1176	1176	80.43	457.0	457.0	457.0	457.0	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	0.2349	2.973	2.973	0.2419	1.373	1.373	1.373	1.373	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-3.730e+005	-3.931e+006	-3.931e+006	-3.217e+005	-1.766e+006	-2.089e+006	-2.089e+006	-2.089e+006	2.684e+005	2.869e+005	2.040e+005	2.466e+005
Molar Enthalpy [kJ/kgmole]	-9.455e+004	-1.040e+005	-1.040e+005	-7.882e+004	-7.871e+004	-9.309e+004	-9.308e+004	-9.309e+004	<empty>	<empty>	<empty>	<empty>
Name	QAC-1	** New **										
Vapour Fraction	<empty>											
Temperature [C]	<empty>											
Pressure [kPa]	<empty>											
Molar Flow [kgmole/h]	<empty>											
Mass Flow [kg/h]	<empty>											
Std Ideal Liq Vol Flow [m3/h]	<empty>											
Heat Flow [kJ/h]	314.3											
Molar Enthalpy [kJ/kgmole]	<empty>											

B.4 Modification 3 (Addition of Cooler at Stream 8)

Name	S15	S13	S1	S2	S3	S4	S5	S12	S6	S16	S17	S18 (BOG)
Vapour Fraction	0.9468	1.0000	1.0000	1.0000	1.0000	1.0000	0.3708	0.6243	1.0000	0.0000	0.0250	1.0000
Temperature [C]	-23.08	29.00	113.1	29.00	98.45	31.00	-35.00	-74.49	-35.18	-147.0	-148.3	-149.4
Pressure [kPa]	4980	290.0	990.0	970.0	2620	2600	2580	310.0	2560	4960	250.0	230.0
Molar Flow [kgmole/h]	12.10	60.25	60.25	60.25	60.25	60.25	60.25	60.25	22.44	12.10	12.10	0.4052
Mass Flow [kg/h]	238.4	1633	1633	1633	1633	1633	1633	1633	457.0	238.4	238.4	6.949
Std Ideal Liq Vol Flow [m3/h]	0.7172	4.346	4.346	4.346	4.346	4.346	4.346	4.346	1.373	0.7172	0.7172	2.101e-002
Heat Flow [kJ/h]	-9.936e+005	-5.032e+006	-4.764e+006	-5.051e+006	-4.846e+006	-5.093e+006	-5.718e+006	-5.697e+006	-1.787e+006	-1.136e+006	-1.136e+006	-2.997e+004
Molar Enthalpy [kJ/kgmole]	-8.212e+004	-8.352e+004	-7.906e+004	-8.383e+004	-8.044e+004	-8.453e+004	-9.491e+004	-9.456e+004	-7.966e+004	-9.387e+004	-9.387e+004	-7.398e+004
Name	S19 (LNG)	S10	S11	S14	S9	S7	S8A	S8	P-1	QWC-1	P-2	QWC-2
Vapour Fraction	0.0000	0.0000	0.0024	1.0000	1.0000	0.0000	0.0000	0.0277	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-149.4	-35.18	-35.34	32.00	-35.40	-148.0	-228.0	-150.0	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	230.0	2560	2540	5000	310.0	2510	241.0	261.5	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	11.69	37.81	37.81	12.10	22.44	22.44	22.44	22.44	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	231.5	1176	1176	238.4	457.0	457.0	457.0	457.0	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	0.6962	2.973	2.973	0.7172	1.373	1.373	1.373	1.373	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-1.106e+006	-3.931e+006	-3.931e+006	-9.537e+005	-1.766e+006	-2.086e+006	-2.207e+006	-2.086e+006	2.684e+005	2.869e+005	2.040e+005	2.466e+005
Molar Enthalpy [kJ/kgmole]	-9.455e+004	-1.040e+005	-1.040e+005	-7.882e+004	-7.871e+004	-9.297e+004	-9.836e+004	-9.297e+004	<empty>	<empty>	<empty>	<empty>
Name	QAC-1	** New **										
Vapour Fraction	<empty>											
Temperature [C]	<empty>											
Pressure [kPa]	<empty>											
Molar Flow [kgmole/h]	<empty>											
Mass Flow [kg/h]	<empty>											
Std Ideal Liq Vol Flow [m3/h]	<empty>											
Heat Flow [kJ/h]	1.210e+005											
Molar Enthalpy [kJ/kgmole]	<empty>											

B.5 Modification 4 (Addition of Cooler at Stream 9)

Name	S15	S13	S1	S2	S3	S4	S5	S12	S6	S16	S17	S18 (BOG)
Vapour Fraction	0.0054	1.0000	1.0000	1.0000	1.0000	1.0000	0.3708	0.6243	1.0000	0.0000	0.0250	1.0000
Temperature [C]	-69.90	29.00	113.1	29.00	98.45	31.00	-35.00	-74.49	-35.18	-147.0	-148.3	-149.4
Pressure [kPa]	4980	290.0	990.0	970.0	2620	2600	2580	310.0	2560	4960	250.0	230.0
Molar Flow [kgmole/h]	4.085	60.25	60.25	60.25	60.25	60.25	60.25	60.25	22.44	4.085	4.085	0.1368
Mass Flow [kg/h]	80.49	1633	1633	1633	1633	1633	1633	1633	457.0	80.49	80.49	2.346
Std Ideal Liq Vol Flow [m3/h]	0.2421	4.346	4.346	4.346	4.346	4.346	4.346	4.346	1.373	0.2421	0.2421	7.094e-003
Heat Flow [kJ/h]	-3.619e+005	-5.032e+006	-4.764e+006	-5.051e+006	-4.846e+006	-5.093e+006	-5.718e+006	-5.697e+006	-1.787e+006	-3.834e+005	-3.834e+005	-1.012e+004
Molar Enthalpy [kJ/kgmole]	-8.860e+004	-8.352e+004	-7.906e+004	-8.383e+004	-8.044e+004	-8.453e+004	-9.491e+004	-9.456e+004	-7.966e+004	-9.387e+004	-9.387e+004	-7.398e+004
Name	S19 (LNG)	S10	S11	S14	S9	S8	S7	S9A	P-1	QWC-1	P-2	QWC-2
Vapour Fraction	0.0000	0.0000	0.0024	1.0000	1.0000	0.0277	0.0000	1.0000	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-149.4	-35.18	-35.34	32.00	-35.40	-150.0	-148.0	-35.23	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	230.0	2560	2540	5000	310.0	261.5	2510	290.0	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	3.948	37.81	37.81	4.085	22.44	22.44	22.44	22.44	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	78.14	1176	1176	80.49	457.0	457.0	457.0	457.0	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	0.2350	2.973	2.973	0.2421	1.373	1.373	1.373	1.373	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-3.733e+005	-3.931e+006	-3.931e+006	-3.220e+005	-1.766e+006	-2.086e+006	-2.086e+006	-1.766e+006	2.684e+005	2.869e+005	2.040e+005	2.466e+005
Molar Enthalpy [kJ/kgmole]	-9.455e+004	-1.040e+005	-1.040e+005	-7.882e+004	-7.871e+004	-9.297e+004	-9.297e+004	-7.870e+004	<empty>	<empty>	<empty>	<empty>
Name	QAC-1	** New **										
Vapour Fraction	<empty>											
Temperature [C]	<empty>											
Pressure [kPa]	<empty>											
Molar Flow [kgmole/h]	<empty>											
Mass Flow [kg/h]	<empty>											
Std Ideal Liq Vol Flow [m3/h]	<empty>											
Heat Flow [kJ/h]	308.0											
Molar Enthalpy [kJ/kgmole]	<empty>											

B.6 Modification 5 (Addition of Cooler at Stream 15)

Name	S15	S13	S1	S2	S3	S4	S5	S12	S6	S16	S17	S18 (BOG)
Vapour Fraction	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.3708	0.6243	1.0000	0.0000	0.0250	1.0000
Temperature [C]	28.14	29.00	113.1	29.00	98.45	31.00	-35.00	-74.49	-35.18	-147.0	-148.3	-149.4
Pressure [kPa]	4980	290.0	990.0	970.0	2620	2600	2580	310.0	2560	4960	250.0	230.0
Molar Flow [kgmole/h]	214.5	60.25	60.25	60.25	60.25	60.25	60.25	60.25	22.44	214.5	214.5	7.182
Mass Flow [kg/h]	4226	1633	1633	1633	1633	1633	1633	1633	457.0	4226	4226	123.2
Std Ideal Liq Vol Flow [m3/h]	12.71	4.346	4.346	4.346	4.346	4.346	4.346	4.346	1.373	12.71	12.71	0.3725
Heat Flow [kJ/h]	-1.695e+007	-5.032e+006	-4.764e+006	-5.051e+006	-4.846e+006	-5.093e+006	-5.718e+006	-5.697e+006	-1.787e+006	-2.013e+007	-2.013e+007	-5.314e+005
Molar Enthalpy [kJ/kgmole]	-7.900e+004	-8.352e+004	-7.906e+004	-8.383e+004	-8.044e+004	-8.453e+004	-9.491e+004	-9.456e+004	-7.966e+004	-9.387e+004	-9.387e+004	-7.398e+004
Name	S19 (LNG)	S10	S11	S14	S9	S8	S7	S15A	P-1	QWC-1	P-2	QWC-2
Vapour Fraction	0.0000	0.0000	0.0024	1.0000	1.0000	0.0277	0.0000	0.0000	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-149.4	-35.18	-35.34	32.00	-35.40	-150.0	-148.0	-145.3	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	230.0	2560	2540	5000	310.0	261.5	2510	4960	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	207.3	37.81	37.81	214.5	22.44	22.44	22.44	214.5	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	4103	1176	1176	4226	457.0	457.0	457.0	4226	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	12.34	2.973	2.973	12.71	1.373	1.373	1.373	12.71	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-1.960e+007	-3.931e+006	-3.931e+006	-1.691e+007	-1.766e+006	-2.086e+006	-2.086e+006	-2.011e+007	2.684e+005	2.869e+005	2.040e+005	2.466e+005
Molar Enthalpy [kJ/kgmole]	-9.455e+004	-1.040e+005	-1.040e+005	-7.882e+004	-7.871e+004	-9.297e+004	-9.297e+004	-9.377e+004	<empty>	<empty>	<empty>	<empty>
Name	QAC-1	** New **										
Vapour Fraction	<empty>											
Temperature [C]	<empty>											
Pressure [kPa]	<empty>											
Molar Flow [kgmole/h]	<empty>											
Mass Flow [kg/h]	<empty>											
Std Ideal Liq Vol Flow [m3/h]	<empty>											
Heat Flow [kJ/h]	-3.167e+006											
Molar Enthalpy [kJ/kgmole]	<empty>											

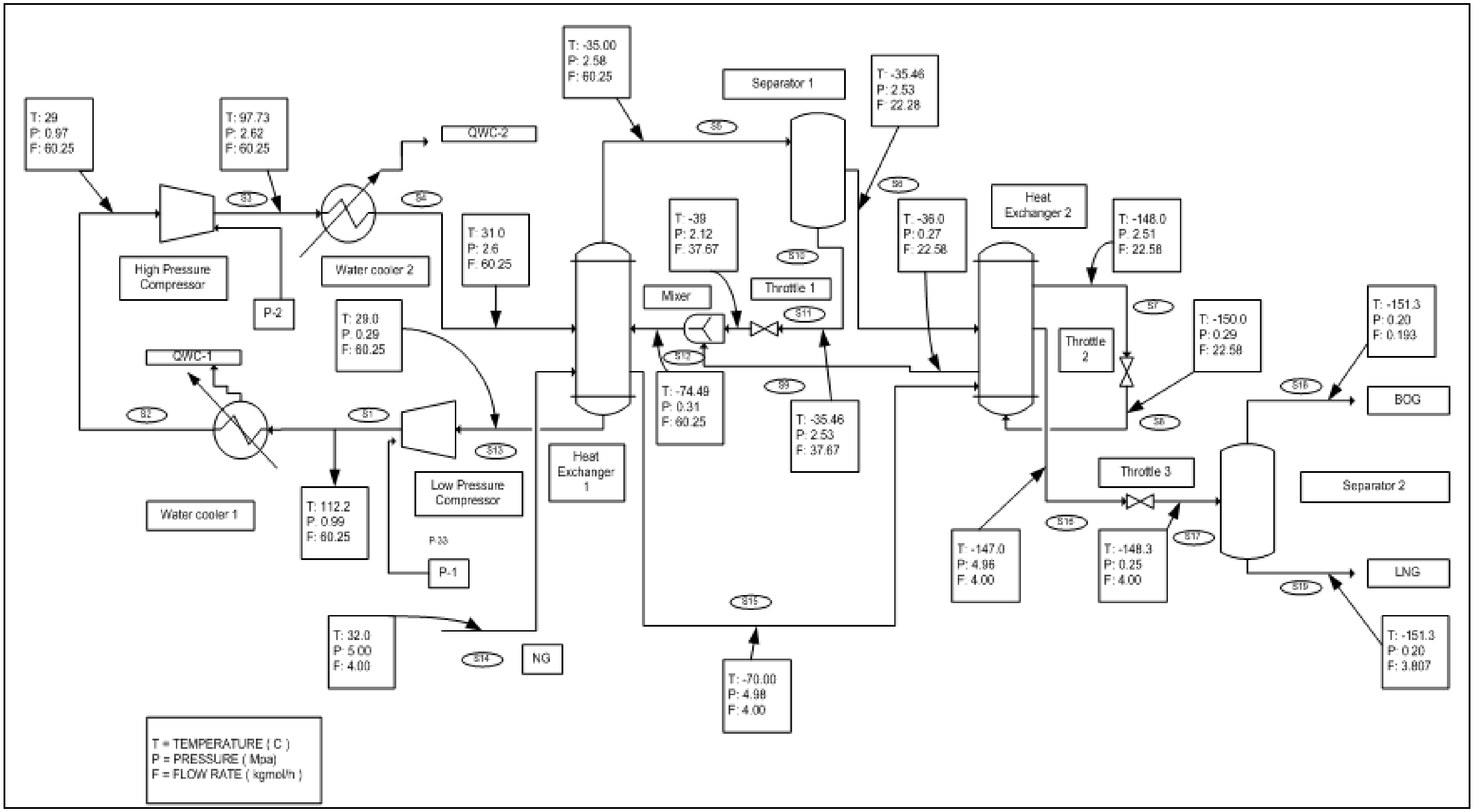

Figure 3.1: The Mixed-Refrigerant Cycle Liquefaction Process with Specification Data (Analysis of Optimal Power Reduction Schemes for an LNG Plant) - APPENDIX C

APPENDIX D

Table 4.1: Summary of Economic Analysis

Parameters	Existing Plant	Modification 1	Modification 2	Modification 3	Modification 4	Modification 5
1. Production of LNG (MMbtu/year)	32502.30	32436.27	32551.817	95823.17	32593.08	1710951.23
2. Revenue of LNG sold per year	RM 505735.72	RM 504708.33	RM 506506.27	RM 1491008.57	RM 507148.40	RM 2662240.15
3. Power Consumption of Compressor (kW)	131.23	131.23	131.23	131.23	131.23	131.23
4. Cost of Power Consumption per year	RM 28503.16	RM 28503.16	RM 28503.16	RM 28503.16	RM 28503.16	RM 28503.16
5. Cooling Duty (kJ/hr)	1518500	1518405.64	1521614.30	1760400	1519008	4685400
6. Cost of Cooling Duty per year	RM 26458.34	RM 26456.70	RM 26512.60	RM 30673.21	RM 26467.20	RM 81638.40
7. Cost of Additional Cooler	-	RM 24122.00	RM 120000.00	RM 132774.00	RM 160938.00	RM 400000.00
8. Production Revenue	RM 450774.22	RM 425626.47	RM 357255.00	RM 1299058.20	RM 291240.00	RM 2152098.60
9. Production Revenue (After Debottlenecking Process)	RM 450774.22	RM 449748.47	RM 451490.50	RM 1431832.20	RM 452178.00	RM 2552098.60