HYSYS SIMULATION AND OPTIMIZATION OF AN LNG PLANT'S BACK-END PROCESS

by:

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13968

Dissertation submitted in partial fulfilment of

the requirement for the

Bachelor of Engineering (Hons)

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CERTIFICATION OF APPROVAL

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Approved by,

………………………………

(AP Dr. Shuhaimi Mahadzir)

UNIVERSITI TEKNOLOGI PETRONAS

TRONOH, PERAK

September, 2014

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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(Carmelo Ciriaco Esono Etetere)

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ABSTRACT

End-flash system is a mechanism applied in LNG processes such as the ConocoPhillips optimized cascade process to reject nitrogen content in the liquefied natural gas as consequence of the storage tank blanketing. In order to meet client LNG quality requirement the end-flash system reduces nitrogen content by rejecting the nitrogen rich natural gas as fuel to fuel up the heavy gas turbines. This project aim is to simulate and optimize a base case simulation fig. 9 by modifying the base case with the ultimate objective to increase produced LNG which could increase plant benefits and to reduce the fuel gas production. ConocoPhillips optimized cascade back-end process simulation is the base case for this study. With a start-up feed of 50000kg/hr which is computed and reduced to 15440kg/hr once the recycle flow joins and adjusts the feed, a production of LNG 13500kg/hr (87.44% feed) and fuel gas 1825 kg/hr (11.82% adjusted feed) yielding a specific power of 903kJ per Kg of LNG produced. Modified simulations have been performed exploring the opportunity to improve the correlation of LNG production and fuel gas efficiency. Two approaches has been tackled by modifying the number of sub-cooling stages in one direction by reducing number of cooling stages from three to a single stage and achieving an improvement of 9680kg/hr of more LNG production and fuel production reduced to 5.44%. On the other direction the number of sub-cooling stages was increased from three to four stages and this approach yield results of 13660kg/hr of more produced LNG and reduced the fuel gas production to 4.04%. This approach presents an overall improvement of 61% reduction of the required power to produce 1Kg of LNG, yielding to 349 KJ/Kg. Increasing the number of sub-cooling stages resulted to be the most efficient approach with optimal results.

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CHAPTER 1: INTRODUCTION

1.1 BACKGROUND

Liquefied natural gas (LNG) is stored in pressures slightly above atmospheric pressure just to maintain positive pressure at all times eliminating the possibility of oxygen presence in LNG storage tanks. However boiling gases are produced constantly at the top side of these tanks and inert gases are used to keep the positive pressure inside those tanks and the boil off gas(BOG) mixed with nitrogen are constantly taken out of the tank via a BOG compressor and LNG is then recovered in the liquefaction area. With nitrogen being injected into the methane cycle at the top side of LNG storage tanks, it is required to get rid of that nitrogen already in the system later on in order to limit the build-up of nitrogen in the methane cycle, in this way LNG client's specification is maintained within target (Vink, 1998). The way LNG processes achieve this is by rejecting the nitrogen in the first stage of methane compression in the flash-end gas stage. In order not to waste natural gas with high concentration of nitrogen it is instead conditioned and used as fuel for the process. LNG processes produce the required fuel used in the liquefaction process.

This study will be carried out using ConocoPhillips optimized cascade process flowsheet for simulation in Hysys. These plants are designed to perform with 95% LNG production efficiency and not to exceed 5% of fuel gas production.

1.2 PROBLEM STATEMENT

Five percent is the design limit for fuel gas production in LNG plants as the strategy to reject nitrogen out of the system. However, the fuel gas produced has its economic value. Is it possible to recover part of that LNG being used as fuel and increasing LNG sales and still meet the process fuel demand? Those are the questions this project is intending to answer with Hysys simulations.

1.3 OBJECTIVES

The objectives of this project FYP 1 are:

- To produce a simulation model using Aspen Hysys as base case of LNG back end process based on the ConocoPhillips process;
- To explore opportunity to increase energy and LNG production efficiency through flowsheet modification of the base case.

1.4 SCOPE OF STUDY

The scope of this study covers the back end process of an LNG cascade process (COPOC). Feed pre-treatment, refrigerant loops but methane, LNG storage and loading facilities are not included in the scope of this study. The about 5% fuel gas production is the focus of this study. The table below shows assumed mass fraction composition of LNG for this study:

Table 1: LNG mol fraction composition

LNG and generated BOGs' compositions considered in this study. Querol et al. and Aspen Plus™.

	LNG Light	Medium	Heavy
CH_4 [% mol]	98.60	92.30	85.87
C_2H_6 [% mol]	1.18	5.00	8.40
C_3H_8 [% mol]	0.10	1.50	3.00
C_4H_{10} [% mol]	0.02	0.60	1.20
C_5H_{12} [% mol]	0.00	0.10	0.23
N_2 [% mol]	0.10	0.50	1.30
Storage temperature [°C]	−159.95	−160.54	−163.23
	BOG Light	Medium	Heavy
CH_4 [% mol]	97.86	87.88	67.61
C_2H_6 [% mol]	0.00	0.01	0.01
C_3H_8 [% mol]	0.00	0.00	0.00
C_4H_{10} [% mol]	0.00	0.00	0.00
C_5H_{12} [% mol]	0.00	0.00	0.00
N_2 [% mol]	2.13	12.11	32.38
$\Delta H_{vap,BOG}$ [kJ/kg]	495.00	436.00	341.00

For the purpose of this study the capacity of the unit is assumed by maintaining the philosophy of 95% LNG production and 5% fuel gas production. LNG product is assumed to be at -161°C and above atmospheric pressure about 70 mbar gauge.

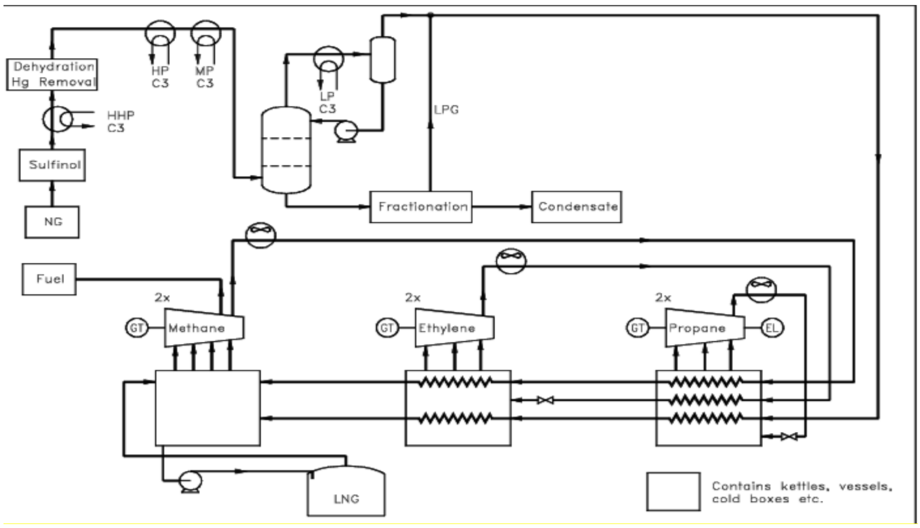
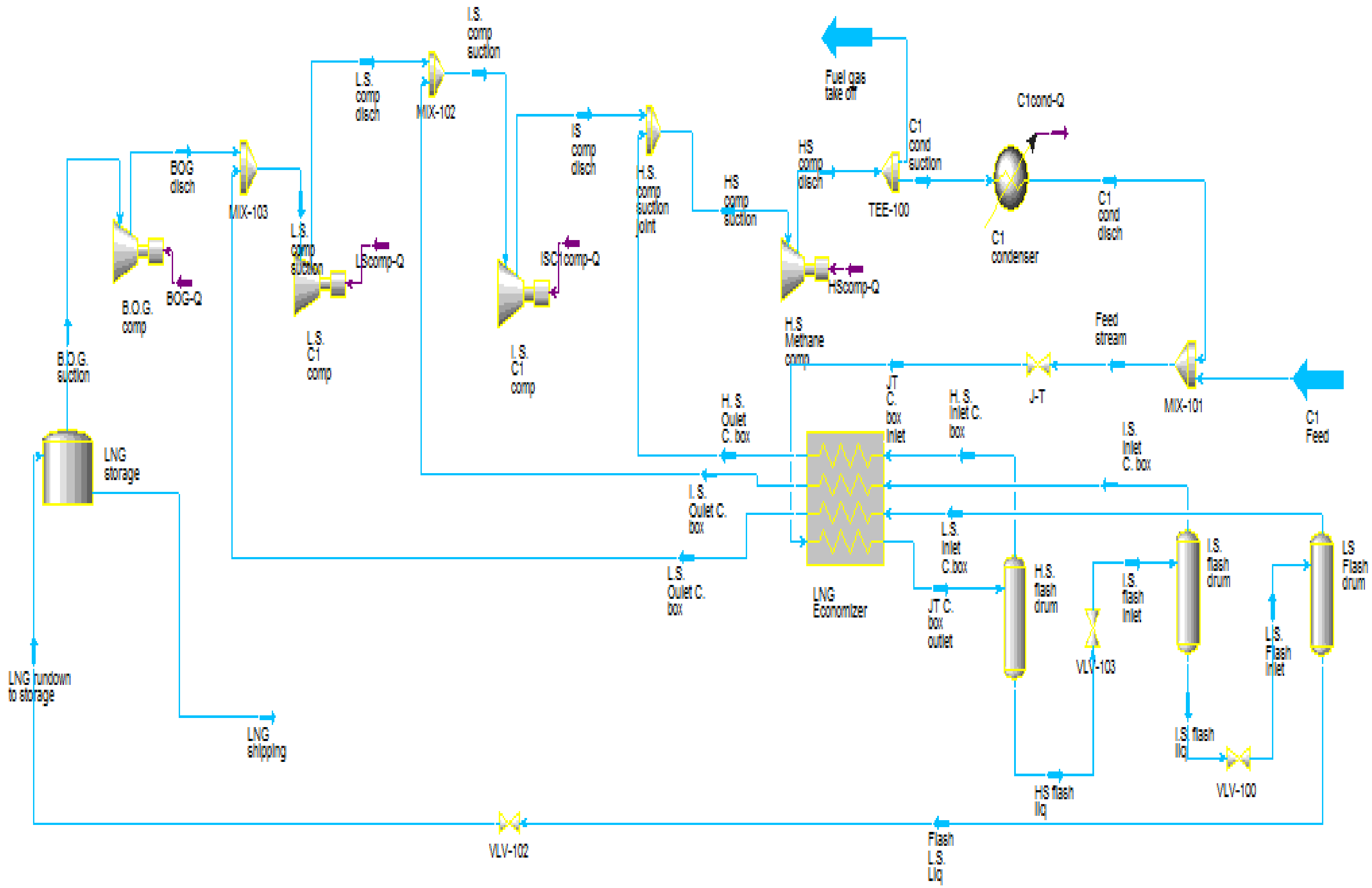


Figure 1: Simplified flowsheet of the cascade cycle (Vink, 1998)

Figure 1 illustrates an example of a COPOC flowsheet with 3 refrigerants cycles including propane, ethylene and methane. However the scope of this project is identified within the methane cycle. Figure 2 shows a detailed Hysys flowsheet of the methane cycle to be analysed focusing on the correlation of fuel and production streams efficiency.

Figure 2: HYSYS LNG back end process flowsheet draft 1



CHAPTER 2: LITERATURE REVIEW

2.1 THE CARNOT CYCLE

When working on refrigeration system designs the Carnot cycle theory is the reference the design although the efficiency of Carnot cycle is partially theoretical. The following pictures illustrate the Carnot cycle

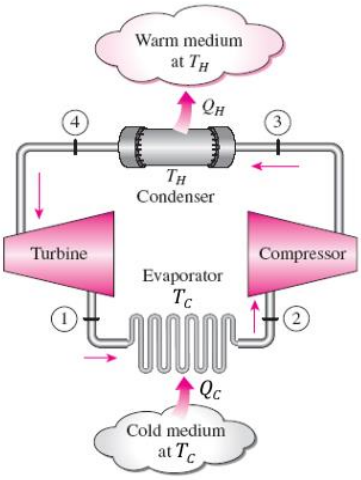


Figure 3: Picture extracted from Chemical engineering Thermodynamics lecture notes, Universiti Teknologi PETRONAS (2014)

In a continuous refrigeration process, the heat absorbed at a low temperature is continuously rejected to the surroundings at a higher temperature. Basically, a refrigeration cycle is a reversed heat-engine cycle. Refrigerators and heat engines operate on a Carnot cycle, consisting in this case of two isothermal steps in which heat $|Q_C|$ is absorbed at the lower temperature T_c and heat $|Q_H|$ is rejected at the higher temperature T_H, and two adiabatic steps. The cycle requires the addition of net work W to the system. Because ΔU of the working fluid is zero for the cycle, the first law is written as

$$W = |Q_H| - |Q_C| \qquad (2\text{-}1)$$

And the measure of the effectiveness of a refrigerator is its coefficient of performance ω,

Applicable to refrigeration operating on a Carnot cycle

$$\omega = \frac{T_C}{T_H - T_C} \qquad (2\text{-}2)$$

2.2 THE VAPOR-COMPRESSION CYCLE

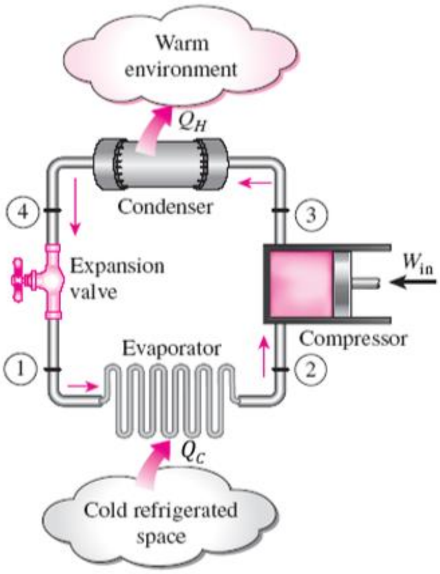


Figure 4: Vapor-compression refrigeration cycle, picture extracted from Chemical engineering Thermodynamics lecture notes, Universiti Teknologi PETRONAS (2014)

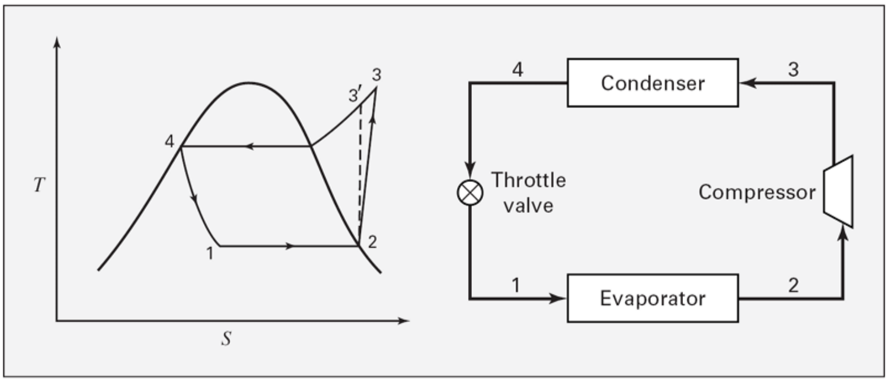


Figure 5: Vapor-compression refrigeration cycle T-S diagram, picture extracted from Chemical engineering Thermodynamics lecture notes, Universiti Teknologi PETRONAS (2014)

The vapor-compression refrigeration cycle is represented in Fig. 2, is the ideal model for refrigeration systems. Shown on the T S diagram are the four steps of the process. Unlike the reversed Carnot cycle Fig. 3, the refrigerant is vaporized completely before it is compressed and the turbine is replaced with a throttling device. A liquid evaporating at constant pressure (line 1→2) provides a means for heat absorption at a low constant temperature. The vapor produced is compressed to a higher pressure, and is then cooled and condensed at constant pressure with rejection of heat at a higher temperature level. Liquid from the condenser returns to its original pressure by an expansion process. In principle, this can be carried out in an expander from which work is obtained, but for practical reasons is usually accomplished by throttling through a partly open valve. The pressure drop in this irreversible process results from fluid friction in the valve, at constant enthalpy. Line (4 → 1) represents this throttling process. The dashed line (2 → 3') is the path of isentropic compression. Line (2 →3) represents the actual compression process, slopes in the direction of increasing entropy, reflecting inherent irreversibilities (Smith et al., 2005).

On the basis of a unit mass of fluid, the equation for the heat absorbed in the evaporator and the heat rejected in the condenser are

$$|Q_C| = H_2 - H_1 \qquad \text{and} \qquad |Q_H| = H_3 - H_4 \tag{2-3}$$

The work of compression is simply: $W = H_3 - H_2$, and the coefficient of performance is

$$\omega = \frac{H_2 - H_1}{H_3 - H_2} \tag{2-4}$$

To design the evaporator, compressor, condenser, and auxiliary equipment one must know the rate of circulation of refrigerant $\dot{m}$. This is determined from the rate of heat absorption in the evaporator by the equation (Smith et al., 2005):

$$\dot{m} = \frac{|Q_C|}{H_2 - H_1} \tag{2-5}$$

2.3 LNG LIQUEFACTION

The liquefaction process is the key element of an LNG plant. Liquefaction is based on a refrigeration cycle, where a refrigerant by means of successive expansion and compression, transport heat from the process side to where the natural gas is (Xiuli, 2009).

Xiuli (2009) points out that “The basic principles for cooling and liquefying the gas using refrigerants, involve matching as closely as possible the cooling/heating curves of process gas and refrigerant. These principles result in a more efficient thermodynamic process, requiring less power per unit of LNG produced, and they apply to all liquefaction processes”.

The following is the referenced LNG cooling curve mentioned above:

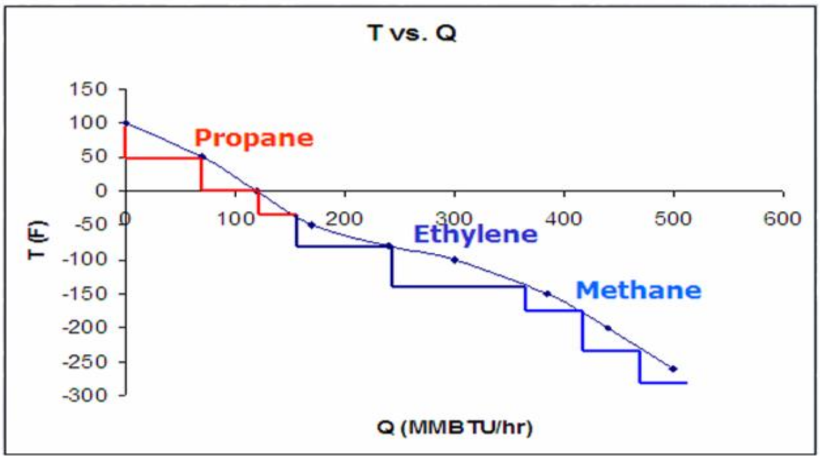


Figure 6: Phillips Cascade LNG cooling curve (Muhannad et al., 2013)

Muhannad et al. (2013) point out that: The crossing blue line with dots in Fig 3 represents how natural gas is 100% efficient (ideal) cooling process should behave; the block drawing under that blue line represents the refrigeration in the cascade process and the area between them represents the heat loss by the system.

In the book “LNG: basics of liquefied natural gas” by Stanley et al. (2007) the COPOC process similar to Figure 1, is described by the authors as a process having three refrigeration loops using propane, ethylene, and methane as refrigerants. The propane and ethylene are two separate closed-loop refrigerant systems while the methane is an open-loop refrigerant system. This loop is open to the high methane content feed stream (condensed feed gas). The methane loop works by flashing the condensed, high-pressure process stream to progressively lower pressures in the stages(high stage, intermediate stage and lower stage), each with recompression and recirculation of the flashed vapors.

In the paper “The Phillips optimized cascade LNG process: a quarter century of improvements” by Andres D. L. (1996), the author ended his paper with a list of special features of the Philips optimized cascade LNG process and I want to highlight the ones that apply to the objectives of this project:

- *Nitrogen removal*, removal of nitrogen from the feed gas minimizes the power requirement per billion Btu of product and lowers marine transportation cost. Nitrogen is removed in a unique rejection scheme. And fuel is provided in a manner that eliminates the need of dedicated fuel gas unit for the compressors;

- *Vapor recovery*, storage tank vapor is returned to the methane refrigeration system to recover both the vapor and its refrigeration. No especial equipment other than a vapor blower in the case of this project and in most COPOC processes a B.O.G compressor is used for this purpose and the vapor is processed through existing liquefaction equipment;

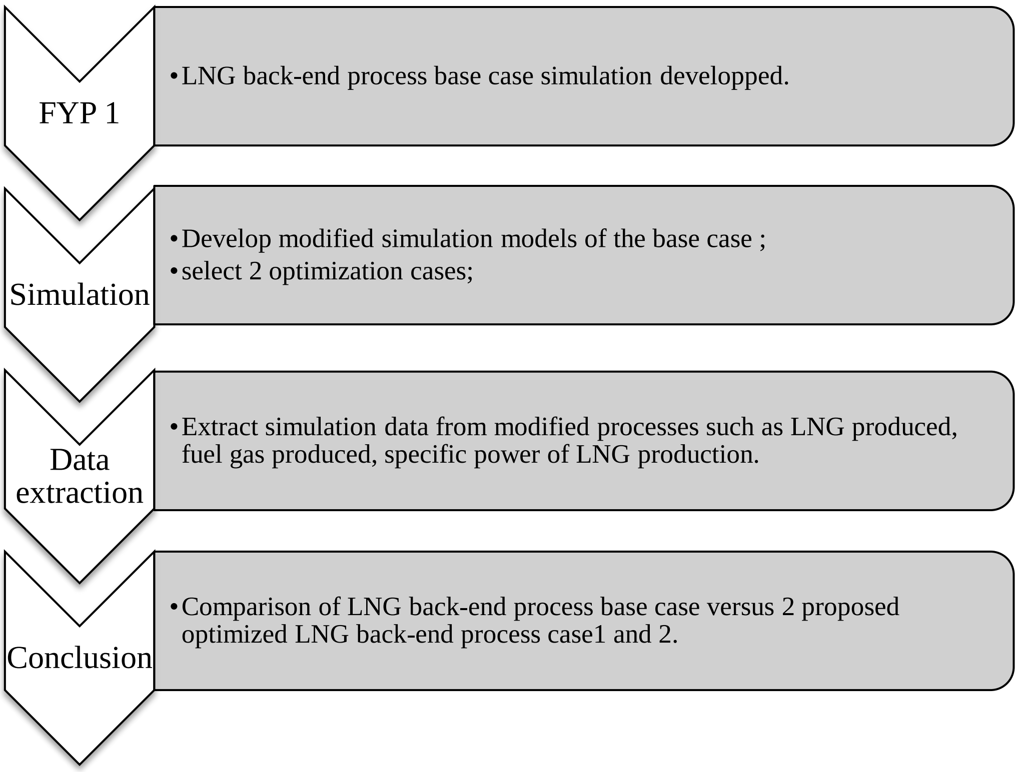
- *Ease of operation*, COPOC processes utilizes pure component refrigerants of essentially constant molecular weight. This fact greatly simplifies the operation of the compression systems and makes the COPOC processes one the simplest operating design.

Castillo et al. (2012) in their paper "Conceptual analysis of the precooling stage for LNG processes" made a comparison between different precooling cycles for LNG processes which were carried out through computational simulation using Aspen HYSYS. The aim of the paper was to provide future development with a clear idea of the technical advantages and disadvantages involved in the selection of the process for the precooling cycle. The results of the research revealed that, 3 stages propane precooled was found to be the most energetically efficient among studied cases, even better than a two stage mixed refrigerant process (C2/C3) for both climate conditions, warm (25°C) and cold (6°C) respectively. However, due to the reduced power share that may be reached with a propane cycle temperature restriction, the mixed refrigerant precooling cycle is the preferred alternative under a cold climate conditions.

Boil off gases (BOG) in LNG storage contribute directly to the nitrogen addition in the methane cycle and therefore it represents the fuel gas produced as the strategy to reject back the nitrogen. Querol et al. (2010), studied the behaviour of BOG in an LNG process in their paper "Boil off gas (BOG) management in Spain liquid natural gas (LNG) terminals". The paper states that most common LNG tank installed in Spain correspond to a fully contained system (tank inside a tank) with storage capacity of 150,000m^3. Those thanks have been designed to maintain the storage temperature (-163) taking into account some liquid vaporization, limited to daily maximum of 0.05% of the stored liquid. The BOG produced must be removed to maintain the tank desired pressure. Furthermore, the paper confirms that evaluation of than BOG is usually done considering LNG as methane. In addition, the Querol et al. (2010) confirms that not tanks but LNG piping system also receive heat from the outside which will contribute in the BOG generation. This makes it necessary a constant LNG flow through the system to keep the lines at low and required temperature around -160°C and ready for use whenever needed.

CHAPTER 3: METHODOLOGY

3.1 PROJECT FLOWCHART



The above flowchart illustrates briefly the flow methodology starting with the base case and resulting in two new optimized cases, followed by a comparison study among the three cases that will be addressed in detail in the next sections.

Final year project II was started with the developed LNG back end process fig. 12 which is the base case of this study. The base case simulation process for this study features the following characteristics:

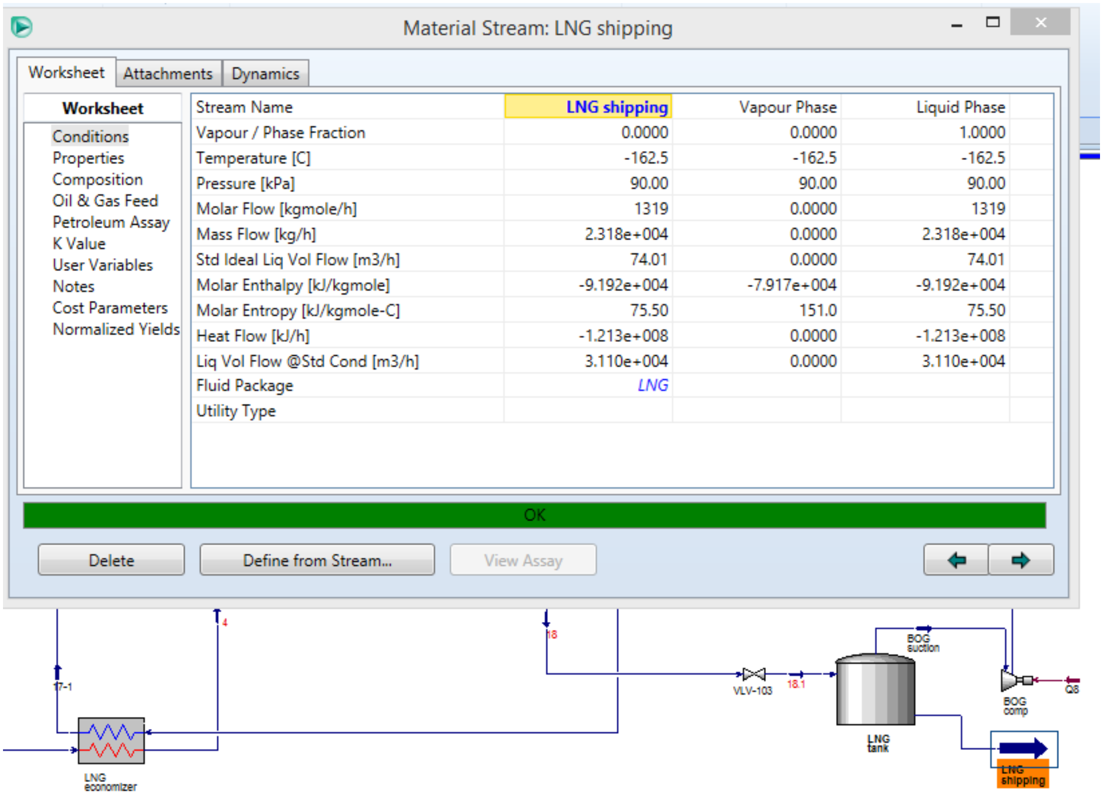
Table 2: Base case performance features

Parameters	
Condensed feed, kg/hr	50000
Feed’s Nitrogen content, mole%	0.0050
Produced LNG, kg/hr	13500
LNG’s nitrogen content	0.0003
Produced fuel gas, kg/hr	1825
Specific power of production, KJ/kg	903

As per the project flowchart the base case is to be modified by exploring the opportunity of increasing LNG production and reducing fuel gas production. Two proposed optimized cases are to be proposed.

EXTRACTION OF DATA FROM ASPEN HYSYS V8.5

Upon following the procedure to develop an optimized simulation case as presented in the fig.13, next will be to extract and analyse data from the simulation that is used to confirm the efficiency of the design simulation to be proposed. For instance, the path of reducing number of stages was followed and the new optimized simulation case produces 9688 kg/hr of LNG and production fuel is reduced from 1825 kg/hr (base case) to 1338 kg/hr (optimized case 1) as seen in the screen shots of the Aspen HYSYS simulation:



Stream Name	LNG shipping	Vapour Phase	Liquid Phase
Vapour / Phase Fraction	0.0000	0.0000	1.0000
Temperature [C]	-162.5	-162.5	-162.5
Pressure [kPa]	90.00	90.00	90.00
Molar Flow [kgmole/h]	1319	0.0000	1319
Mass Flow [kg/h]	2.318e+004	0.0000	2.318e+004
Std Ideal Liq Vol Flow [m3/h]	74.01	0.0000	74.01
Molar Enthalpy [kJ/kgmole]	-9.192e+004	-7.917e+004	-9.192e+004
Molar Entropy [kJ/kgmole-C]	75.50	151.0	75.50
Heat Flow [kJ/h]	-1.213e+008	0.0000	-1.213e+008
Liq Vol Flow @Std Cond [m3/h]	3.110e+004	0.0000	3.110e+004
Fluid Package	LNG		
Utility Type			

Figure 7: Optimized simulation proposal 1 rundown stream data

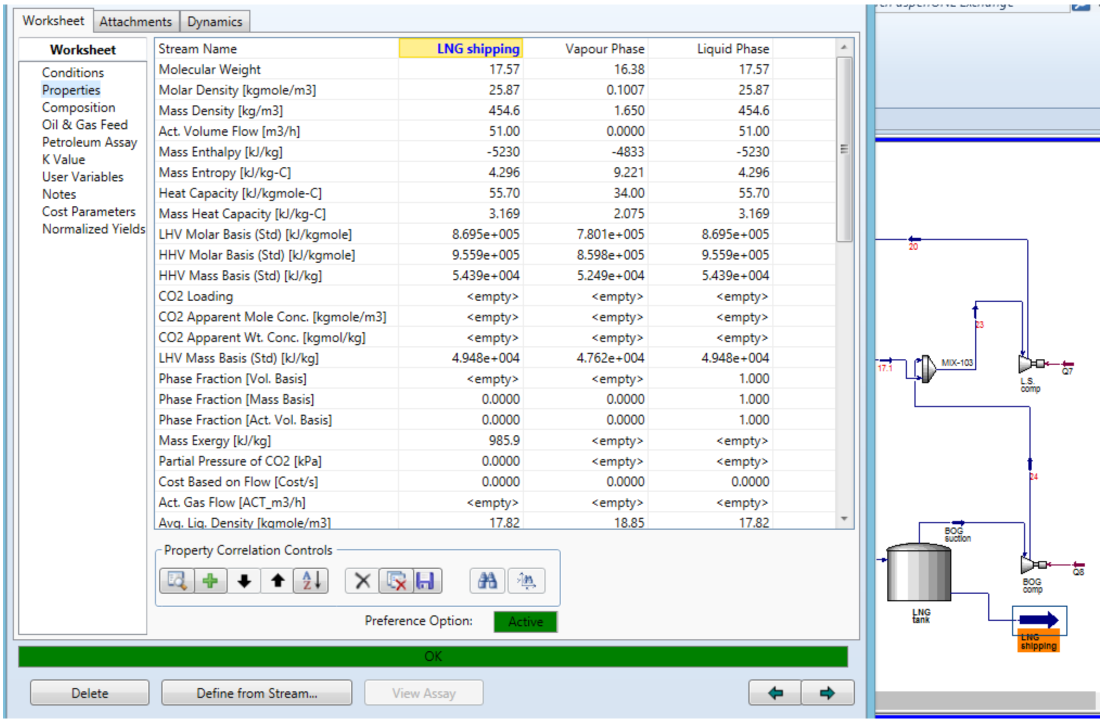


Figure 8: Optimized simulation proposal 1 rundown properties

In the same way by clicking the fuel gas stream the same information can be extracted such as produced fuel gas flowrate which is 1338 kg/hr for this simulation, the properties, etc.

In an efficient way is possible to extract overall data of the process in a single table to better analyse the overall performance of the unit. A right click with the mouse on the simulation screen brings up this menu options:

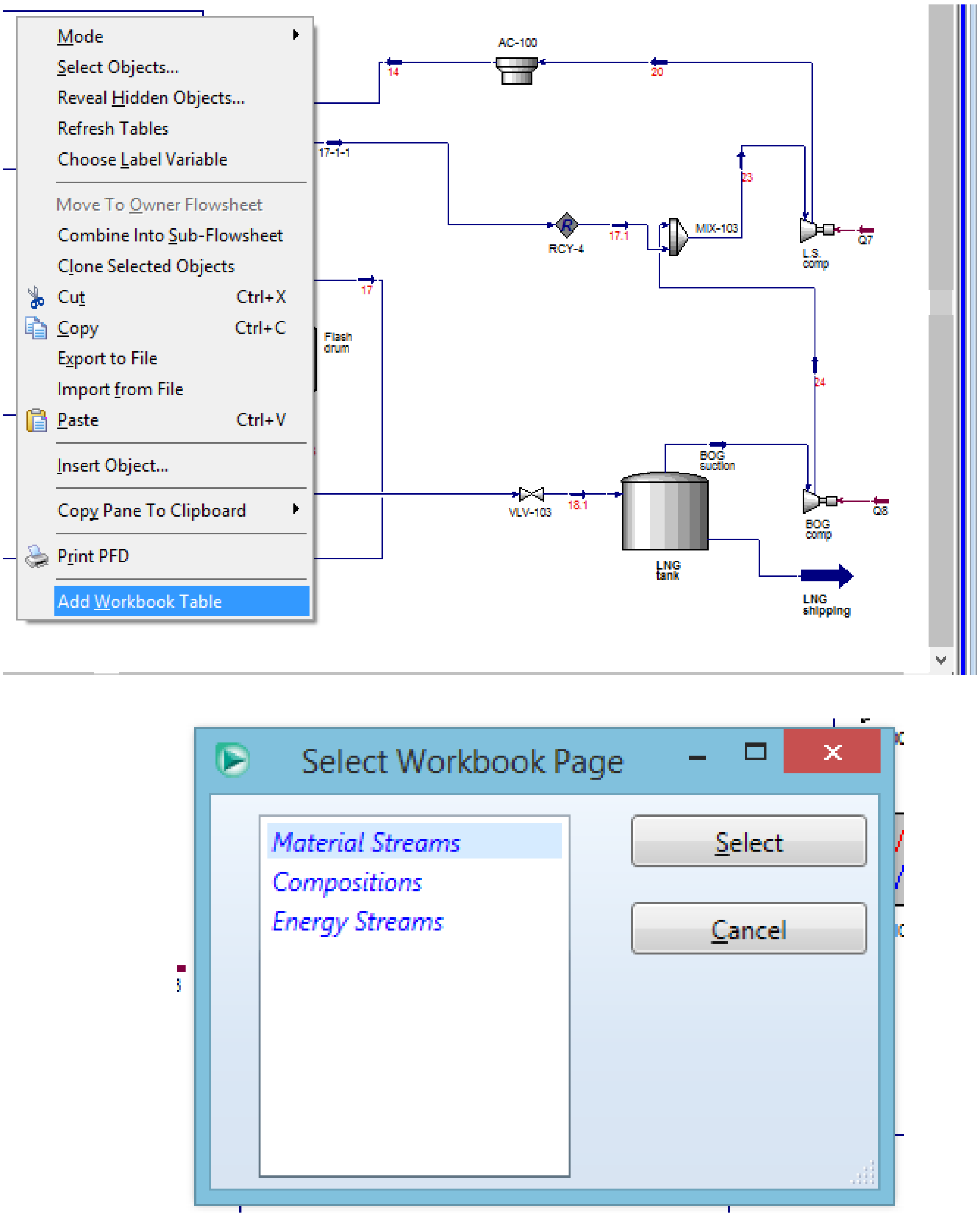
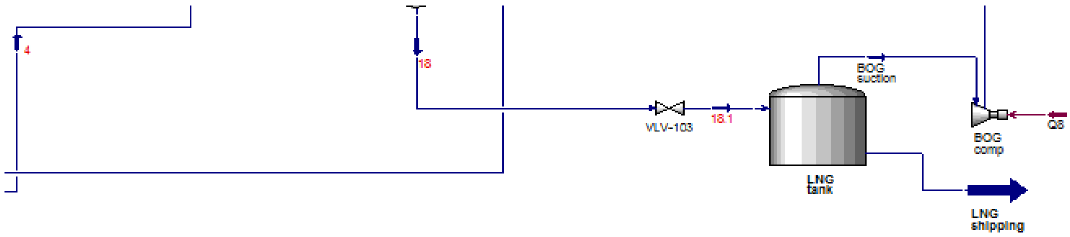


Figure 9: Overall process data extraction procedure on Aspen HYSYS

And menu "add workbook table" brings up 3 options of overall process data such as the all the material streams table, or the entire streams compositions table, and the table of all the energy stream is the third option and any of those table will be displayed as in the below screen shot:



Material Streams										
		Condensed feed gas	2	HS comp disch	HP fuel gas	C1 recycle	8	13	14	15
Vapour Fraction		0.0000	0.0000	1.0000	1.0000	1.0000	0.5153	1.0000	1.0000	0.0000
Temperature	C	-90.00	-87.04	135.7	135.7	135.7	-152.4	30.00	124.4	-88.25
Pressure	kPa	4500	4500	4550	4550	4550	200.0	1600	400.0	4500
Molar Flow	kgmole/h	1402	2926	1604	80.18	1523	2926	1604	1604	1523
Mass Flow	kg/h	2.458e+004	5.000e+004	2.675e+004	1338	2.542e+004	5.000e+004	2.675e+004	2.675e+004	2.542e+004
Liquid Volume Flow	m3/h	78.32	158.5	84.32	4.216	80.11	158.5	84.32	84.32	80.11
Heat Flow	kJ/h	-1.216e+008	-2.427e+008	-1.077e+008	-5.383e+006	-1.023e+008	-2.449e+008	-1.138e+008	-1.078e+008	-1.211e+008
		17	18	20	BOG suction	LNG shipping	18.1	23	24	LNG R
Vapour Fraction		1.0000	0.0000	1.0000	1.0000	0.0000	0.0696	1.0000	1.0000	0.0000
Temperature	C	-152.4	-152.4	156.9	-162.5	-162.5	-162.5	18.93	-158.5	-88.25
Pressure	kPa	200.0	200.0	400.0	90.00	90.00	90.00	100.0	100.0	4500
Molar Flow	kgmole/h	1508	1418	1604	98.72	1319	1418	1604	98.72	1524
Mass Flow	kg/h	2.520e+004	2.480e+004	2.675e+004	1617	2.318e+004	2.480e+004	2.675e+004	1617	2.542e+004
Liquid Volume Flow	m3/h	79.20	79.25	84.32	5.238	74.01	79.25	84.32	5.238	80.13
Heat Flow	kJ/h	-1.158e+008	-1.291e+008	-1.056e+008	-7.816e+006	-1.213e+008	-1.291e+008	-1.141e+008	-7.803e+006	-1.212e+008
		17.1	3	4	14.1	17-1	17-1-1	14.1.1		
Vapour Fraction		1.0000	0.3399	0.2148	1.0000	1.0000	1.0000	1.0000		
Temperature	C	30.00	-107.7	-108.5	137.3	-112.1	30.00	3.108		
Pressure	kPa	100.0	2000	2000	1600	100.0	100.0	400.0		
Molar Flow	kgmole/h	1505	2926	2926	1604	1508	1508	1604		
Mass Flow	kg/h	2.514e+004	5.000e+004	5.000e+004	2.675e+004	2.520e+004	2.520e+004	2.675e+004		
Liquid Volume Flow	m3/h	79.08	158.5	158.5	84.32	79.20	79.20	84.32		
Heat Flow	kJ/h	-1.063e+008	-2.427e+008	-2.449e+008	-1.071e+008	-1.137e+008	-1.064e+008	-1.151e+008		

Figure 10: Optimized simulation proposal 1 material stream data extraction

Another important information these end flash processes is the performance of the multiple stream heat exchangers. The step of converging the LNG heat exchangers is crucial and usually the last to converge in the whole simulation. The wrong minimum approach temperature will result in temperature cross and by adjusting the inlet and outlet temperatures of LNG heat exchangers and maintaining a positive minimum approach temperature difference between streams should converge the heat exchanger. This knowledge is supported by H.M. Chang et al. (June, 2012) paper “Effect of multi-stream heat exchanger on performance of natural gas liquefaction with mixed refrigerant” and the paper states that –a simple and widely used method in process simulation is to assume that two hot stream (H and F) have the same temperature approach between hot and cold streams:

$$T_H - T_L = T_F - T_L \geq \Delta T_{min}$$

In the performance menu of the LNG heat exchanger the overall performance can be extracted and evaluated from Temperature vs Heat flow plot which illustrates the minimum approach temperature between the hot and cold streams as shown in the next screen shot:

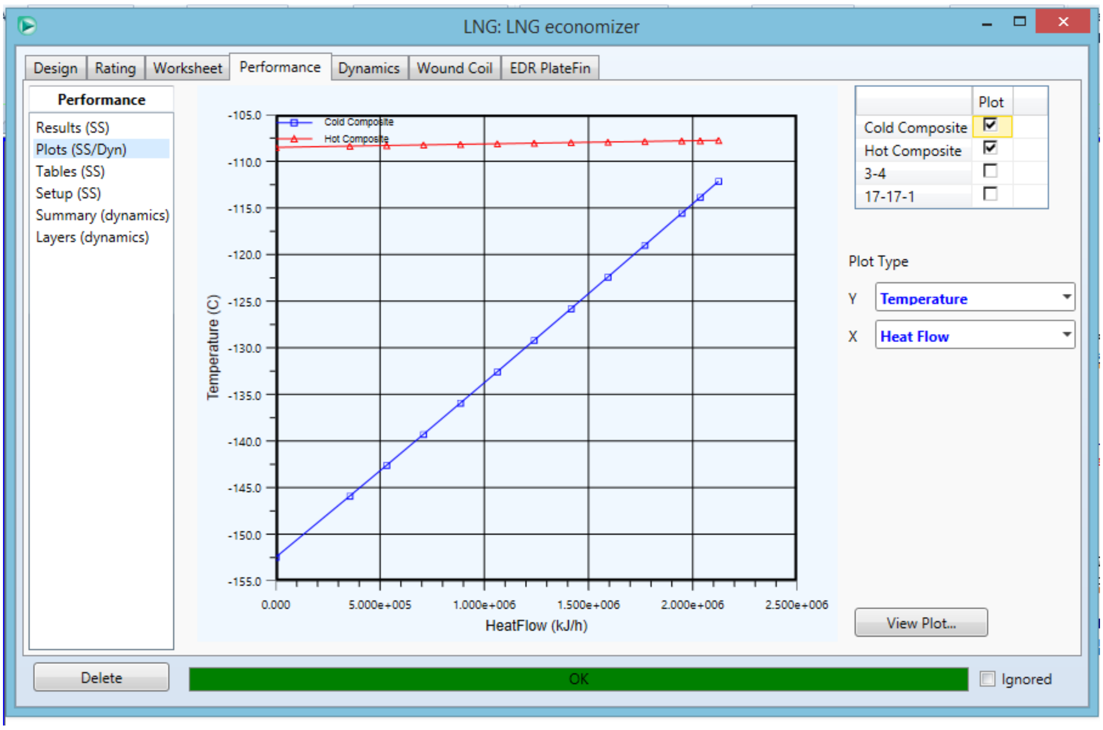


Figure 11: Optimized simulation proposal 1 LNG economizer temperature performance

The above screen shots are illustrations on how process information, process deliverables are extracted from Aspen HYSYS simulation and analysis of the performance can be conducted and different simulations performance can be compared respectively.

3.2 GANTT CHART AND KEY MILESTONE

Table 3: Timelines for FYP 2

No.	Detail/Week	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
1	Project work continues															
	✓ Met with supervisor and project planning discussed for FYP2															
	✓ Develop modified HYSYS models compared to base case(FYP1)															
	✓ Analysis of cases comparison and report writing															
	✓ Identify a fuel and production efficient case if any															
2	Submission of Progress report															
3	Project work continues															
	✓ Completion of Final draft report															
4	Pre-SEDEX															
5	Submission of Draft Final Report															
6	Submission of Dissertation (Soft bound)															
7	Submission of Technical paper															
8	Oral presentation															
9	Submission of Project Dissertation (Hard bound)															

Suggested process

Suggested milestone

Process completion

3.3 RESEARCH METHODOLOGY

BASE CASE PROCESS DESCRIPTION

The base case simulation of this project was simulated during the FYP1. Therefore for FYP2 we are recalling the process description to define the start point of this project. Modifications are to be applied with the goal to optimize the base case developing 2 new optimized proposal models with deliverables parameters as increase of LNG production, reduction of fuel gas production and less power required for production yielding in more efficient LNG back end processes.

The base case HYSYS flowsheet illustrated in fig. 12, the process starts with a condensed feed gas assumed to come out of the ethylene cycle feed condensers, the condensed feed gas is received at a temperature of -90°C and elevated pressure of about 45 bar gauge, this line is joint with 5 degree warmer LNG recycled at about -85°C and at almost same pressure, the joint lines yield the stream 2 at temperature of about -83°C and 45 bar gauge. Stream 2 is then expanded under a Joule-Thomson effect valve (JT) breaking down de pressure from 45 barg to 17 barg and cooled to about -111°C, yielding stream 3. Stream 3 will then undergo further precooling on the LNG economizer and yield stream 4 at 15 barg and about -114.5°C. At this point, as indicated in table 2, 15 barg defines the High stage of methane cycle for this simulation. 15 barg was predefined as first stage cooling conditions and Aspen HYSYS has automatically calculated the temperature for this stage at -114.5°C. Stream 4 continues as stream 5 with no changes, stream 5 then undergoes a flashing process where vapors are flashed with the ultimate intention to reject nitrogen. H.S. flash drum then rejects nitrogen by flashing the vapors of stream 5, and the vapors will be warmed up in the LNG economizer before being sent to H.S. compressor at almost ambient temperature. The liquid portion from the H.S. flash drum stream 7 then undergoes further subcooling in the next drum. Pressure is again breakdown to 4 barg as predefined for I.S. stage.

HYSYS has calculated the temperature for this stage (I.S.) with is -141°C; same flashing process occurs here, nitrogen is rejected via the vapors that will be warmed up in the LNG economizer before going to the I.S. compressor at almost ambient temperature. Liquid fraction from the I.S. drum now stream 10, is subcooled at about -160°C and the pressure has been breakdown to 1 barg in the L.S. flash drum, yielding the desired conditions defined for the LNG product. Note that these three stages, H.S., I.S. and L.S. are the core of the liquefaction process, the three stages consecutively subcool the condensed feed and at the same time reject the extra nitrogen that is added in the system in the storage vessel, LNG tank. The LNG economizer on the other hand takes and recovers energy for the hot stream from the cold streams. LNG is stored at about -162.2°C in this simulation.

All vapor streams including 6, 9, 17 and the B.O.G. are all recovered in the compressor three stages respectively and a fraction of 5% of the final stage, H.S. compressor discharge is separated for fuel gas production. Detailed information of this process parameters are illustrated in table 4.

Overall with a condensed feed of 50,000kg in stream 2, the simulated process consumes about 903KJ to produce 1Kg of LNG, 13500kg/h of LNG production. In addition the 3 compressor stages which in fact represent a multistage compressor consume overall 3.3MW of power.

Figure 12: Base case simulation (COPOC process)

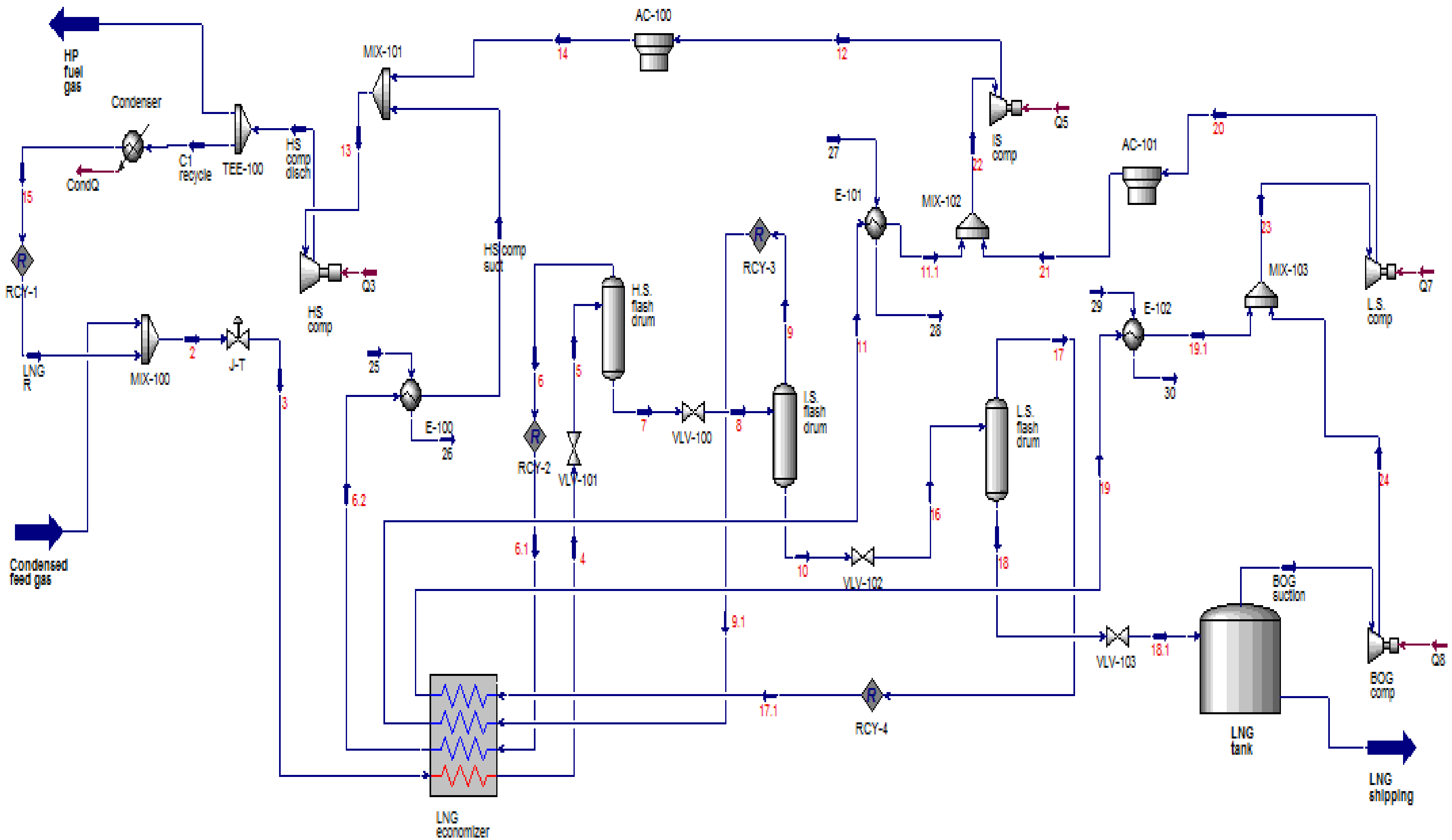


Table 4: Base case study material balance extracted from HYSYS

Material Streams												
		Condensed feed gas	2	3	4	5	6	7	6.2	HS comp disch	HP fuel gas	C1 recycle
Vapour Fraction		0.0000	0.4303	0.6327	0.6089	0.6089	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000
Temperature	C	-90.00	-82.88	-111.1	-114.4	-114.4	-114.4	-114.4	-112.0	144.2	144.2	144.2
Pressure	kPa	4500	4500	1700	1500	1500	1500	1500	1470	4550	4550	4550
Molar Flow	kgmole/h	880.4	2975	2975	2975	2975	1812	1164	1811	2212	110.6	2101
Mass Flow	kg/h	1.544e+004	5.000e+004	5.000e+004	5.000e+004	5.000e+004	2.998e+004	2.002e+004	2.996e+004	3.649e+004	1825	3.467e+004
Liquid Volume Flow	m3/h	49.19	160.1	160.1	160.1	160.1	95.80	64.30	95.74	117.1	5.855	111.2
Heat Flow	kJ/h	-7.634e+007	-2.417e+008	-2.417e+008	-2.424e+008	-2.424e+008	-1.403e+008	-1.021e+008	-1.400e+008	-1.505e+008	-7.525e+006	-1.430e+008
		8	9	10	11	12	13	14	15	16	17	18
Vapour Fraction		0.2305	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.1403	1.0000	0.0000
Temperature	C	-141.0	-141.0	-141.0	-112.0	168.9	30.00	30.14	-85.27	-160.9	-160.9	-160.9
Pressure	kPa	400.0	400.0	400.0	370.0	1500	1470	1500	4500	100.0	100.0	100.0
Molar Flow	kgmole/h	1164	268.2	895.5	268.8	401.2	2212	401.2	2101	895.5	125.6	769.9
Mass Flow	kg/h	2.002e+004	4382	1.564e+004	4392	6533	3.649e+004	6533	3.467e+004	1.564e+004	2031	1.361e+004
Liquid Volume Flow	m3/h	64.30	14.25	50.05	14.28	21.35	117.1	21.35	111.2	50.05	6.708	43.34
Heat Flow	kJ/h	-1.021e+008	-2.117e+007	-8.098e+007	-2.094e+007	-2.726e+007	-1.598e+008	-2.949e+007	-1.659e+008	-8.098e+007	-1.011e+007	-7.087e+007
		19	20	21	22	BOG suction	LNG shipping	18.1	23	24	LNG R	6.1
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0086	1.0000	1.0000	1.0000	1.0000
Temperature	C	-112.0	152.1	28.15	28.00	-162.2	-162.2	-162.2	16.32	-158.2	-85.26	-114.4
Pressure	kPa	100.0	400.0	400.0	370.0	90.00	90.00	90.00	100.0	100.0	4500	1500
Molar Flow	kgmole/h	125.8	132.4	132.4	401.2	6.596	763.3	769.9	132.4	6.596	2095	1811
Mass Flow	kg/h	2035	2141	2141	6533	106.5	1.350e+004	1.361e+004	2141	106.5	3.456e+004	2.996e+004
Liquid Volume Flow	m3/h	6.718	7.071	7.071	21.35	0.3524	42.99	43.34	7.071	0.3524	110.9	95.74
Heat Flow	kJ/h	-9.919e+006	-9.170e+006	-9.811e+006	-2.944e+007	-5.319e+005	-7.033e+007	-7.087e+007	-9.860e+006	-5.310e+005	-1.654e+008	-1.402e+008
		9.1	17.1	25	26	HS comp suct	27	28	11.1	29	30	19.1
Vapour Fraction		1.0000	1.0000	1.0000	0.9154	1.0000	1.0000	0.9886	1.0000	1.0000	0.9949	1.0000
Temperature	C	-141.0	-160.9	100.0	99.96	30.00	100.0	99.96	28.00	100.0	99.96	25.00
Pressure	kPa	400.0	100.0	101.3	101.3	1470	101.3	101.3	370.0	101.3	101.3	100.0
Molar Flow	kgmole/h	268.8	125.8	2775	2775	1811	2775	2775	268.8	2775	2775	125.8
Mass Flow	kg/h	4392	2035	5.000e+004	5.000e+004	2.996e+004	5.000e+004	5.000e+004	4392	5.000e+004	5.000e+004	2035
Liquid Volume Flow	m3/h	14.28	6.718	50.10	50.10	95.74	50.10	50.10	14.28	50.10	50.10	6.718
Heat Flow	kJ/h	-2.121e+007	-1.013e+007	-6.643e+008	-6.739e+008	-1.303e+008	-6.643e+008	-6.656e+008	-1.963e+007	-6.643e+008	-6.649e+008	-9.329e+006

BASE CASE OPTIMIZATION METHODOLOGY

In order to develop optimized simulation models, two approaches have been consider, one by reducing the number of sub-cooling stages and two by increasing the number of sub-cooling stages. The flash end system is a mechanism mainly to reject nitrogen content via the flashing process and ultimately use the nitrogen rich natural gas as fuel gas. The following flowchart was followed to develop the proposed optimized simulation models which will be discussed in detail in the results and discussion section.

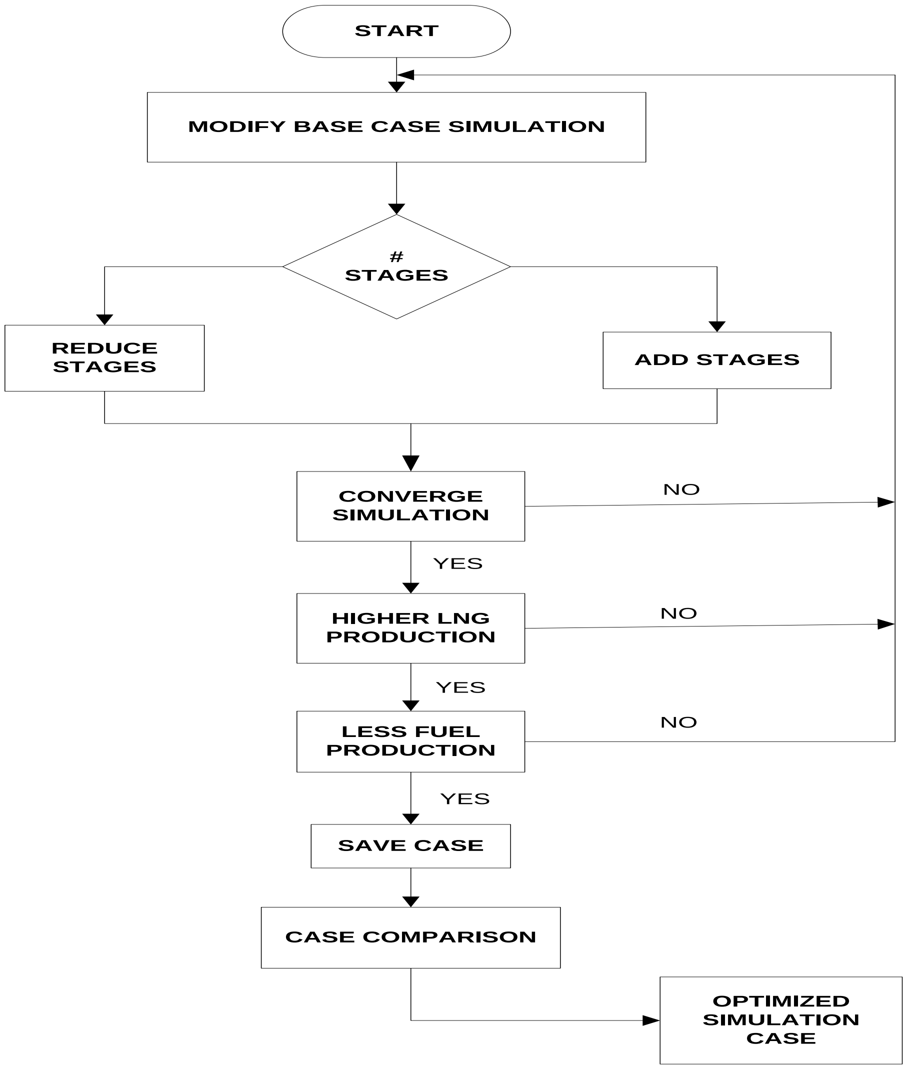


Figure 13: Base case modification flowchart to optimized cases

CHAPTER 4: RESULTS AND DISCUSION

Optimized cases were developed following the two approaches presented in the simulation modification flowchart presented in fig. 13 and, 2 new cases have been developed. As the objective of this study is to explore opportunity to increase LNG production and reduced the fuel gas production analysing as well the specific power required to produced 1Kg of LNG have all been considered in the optimized simulations.

4.1 OPTIMIZED SIMULATION CASE 1

Referring to fig. 14 in the next page, the front of this flowsheet starts with a Joule Thompson effect breaking down the front pressure from 45 barg in stream 2 to 20 barg on stream 3. The condensed feed is 3 degrees Celsius hotter than in the base case fig. 12 at that point of the process. The process is pretty much similar up to before the condensed feed undergoes further cooling in the LNG economizer 1. The first big modification is the LNG economizer, for this optimization proposal fig. 14 the LNG economizers only have a single hot stream and single cold stream as a result of the biggest modification of the process which is the reduction of multiple subcooling stages into a single stage. What characterizes optimization simulation proposal 1 is that is a single stage cooling process, therefore a single thus bigger flash drum does the work of rejecting the nitrogen in the system. Optimization simulation proposal 1 handles it single flashed vapor stream in the low pressure compressor (L.P. Comp.) and 2 extra booster compressors are used in order to boost up the pressure back to feed pressure of 45 bars. This optimized proposal 1 produces 23180 kg/hr of LNG at -162.5°C. The storage condition pretty much the same. Optimization proposal 1 is simpler but with bigger equipment.

Figure 14: LNG back-end process Optimized simulation CASE 1

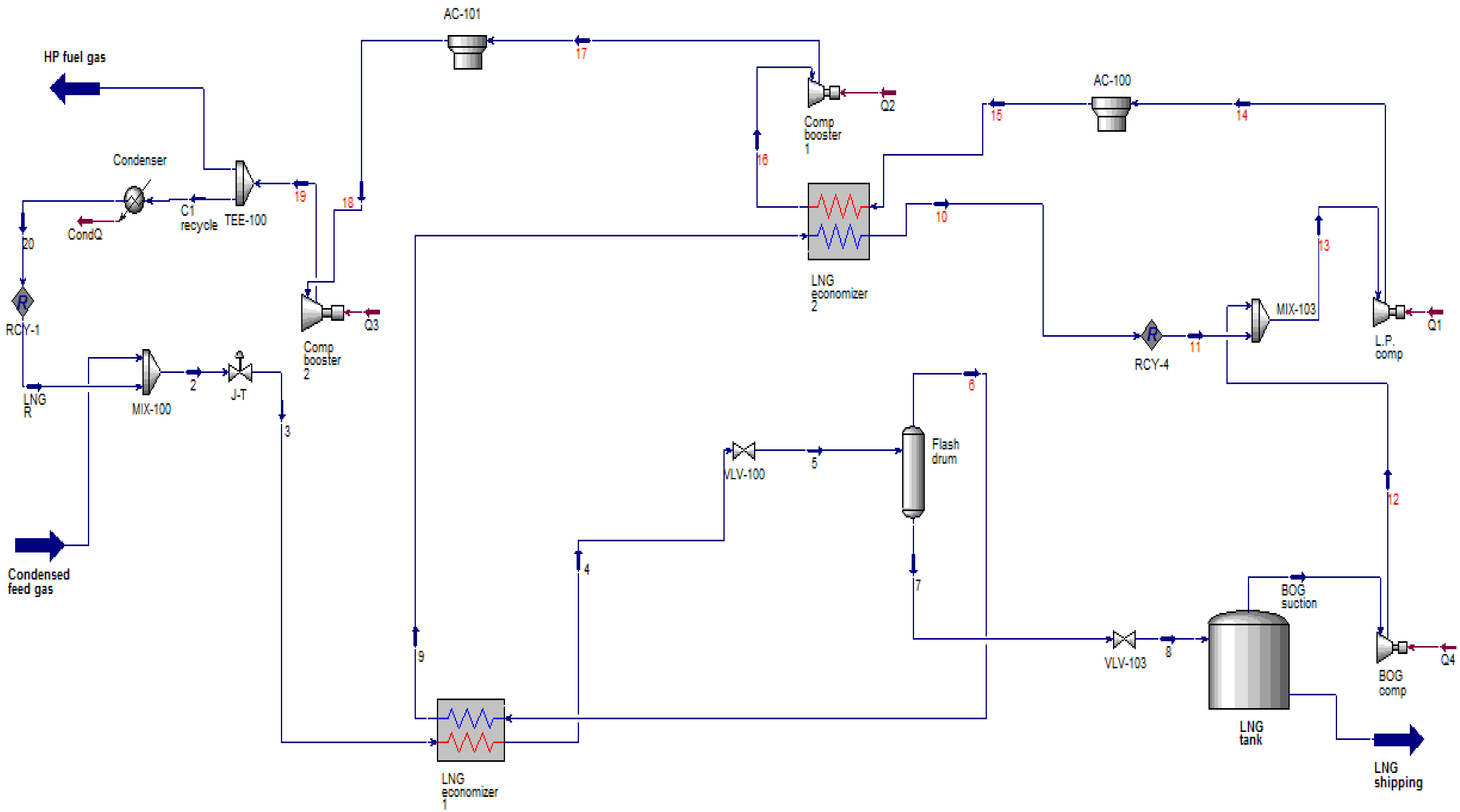


Table 5: CASE 1 material balance extracted from HYSYS

Material Streams										
		Condensed feed gas	2	19	HP fuel gas	C1 recycle	5	18	15	20
Vapour Fraction		0.0000	0.0000	1.0000	1.0000	1.0000	0.5153	1.0000	1.0000	0.0000
Temperature	C	-90.00	-87.04	135.7	135.7	135.7	-152.4	30.00	124.4	-88.25
Pressure	kPa	4500	4500	4550	4550	4550	200.0	1600	400.0	4500
Molar Flow	kgmole/h	1402	2926	1604	80.18	1523	2926	1604	1604	1523
Mass Flow	kg/h	2.458e+004	5.000e+004	2.675e+004	1338	2.542e+004	5.000e+004	2.675e+004	2.675e+004	2.542e+004
Liquid Volume Flow	m3/h	78.32	158.5	84.32	4.216	80.11	158.5	84.32	84.32	80.11
Heat Flow	kJ/h	-1.216e+008	-2.427e+008	-1.077e+008	-5.383e+006	-1.023e+008	-2.449e+008	-1.138e+008	-1.078e+008	-1.211e+008
		6	7	14	BOG suction	LNG shipping	8	13	12	LNG R
Vapour Fraction		1.0000	0.0000	1.0000	1.0000	0.0000	0.0696	1.0000	1.0000	0.0000
Temperature	C	-152.4	-152.4	156.9	-162.5	-162.5	-162.5	18.93	-158.5	-88.25
Pressure	kPa	200.0	200.0	400.0	90.00	90.00	90.00	100.0	100.0	4500
Molar Flow	kgmole/h	1508	1418	1604	98.72	1319	1418	1604	98.72	1524
Mass Flow	kg/h	2.520e+004	2.480e+004	2.675e+004	1617	2.318e+004	2.480e+004	2.675e+004	1617	2.542e+004
Liquid Volume Flow	m3/h	79.20	79.25	84.32	5.238	74.01	79.25	84.32	5.238	80.13
Heat Flow	kJ/h	-1.158e+008	-1.291e+008	-1.056e+008	-7.816e+006	-1.213e+008	-1.291e+008	-1.141e+008	-7.803e+006	-1.212e+008
		11	3	4	17	9	10	16		
Vapour Fraction		1.0000	0.3399	0.2148	1.0000	1.0000	1.0000	1.0000		
Temperature	C	30.00	-107.7	-108.5	137.3	-112.1	30.00	3.108		
Pressure	kPa	100.0	2000	2000	1600	100.0	100.0	400.0		
Molar Flow	kgmole/h	1505	2926	2926	1604	1508	1508	1604		
Mass Flow	kg/h	2.514e+004	5.000e+004	5.000e+004	2.675e+004	2.520e+004	2.520e+004	2.675e+004		
Liquid Volume Flow	m3/h	79.08	158.5	158.5	84.32	79.20	79.20	84.32		
Heat Flow	kJ/h	-1.063e+008	-2.427e+008	-2.449e+008	-1.071e+008	-1.137e+008	-1.064e+008	-1.151e+008		

4.2 OPTIMIZED SIMULATION CASE 2

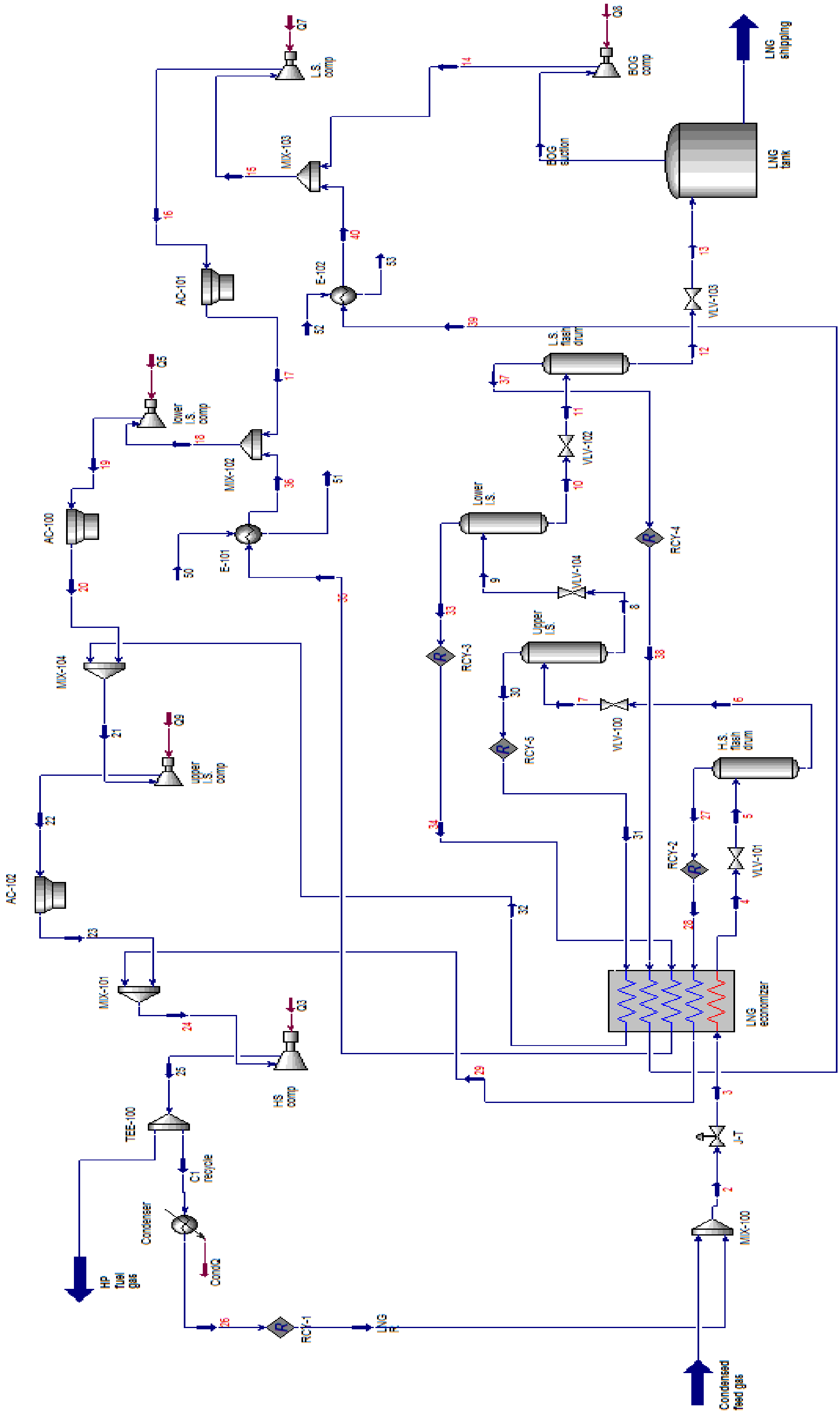


Figure 15: LNG back-end process optimized simulation CASE 2

Table 6: CASE 2 material balances extracted from HYSYS

Material Streams														
		Condensed feed gas	2	3	4	5	27	6	29	25	HP fuel gas	C1 recycle	7	33
Vapour Fraction		0.0000	0.0000	0.0556	0.0119	0.1385	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.1518	1.0000
Temperature	C	-90.00	-98.99	-102.9	-103.5	-112.0	-112.0	-112.0	-112.0	118.6	118.6	118.6	-125.8	-138.4
Pressure	kPa	4500	4500	2700	2700	1900	1900	1900	1900	4550	4550	4550	1000	500.0
Molar Flow	kgmole/h	1617	2868	2868	2868	2868	397.1	2471	397.1	1322	66.10	1256	2471	234.9
Mass Flow	kg/h	2.835e+004	5.000e+004	5.000e+004	5.000e+004	5.000e+004	7084	4.292e+004	7084	2.287e+004	1144	2.173e+004	4.292e+004	4015
Liquid Volume Flow	m3/h	90.34	155.0	155.0	155.0	155.0	20.19	134.8	20.19	68.27	3.414	64.86	134.8	12.20
Heat Flow	kJ/h	-1.402e+008	-2.377e+008	-2.377e+008	-2.384e+008	-2.384e+008	-2.754e+007	-2.108e+008	-2.753e+007	-8.472e+007	-4.236e+006	-8.048e+007	-2.108e+008	-1.741e+007
		10	35	19	24	20	26	11	37	12	39	16	17	18
Vapour Fraction		0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.1059	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000
Temperature	C	-138.4	-112.0	119.0	30.00	125.2	-110.9	-152.4	-152.4	-152.4	-112.0	77.44	28.00	28.00
Pressure	kPa	500.0	400.0	1000	1900	1000	4500	200.0	200.0	200.0	100.0	400.0	400.0	400.0
Molar Flow	kgmole/h	1861	234.9	548.1	1322	548.1	1256	1861	197.0	1664	197.0	313.2	313.2	548.1
Mass Flow	kg/h	3.233e+004	4015	9193	2.287e+004	9193	2.173e+004	3.233e+004	3278	2.906e+004	3278	5178	5178	9193
Liquid Volume Flow	m3/h	103.3	12.20	28.75	68.27	28.75	64.86	103.3	10.37	92.94	10.37	16.55	16.55	28.75
Heat Flow	kJ/h	-1.666e+008	-1.719e+007	-3.670e+007	-8.889e+007	-3.656e+007	-9.788e+007	-1.666e+008	-1.522e+007	-1.514e+008	-1.494e+007	-2.192e+007	-2.250e+007	-3.855e+007
		BOG suction	LNG shipping	13	15	14	LNG R	28	34	38	50	51	36	52
Vapour Fraction		1.0000	0.0000	0.0699	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.9901	1.0000	1.0000
Temperature	C	-162.4	-162.4	-162.4	-41.61	-158.5	-110.9	-112.0	-138.4	-152.4	100.0	99.96	28.00	100.0
Pressure	kPa	90.00	90.00	90.00	100.0	100.0	4500	1900	500.0	200.0	101.3	101.3	400.0	101.3
Molar Flow	kgmole/h	116.3	1548	1664	313.2	116.3	1251	397.1	234.9	197.0	2775	2775	234.9	2775
Mass Flow	kg/h	1900	2.716e+004	2.906e+004	5178	1900	2.165e+004	7084	4015	3278	5.000e+004	5.000e+004	4015	5.000e+004
Liquid Volume Flow	m3/h	6.175	86.76	92.94	16.55	6.175	64.62	20.19	12.20	10.37	50.10	50.10	12.20	50.10
Heat Flow	kJ/h	-9.232e+006	-1.422e+008	-1.514e+008	-2.324e+007	-9.217e+006	-9.751e+007	-2.754e+007	-1.741e+007	-1.522e+007	-6.643e+008	-6.654e+008	-1.605e+007	-6.643e+008
		53	40	30	8	9	32	21	22	23	31			
Vapour Fraction		0.9920	1.0000	1.0000	0.0000	0.1121	1.0000	1.0000	1.0000	1.0000	1.0000			
Temperature	C	99.96	25.00	-125.8	-125.8	-138.4	-114.4	32.00	95.87	90.63	-125.8			
Pressure	kPa	101.3	100.0	1000	1000	500.0	1000	1000	1900	1900	1000			
Molar Flow	kgmole/h	2775	197.0	375.2	2096	2096	376.7	924.8	924.8	924.8	376.7			
Mass Flow	kg/h	5.000e+004	3278	6567	3.635e+004	3.635e+004	6594	1.579e+004	1.579e+004	1.579e+004	6594			
Liquid Volume Flow	m3/h	50.10	10.37	19.26	115.5	115.5	19.33	48.08	48.08	48.08	19.33			
Heat Flow	kJ/h	-6.652e+008	-1.402e+007	-2.678e+007	-1.841e+008	-1.841e+008	-2.672e+007	-6.329e+007	-6.116e+007	-6.135e+007	-2.689e+007			

Optimization simulation case 2 fig. 15 flowsheet starts with Joule Thompson effect valve as previous cases by breaking down the front pressure from 45 barg in stream 2 to 27 barg in stream 3. The condensed feed is about 9 degrees Celsius colder than the base case process' feed. The process is pretty much similar up to before the condensed feed undergoes further cooling in the LNG economizer. The first big modification is a larger LNG economizer since it handles the single hot stream and the four cold vapor streams from the four respective sub-cooling stages: lower stage, lower-intermediate stage, upper-intermediate stage and high stage. What characterizes optimization simulation proposal 2 is the addition of one extra subcooling stage to the process, becoming an Optimized cascade LNG back end process with four sub-cooling stages. Adding an extra subcooling stage results in more equipment yet smaller in size. Smaller equipment such as compressors requires lesser power for the process. However the subcooling chills the process resulting in a colder feed stream of -98.99°C.

The optimized simulation proposal 2 produces 27160 kg/hr of LNG at -162.4°C. The storage conditions are pretty much the same. Optimization proposal 2 is more complex with extra flash drum, extra compressor thus overall a whole extra sub-cooling stage. However the performance of this optimized simulation case 2 is much more efficient with about only half of power required to produced 1kg of LNG compared to the base case. A comparison table among the three cases is presented in the next section.

4.3 COMPARISON OF BASE CASE VERSUS OPTIMIZED SIMULATION CASES 1 and 2.

Table 7 illustrates a brief comparison of the 3 simulation results obtained during this study. *Start-up feed*, a term used in the comparison table 7, refers to the feed required to start-up the unit. With the unit started the vapor recycled, condensed and joining the feed stream this computes the feed resulting in reduction of main feed. The now adjusted feed is referred in the comparison table as *after recycle feed*.

The main objective of this study has been to explore the opportunity to optimize the performance of the base case simulation by reducing fuel gas production and increasing LNG production leading to increment of LNG sales benefits.

H.-M. Chang et al. (2012) also addressed the overall performance of a liquefaction system stating that the thermodynamic performance of a liquefaction system is evaluated in terms of the work required per unit mass of liquefied gas. That performance is address in the next lines and in table 7 as specific power of production.

Optimized simulation case 2, also called Optimized proposal 2 throughout this report fig. 15 appears to be most efficient process with about half reduction of power consumption to produce 1kg of LNG. Optimized simulation case 2 consumes 349 kJ to produce 1kg of LNG compare to 903KJ/Kg base case and 973KJ/Kg optimized simulation case 1.

Optimized simulation case 2 fuel gas production flowrate is 1144Kg/hr which is a reduction of 37% compared to the base case. What is more, case 2 LNG production is 27160kg/hr which yields a production of 13660kg/hr more compared to the base case.

Nitrogen content of LNG produced in case 2 is reduced from 0.0050 mole% (condensed feed) to 0.0008 mole%.

With the key features such as specific power of production reduction to half yet doubling the production of LNG, optimized simulation case 2, fig. 15 is the proposed method to achieve the objective of this project.

Table 7: Comparison of base case versus optimized cases 1 & 2

LNG back-end process	**Base case**	**Optimized case 1**	**Optimized case 2**
Start-up feed, kg/hr	50000	50000	50000
After recycle feed, kg/hr	15440	24580	28350
Produced LNG, kg/hr	13500	23180	27160
Production efficiency (%)	87.44	94.30	95.80
Feed nitrogen content, mole%	0.0050	0.0050	0.0050
LNG nitrogen content, mole%	0.0003	0.0009	0.0008
LNG LHV, KJ/kg	49490	49480	49490
Fuel gas produced, kg/hr	1825	1338	1144
Fuel gas ratio (%)	11.82	5.44	4.04
Fuel gas LHV, kJ/kg	46930	45550	41590
Compressor fuel gas, MW	3.4	6.3	2.6
Additional LNG production, kg/hr	-	9680	13660
Specific power of production, KJ/kg	903	973	349

For the calculation formulas used in the above table refer to result calculations sample in appendix 1.

Figure 16: LNG economizer Temperature (C) vs. Heat flow (KJ/hr) BASE CASE

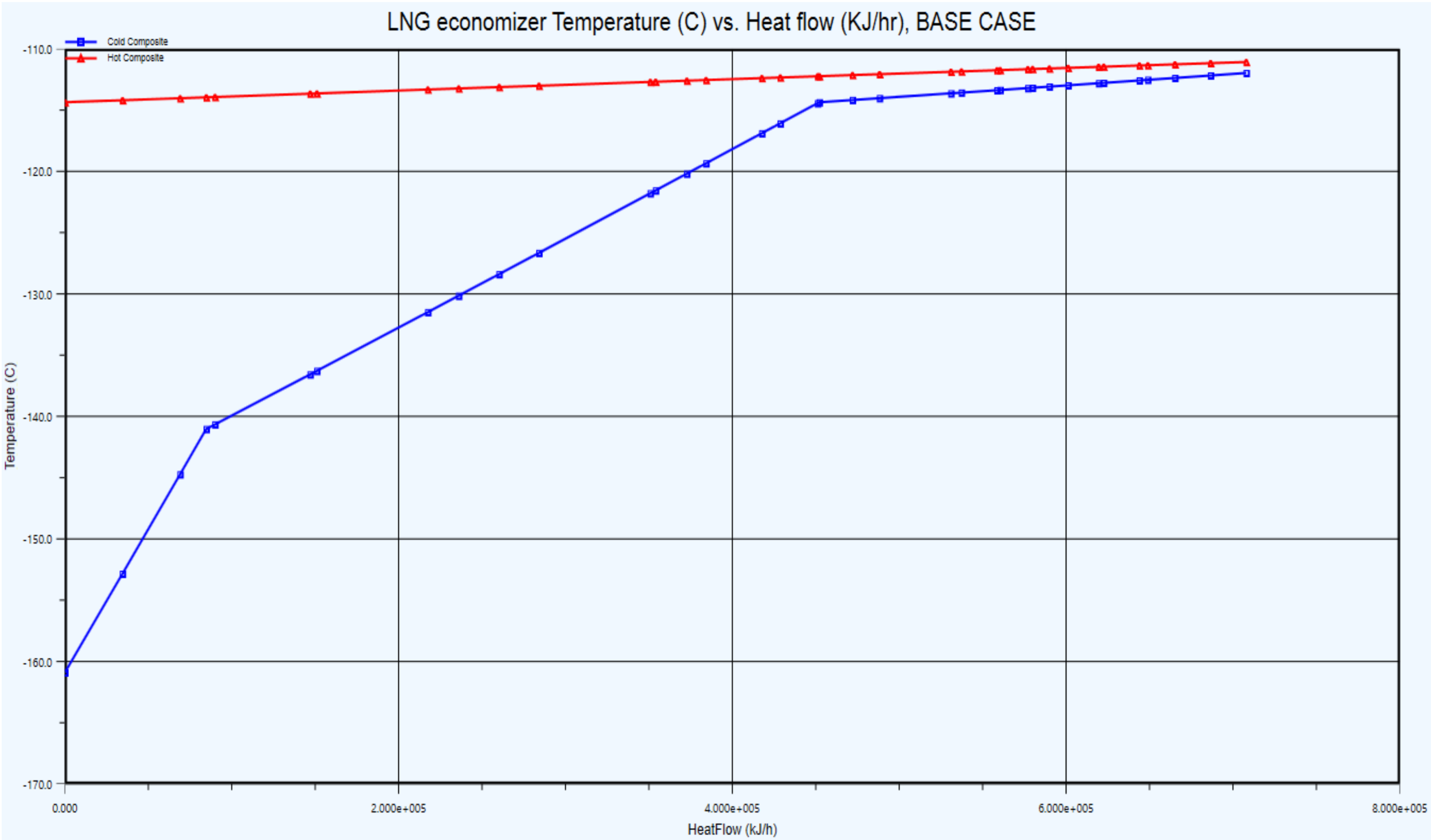


Figure 17: LNG economizer 1 Temperature (C) vs. Heat flow (KJ/hr) CASE 1

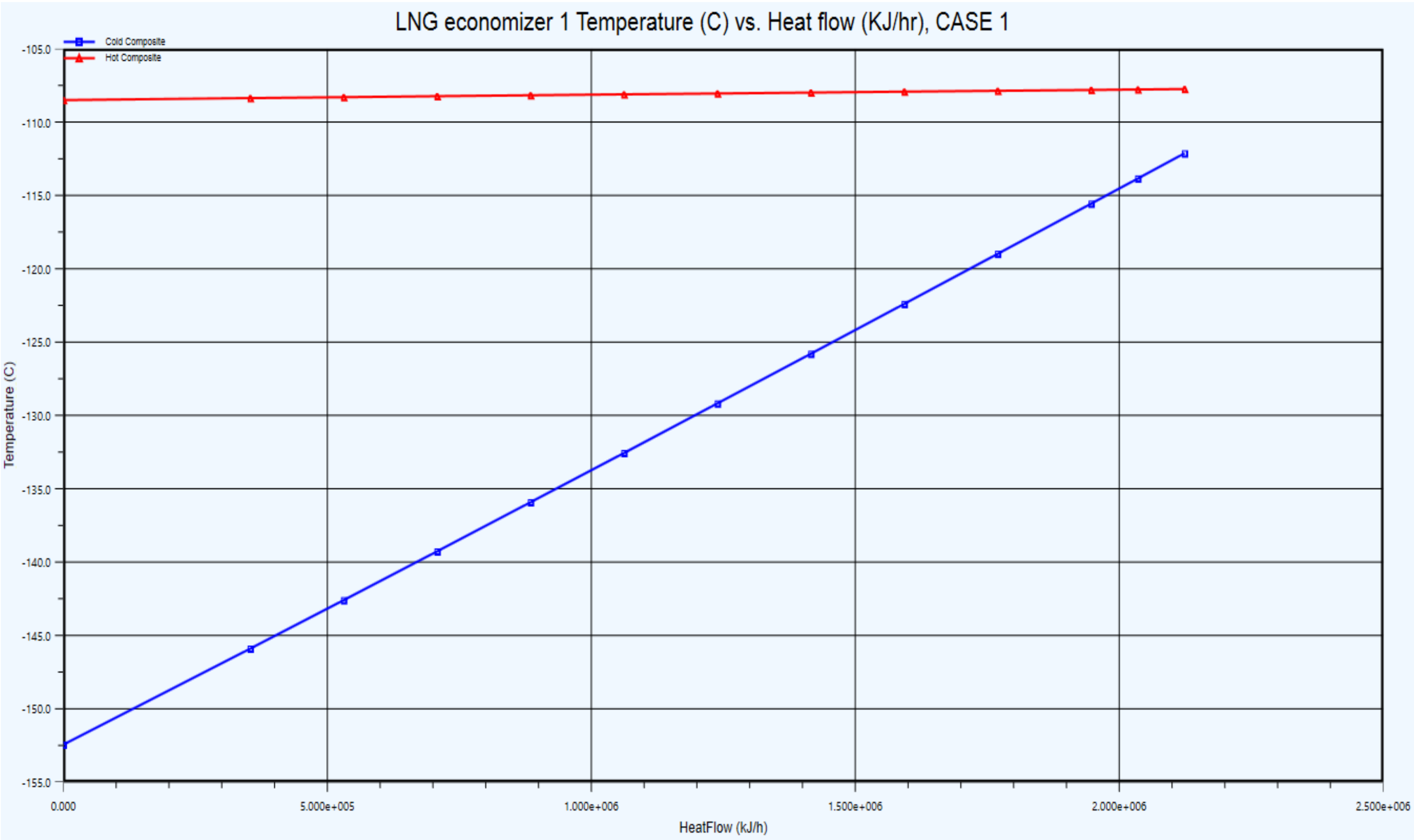


Figure 18: LNG economizer 2 Temperature (C) vs. Heat flow (KJ/hr) CASE 1

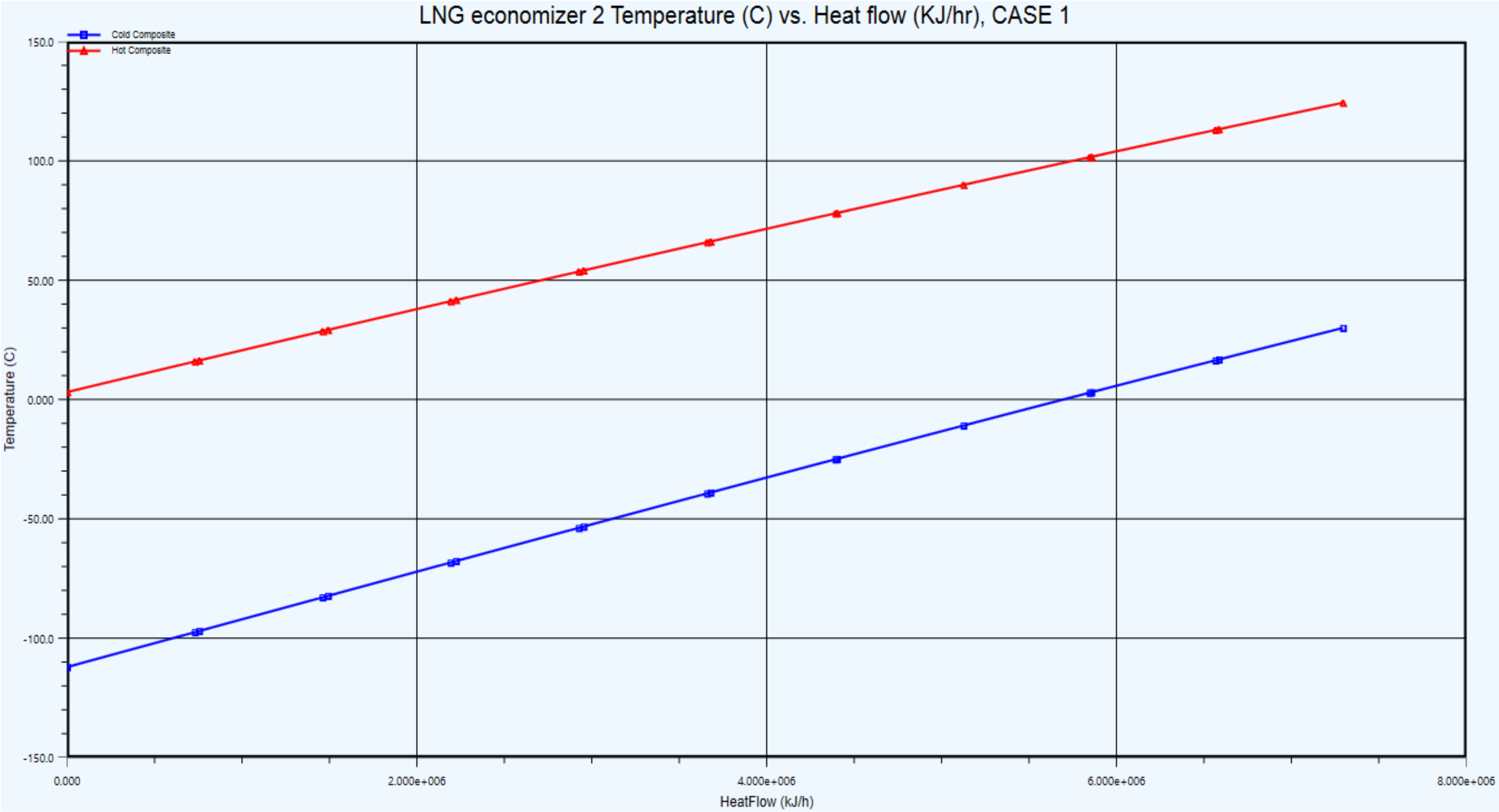
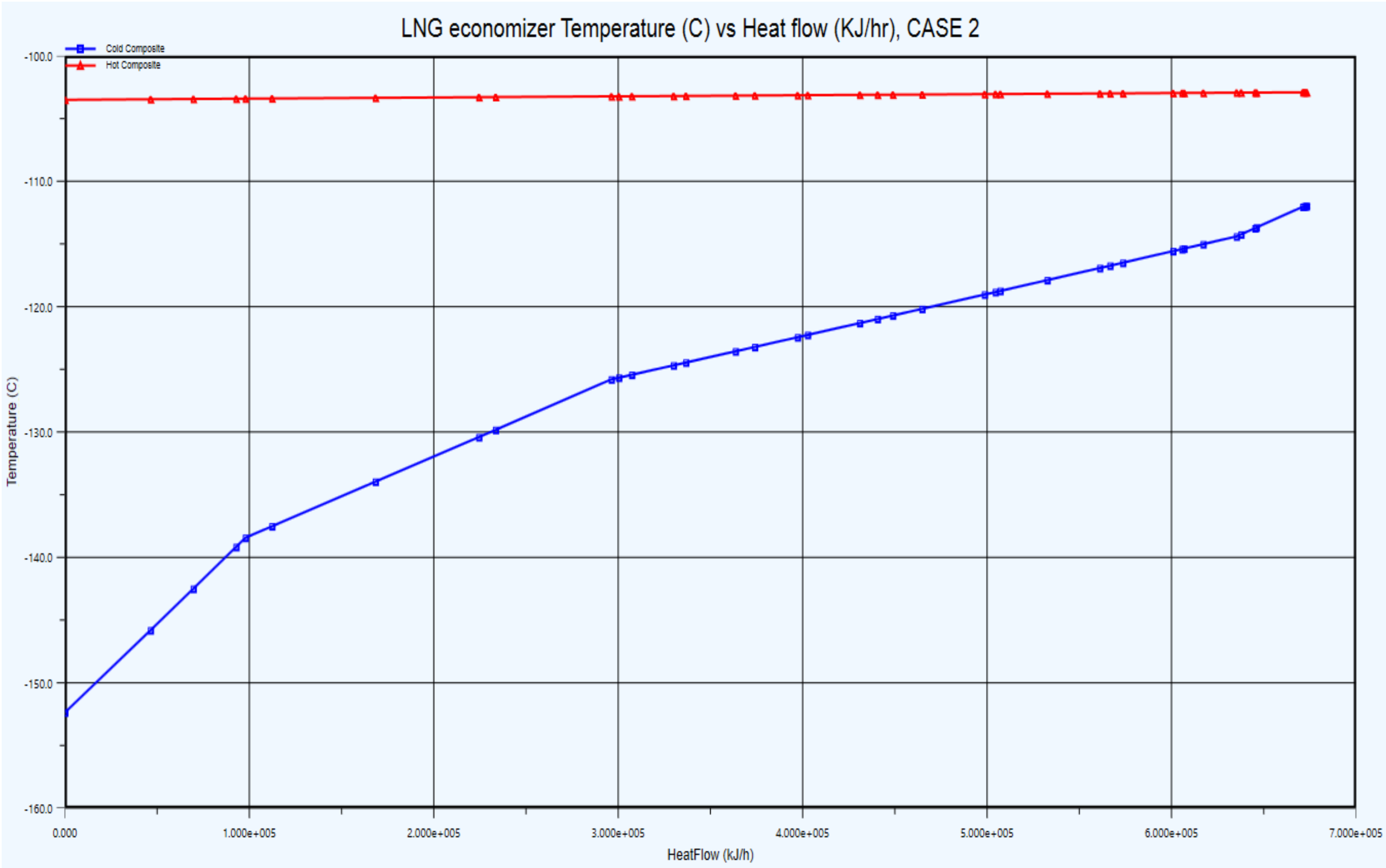


Figure 19: FIGURE 19: LNG economizer Temperature (C) vs. Heat flow (KJ/hr) CASE 2



CHAPTER 5: CONCLUSION AND RECOMENDATION

5.1 CONCLUSION

Reporting to the main objective of this project which has been to explore opportunities of process optimization by improving process efficiency has been achieved by the results of this project. Process efficiency here is defined as the increase of LNG sales production by reducing fuel gas production all with the challenge to maintain or reduce the specific power of LNG production.

The results displayed in table 7 confirmed the results solution of the problem statement of this project and Optimized case 2 simulation is indeed the proposed solution with only 349 KJ of power required to produce 1 kg of LNG which is about 39% of the power requirement in the base case simulation of this study. Table 7 also shows and increase of 13660kg/hr of more LNG produced with the fuel gas production cut down from 1825kg/hr to 1144kg/hr. Cases 1 and 2 are improved process efficiencies of the base case however, Optimized Case 2 in fig. 15 is the solution proposal for this study.

5.2 RECOMMENDATIONS.

We think we have created a solid base in this project for further studies. The simulations were carried out using Aspen HYSYS steady state module which neglects or assume and in fact it calculates for instance the sizing of equipment and without user specifications entry. With dynamic simulations actual sizing takes place and the results are more accurate and closer to reality. Results of this project are a good starting point to transfer the steady state simulation into dynamic simulations and obtain more accurate and complete results.

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APPENDICES

Appendix 1: Result calculations sample

Production efficiency calculation of the base case simulation:

$$Production\ efficiency\ (\%) = \frac{Produced\ LNG}{After\ recycle\ feed} \times 100$$

$$= \frac{13500\ kg/hr}{15440\ kg/hr} \times 100 = \mathbf{87.44}$$

Fuel gas ratio calculation of the Optimized case 1 simulation:

$$Fuel\ gas\ ratio\ (\%) = \frac{Fuel\ gas\ produced}{After\ recycle\ feed} \times 100 = \frac{1338\frac{kg}{hr}}{24580\frac{kg}{hr}} \times 100 = \mathbf{5.44}$$

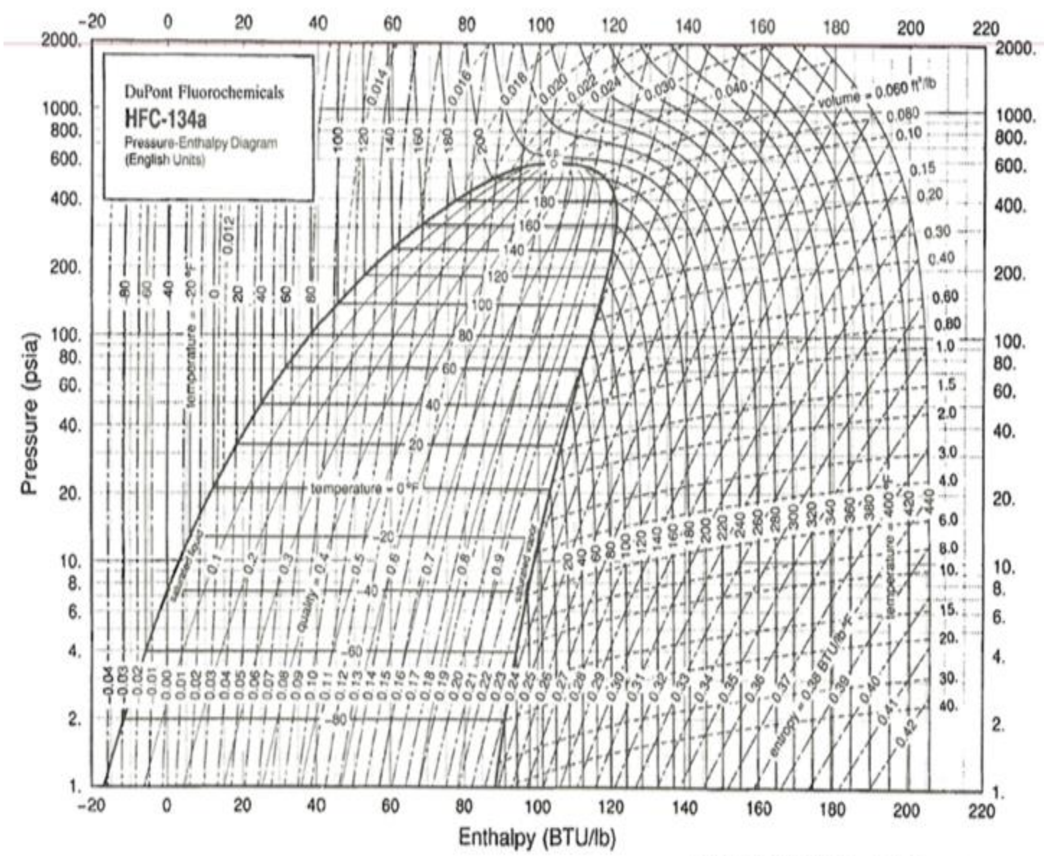
Specific power of LNG production (efficiency) of the Optimized case 2 simulation:

$$specific\ power\ of\ LNG\ production = \frac{Total\ compressors\ power\ consumed}{Produced\ LNG}$$

$$= \frac{2627.938\ Kw}{27160\ kg/hr} = \frac{2627.938\ KJ/s}{7.544\ kg/s} = \mathbf{349\ KJ/kg}$$

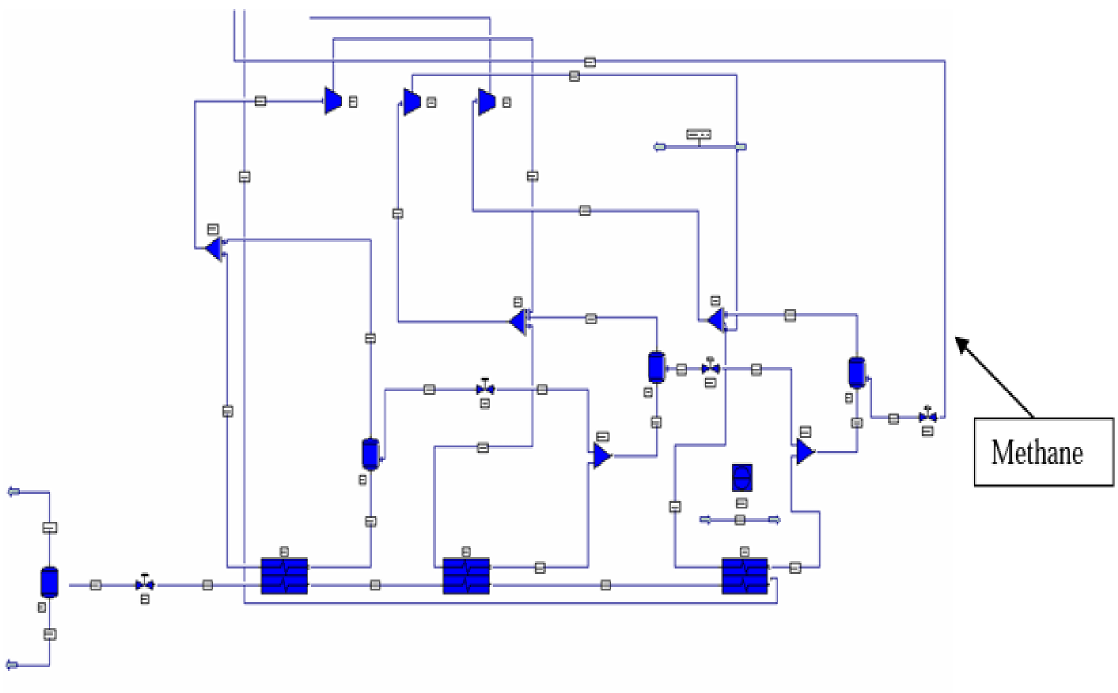
Appendix 2: Pressure-Enthalpy refrigerant loops

Pressure-Enthalpy Diagram for HFC-134a[12]



Courtesy of DuPont HFC-134a Refrigerants

Appendix 3: Example of a methane loop.



Appendix 4: Three-stage propane refrigeration system (courtesy of GPSA)