Sensitivity Analysis of Process Parameters on Upstream Equipment and Downstream Products in a Single Cascade Refrigeration Cycle

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Abstract: **This work was designed to showcase the effect of upstream parameters on downstream products in a simple cascade refrigeration cycle in Niger Delta oil and gas production field. The generalized algorithm was developed and modeled on ASPEN HYSYS version 8.6 using the Peng-Robinson equation of state. The model consists of two segments- liquefaction and the fractionation. The liquefaction segment is a cascade refrigerant cycle which employs the use of an LNG heat exchanger, compressor, throttle valve and condenser process units while the fractionation unit employs the use of fractionating columns. This study models the relationship of the molar flow of propane refrigerant and the power consumed by the compressor, heat flow across the compressor and the condenser heat duty. It also showcases the effect of natural gas feed temperature on the inlet temperature of the fractionators as it relates to the temperature of top and bottom products as a means of providing solutions to specific and specialized natural gas processing; supplies data and process information for both technical and investment decision on the hydrocarbon utilization, optimization and application. The study reveals that a linear relationship exists between the molar flow of propane refrigerant and the power consumed by the compressor, heat flow across the compressor and the condenser heat duty. Also, a linear relationship also exists between the gas feed temperature and the de-methanizer and de-ethanizer fractionators but beyond the de-ethanizer there is no significant effect of the gas feed temperature on fractionators products.**

Key Words: **Natural gas, Parameters, Simulation, Process, Fractionators, Mathematical model**

I. INTRODUCTION

As at 2016, Nigeria's proved crude oil and natural gas reserve is estimated at 37.1 billion barrels and 180.5 trillion cubic feet respectively making her the nation with the 9th largest gas reserve in the world behind Iran, Russia, Qatar, Turkmenistan, United States, Saudi Arabia, United Arab Emirates and Venezuela [(3)] therefore one can conclude that Nigeria is a gas province with pockets of oil. Natural gas as a non renewable resource must be optimally utilized as an act of sustainable development particularly because it is the cleanest source of energy compared to other commercial fossil fuels and has its use ranging from electricity generation, transportation fuel to use as feed stock for petrochemicals industry [(9)]

As part of Nigeria's resolve to become a major international player in the international gas market as well as to lay a solid framework for gas infrastructure expansion within the domestic market, the Nigerian Gas Master Plan was approved on February 13 2008 which is a guide for the commercial exploitation and management of the gas sector with the aim at growing the Nigerian economy with natural gas [(6)].

The main objective of this study is to develop a reliable model for universal evaluation of the natural gas reserve, determine processing requirements and prepare for investments. Data on Nigerian oil and gas is available and can be obtained from the Nigerian national Petroleum Corporation (NNPC) and the Department of Petroleum Resources (DPR) where they are domiciled. This study has therefore taken advantage of these to do the design and simulation of natural gas as they undergo liquefaction and fractionation.

II. LITERATURE REVIEW

Due to the low liquefaction temperature of LNG (-160^{o}C), LNG production has high energy requirement thus high amount of used up for refrigeration which implies high cost of production [(12)]. For this reason, researchers' attention has been drawn to this in the last decade.

2.1 LNG Production Optimization: From history, a lot of designs have been recommended for LNG optimization, they have varied from equipment design to refrigerants to be used. Some of such include turbo expander units, cascade process, dual/single mixed refrigerants (SMR/DMR) process [(11)]

In cascade circle sequential stages are applied with each stage having progressively colder refrigerants, having their own compression system. The refrigerants include C_1-C_3. Using sequence of pressure drops to lower temperature an advanced approach between the cold and hot composite curves is achieved thereby enhancing efficiency and provides better process control since there are few variables to deal with. However, all these leads to high operating costs therefore it is only economical when applied to a large scale process [(12).]

In very recent time, the mixed refrigerant process has been used for over seventy-five percent of new LNG plants because of its low heat exchanger and compressor requirement thus less capital intensive. However, the application of MR process increases process operation and design complexity due to increased thermodynamic which makes the control system very challenging [(2)].

2.2 Investigations of the Cascade Model

Paradowski et al. [(7)] carried out a parametric research on pre-cooled propane mixed refrigerant cycle. In this study, mixed refrigerant composition, propane cycle pressures, pre-cooling temperatures and propane cycle compressor speed were investigated.

Wang et al [(5)] minimized energy consumption by using a synthesis approach in accordance with mixed integer nonlinear programming formulation. The team conducted preliminary thermodynamic analysis, simulation and optimization.

Hamidreza and Saffari [(4)] did an energy optimization of an industrial propane pre-cooled mixed refrigerant LNG base load plant by varying the components and the mole fractions in liquefaction and sub-cooling cycles. This process was simulated using HYSYS software, the Peng Robinson equation of state was used as the fluid package. Two approaches for modelling and optimization were studied and a number of parameters were evaluated.

In 2015, there was an investigation of three selections of variables. In the three sets, the refrigerant compressor suction and compressor discharge pressure were used as variables. The additional variable was the refrigerant flow. It was discovered that using the component molar flow performed slightly better than using the molar fractions and total molar flow and that the use of heat flow provided less success [(8)].

2.3 Literature Gap

Previously published research has provided a strong foundation for further studies. In addition to various parametric studies carried out on the single loop cascade liquefaction system, this research showcases the effect of molar flow rate of a propane refrigerant on the compressor and condenser load. It goes further to reveal the relationship between the natural gas feed temperature into the LNG heat exchanger and the inlet temperature of the fractionators as well as its effect on the temperature of top and bottom products of the C1 C2 and C3 fractionators.

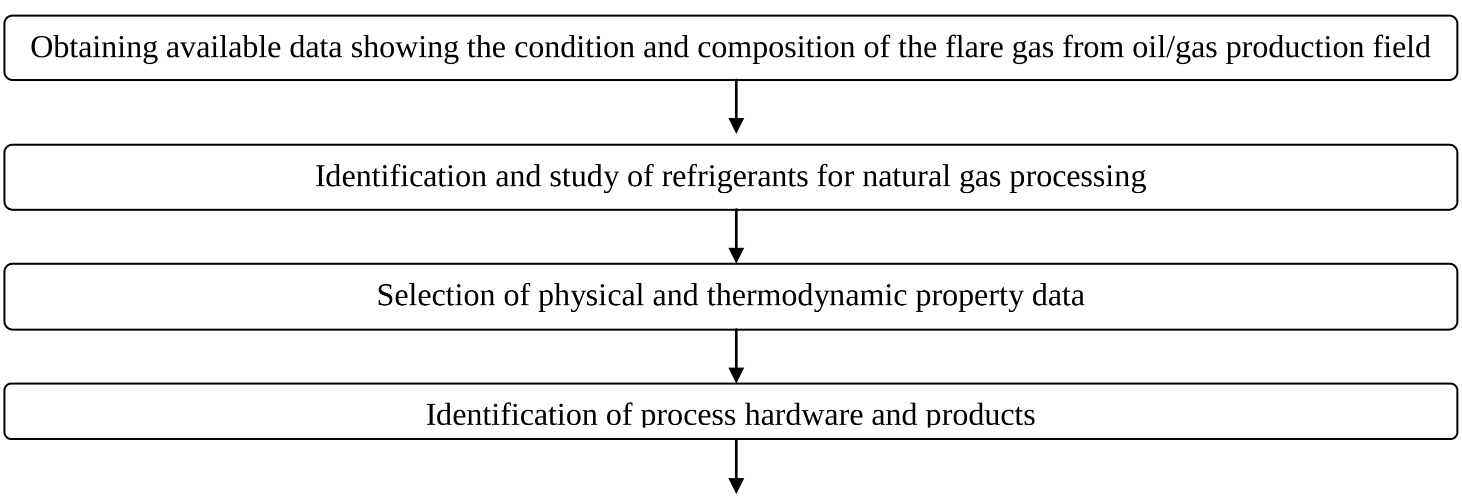
III. METHODOLOGY

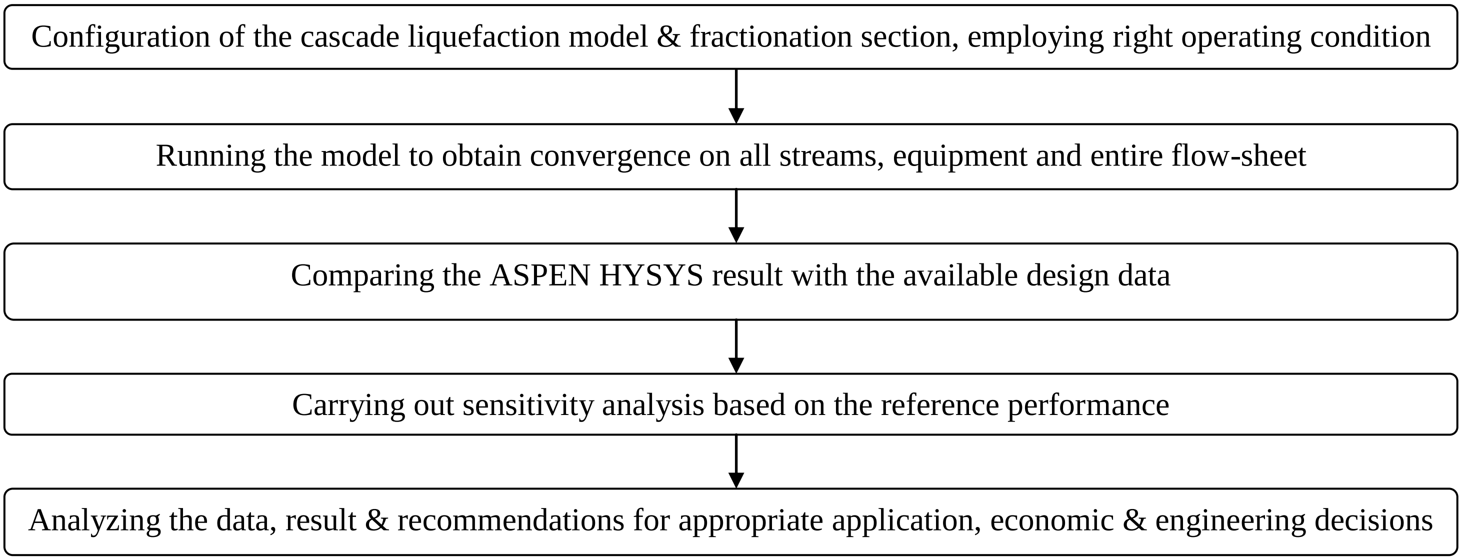
3.1 Overview

This is a stride towards developing a robust model for the technical and commercial evaluation of natural gas. Liquefaction/refrigeration process is a process used to separate natural gas (wet or dry) into its main components. [(1)]The cascade refrigeration system is a system which employs the use of refrigerants (pure or a mixture of compounds) to achieve cooling/freezing through a heat exchanger. The refrigeration employs pressure and temperature changes on a refrigerant to achieve heat transfer. The fractionation segment on the other hand employs the use of fractionating vessels, taking advantage of the difference in boiling points of the gas constituent particularly methane, ethane and propane.

3.2 Procedural Algorithm

A stepwise approach was adopted in this study and is shown below:





3.3 Simulation Tool

There are several simulation packages that are available, but ASPEN HYSYS provides one of the best process modeling environments for conceptual design and operations improvement of petroleum and oil and gas process. This modeling tool has been used over the years by researchers and engineers to achieve improved engineering design and operations and energy efficiency as well as reduce capital cost thus the choice of ASPEN HYSYS version 8.6. Peng-Robinson thermodynamic model is the chosen fluid property package.

3.4 Equipment

The model is a cascade liquefaction/refrigeration approach which employs a compressor, condenser, valve and an LNG unit operation. Downstream equipment includes the fractionating columns, reboilers, pumps and valves.

3.5 Feed Stream Parameters

The plant takes in a single feed stream of pre-treated natural gas from the gas recovery section of a gas flare system. The flare condition and composition (feed) is obtained based on available laboratory data.

Table 2.1: Feed Gas Condition and Composition from a Niger Delta Field

Property	**Value**
Temperature (^{o}C)	30
Pressure (kPa)	2275
Flowrate (SCFD)	32460000

Composition	**Mole fraction**
Nitrogen	0.0025
Carbon dioxide	0.0048
Methane	0.7041
Ethane	0.1921
Propane	0.0706
Iso Butane	0.0112
Normal Butane	0.0085
Iso Pentane	0.0036
Normal Pentane	0.0020
Hexane	0.0003
Heptane	0.0002
Octane	0.0001

3.6 Product Specifications

The main interest is for the modeled refrigeration/liquefaction cycle to be able to bring the feed to the operating conditions of the downstream process facilities. Therefore the feed (natural gas) temperature (about 30^{o}C) at a given pressure on exiting from the LNG heat exchanger unit should meet the temperature and pressure requirement of the fractionators. Also, the final downstream product should meet the specification for end use.

IV. DESCRIPTION OF SIMULATION ENVIRONMENT

4.1 Parameters

Modeling and simulating the liquefaction and fractionation segments of the natural gas process is very technical, especially to converge the distillation columns. Another test of accuracy is that the resulting values of the unspecified conditions and parameters of process units is an accurate representation of reality.

The parameters below are considered in order to achieve a good process simulation results with minimal error:

- Type and molar composition of refrigerant
- Compressor adiabatic efficiency
- Condenser pressure drop
- Pressure drop at LNG heat exchanger
- Condenser outlet temperature
- Natural gas outlet temperature from the LNG heat exchanger
- Column specifications
- Product specifications

4.2 The Single Cascade Model

Here the refrigerant is pure propane [13]. The expansion of the throttle valve greatly reduces the temperature of the pure propane, thereafter it goes into the LNG heat exchanger where it exchanges heat with the hot natural gas stream making the natural gas leave the LNG heat exchanger at a much lower temperature ready to be fed into the fractionators.

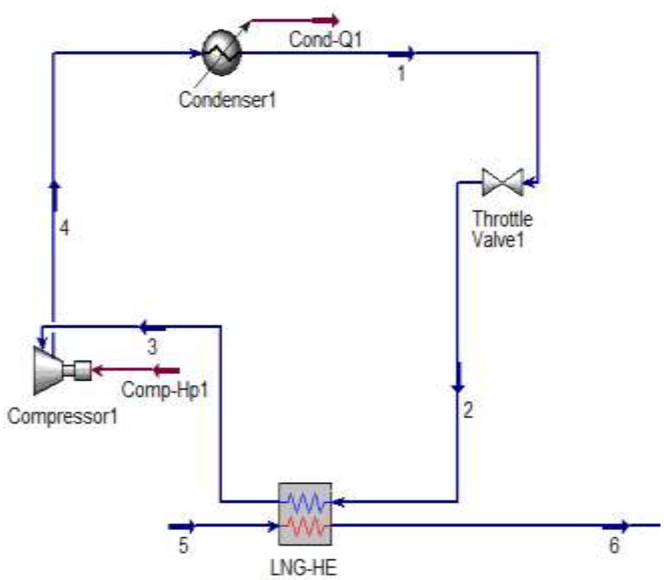


Figure 4.1: Single Cascade Refrigeration Configuration on ASPEN HYSYS

For the above cascade model to converge, care must be taken to ensure that the operating conditions of all unit operations. Typically, the phase of the refrigerant as it passes through the compressor is set as 100% vapour. As a result of the temperature rise that often occurs after compression and the work the condenser will do on the refrigerant, the exit stream of the condenser is usually set at a lower temperature.

4.3 De-Methanizer

Modeling the de-methanizer is the most complex of the fractionation segment. Modeling the de-methanizer in ASPEN HYSYS requires the use of a reboiled absorber column as opposed to the distillation column used for other downstream fractionation process. A stream of the feed (pretreated natural gas from flare). The feed into the de-methanizer is at the operating temperature of the column usually less than -65°C but will differ depending on set parameters. In the de-methanizer, methane is recovered (as the major product with traces of other fractions) as the top product and NGL as the bottom product.

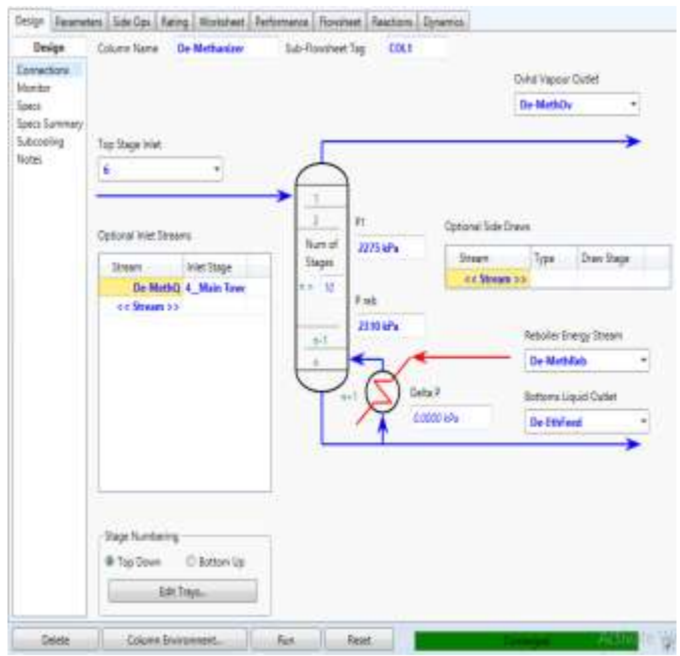


Figure 4.2: ASPEN HYSYS Simulation Showing the De-methanizer Column Specifications

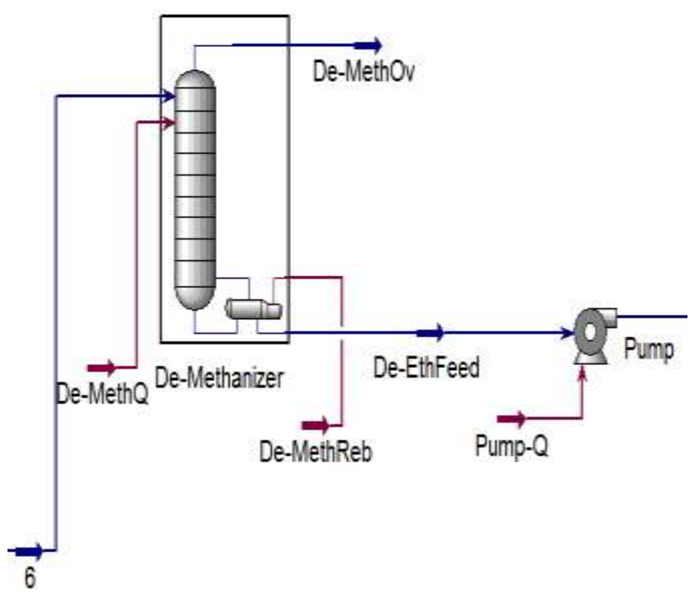


Figure 4.3: ASPEN HYSYS Simulation – Demethanizer

4.4 De-Ethanizer

The de-ethanizer feed conditions are set from text standard texts and research works [10] other front end parameters are gotten through trial and error method. Specifications such as component fraction of the top and bottom products are very important. The convergence of the column is dependent on many factors which must be carefully aligned to meet the columns convergence requirement.

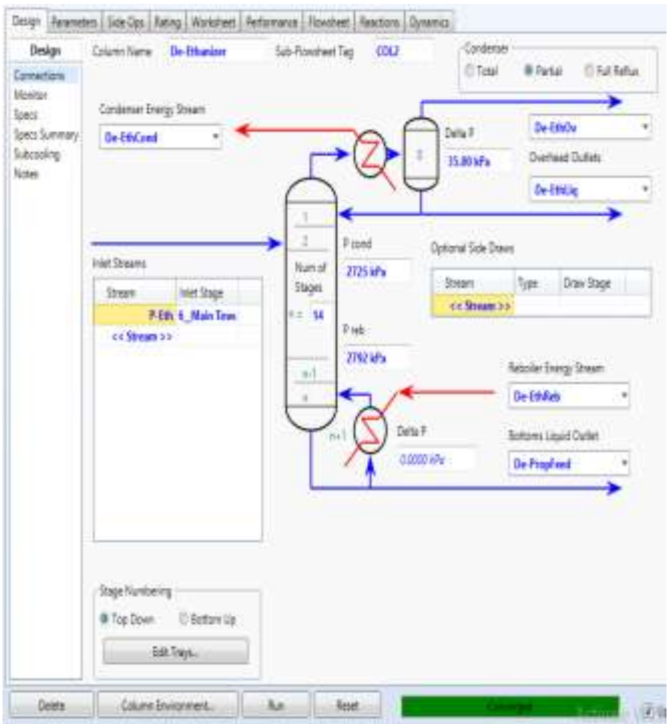


Figure 4.4: ASPEN HYSYS Simulation Showing the De-Ethanizer Column Specification

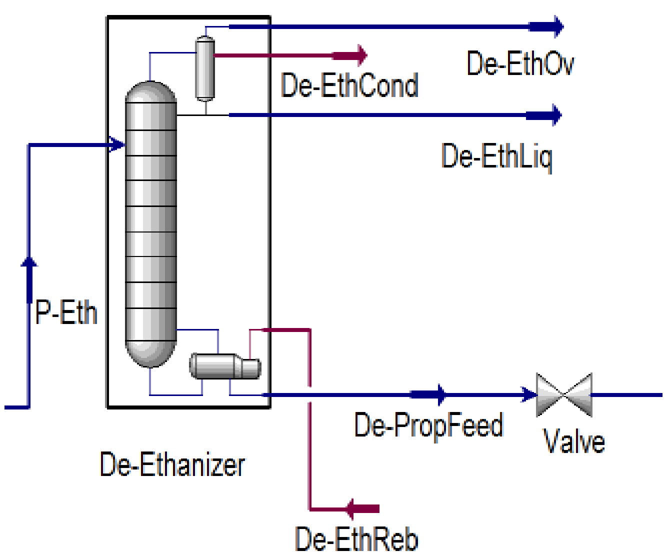


Figure 4.5: ASPEN HYSYS Simulation – De-ethanizer

From the de-ethanizer, ethane is recovered (as the major product with traces of other fractions) at the top product while the de-propanizer feed is the bottom product.

4.5 De-propanizer

This column is used next in order to separate propane from the heavier hydrocarbons fed into it. The overhead product is majorly propane while the other fractions are received as overhead products. The modeling steps are similar to those of the de-ethanizer. Similar to the de-ethanizers, condensers and reboilers are used.

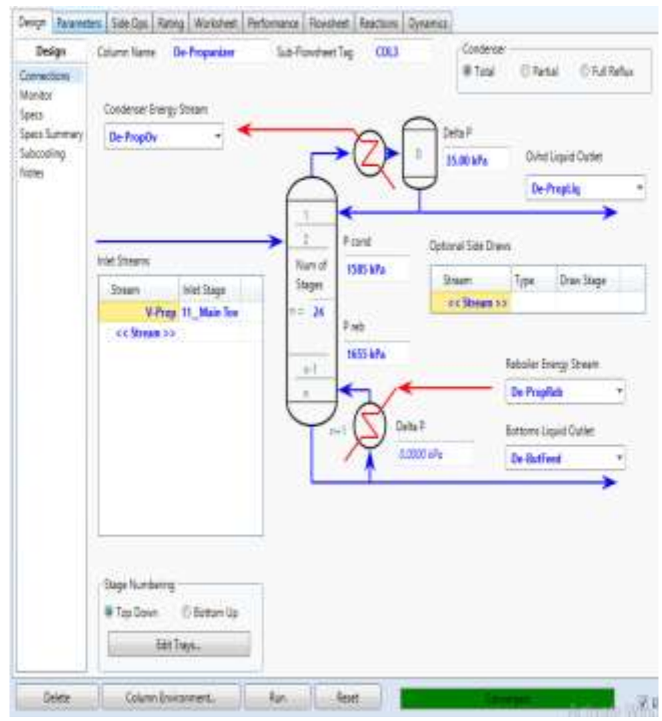


Figure 4.6: SPEN HYSYS Simulation Showing Some De-propanizer Column Specification

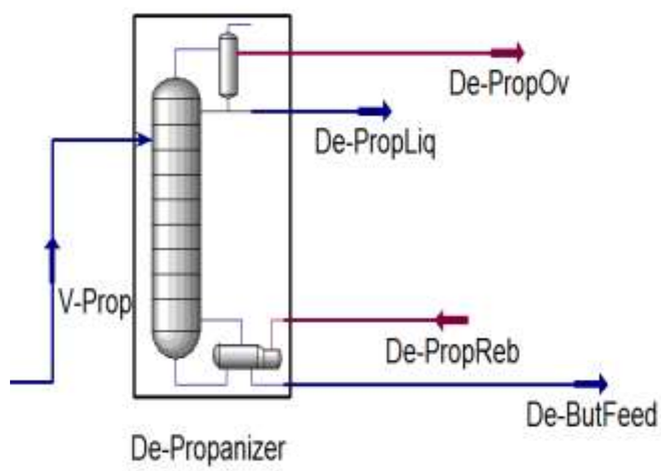


Figure 4.7: ASPEN HYSYS Simulation – De-propanizer

4.6 Reboiler and Condenser

Reboiler and condenser are part of the de-ethanizer and de-butanizer columns. The reboiler is used to maintain the temperature required for the fractionating column to operate. With the reboiler, lighter hydrocarbons will be vaporized unto the top stage. Steam is used as the heat source. The condenser is used to condense the overhead liquid which is entrained in the overhead vapour. Being part of the distillation column, the condenser and reboiler are defined based on the desired product specification.

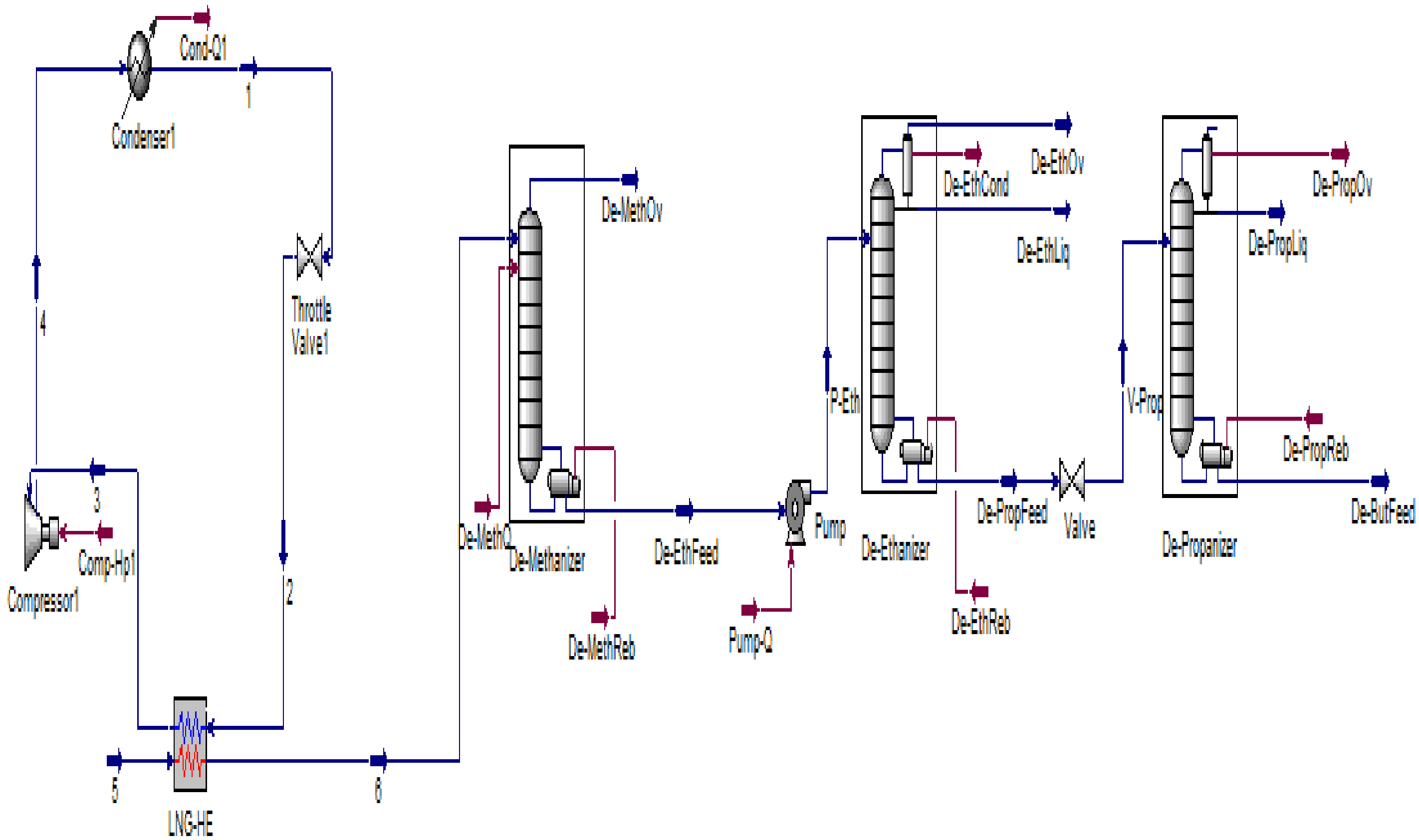


Figure 4.8 ASPEN HYSYS Simulation – The Complete Flowsheet for Single Cascade Model

Assumptions

The following assumptions were made:

a. The feed gas has been pretreated for the removal of contaminants such as acid gases, sulphur compounds, mercury and water.

b. There is no chemical reaction between the gas/refrigerant and the walls of the vessels

V. RESULTS AND DISCUSSION

After several simulation runs were performed for various cases and process conditions, the following results were obtained and discussed below:

5.1The Liquefaction Segment

The rating and performance of the equipment that make up the single and double cascade liquefaction configuration are dependent upon many factors which include refrigerant molar flow through the loops, temperature drop across the condensers, temperature drop across the LNG heat exchangers, the molar composition of the refrigerant (for mixed refrigerant) across the loop(s) and so on. Some of this above factors are addressed and demonstrated with some relationships developed.

Refrigerant Molar Flow Rate and Compressor Power Consumption

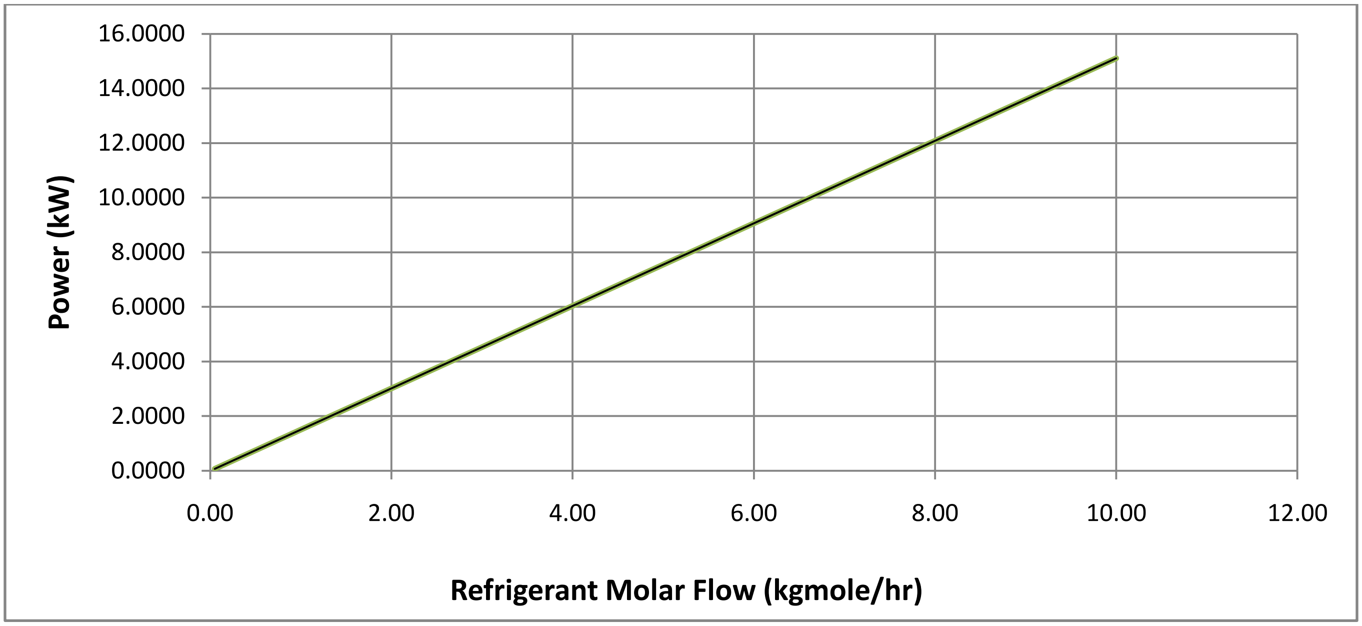


Figure 5.1: Power against Molar flow of Pure Propane Refrigerant

The single cycle cascade which uses pure propane refrigerant, the molar flow rate of the refrigerant across the loop is proportional to the power consumption and the heat flow rate across the compressor. The table showing this detail as well as other key parameters which values were chosen to achieve convergence is as illustrated in Appendix II while the plots are shown in Figure 4.1.

The plot above gives a linear model satisfied by the equation **Y = 1.51x + 5E-05** with a coefficient of determination R^2 =1. This is a pointer which indicates a direct proportionality of the power consumed in the compressor of a cascade process to the molar flow rate of propane refrigerant across that same loop.

Refrigerant Molar Flow Rate and Heat Flow across a Compressor

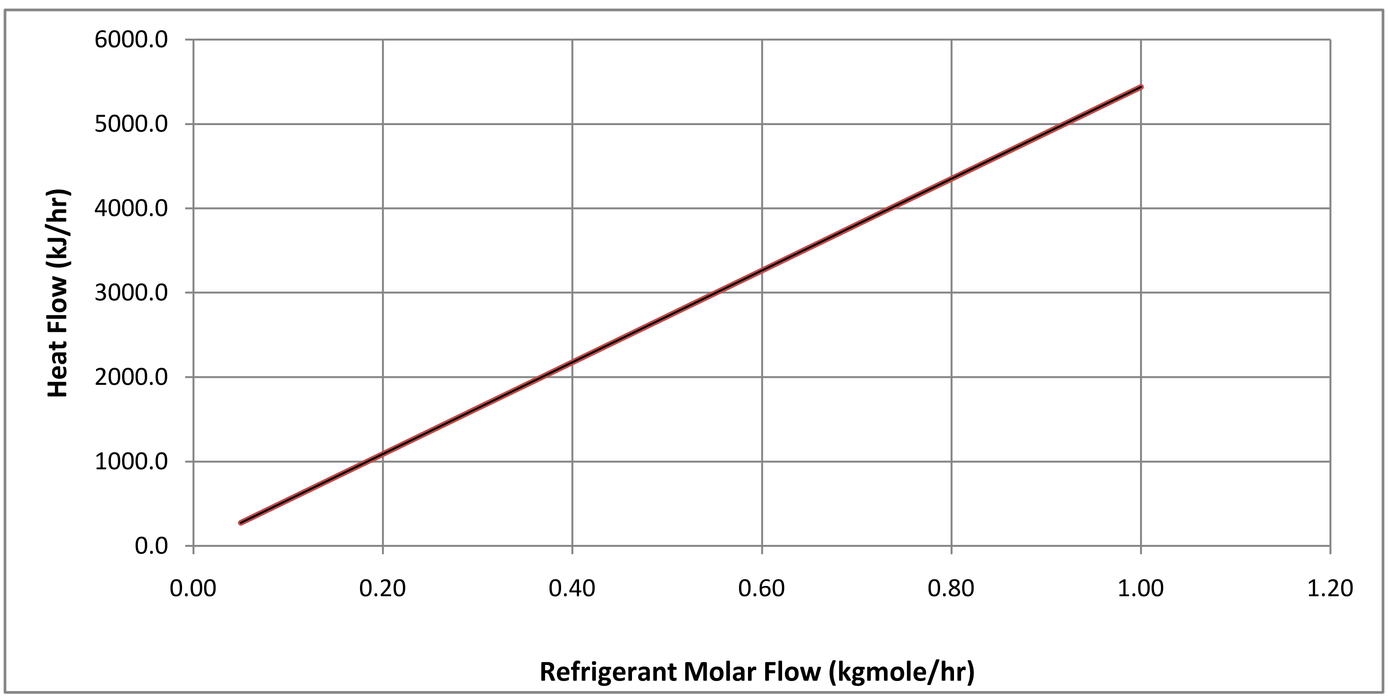


Figure 5.2: Heat Flow against Molar flow of Pure Propane Refrigerant

As expected from basic knowledge of thermodynamics and fluid mechanics, the molar flow rate of propane across the refrigeration loop will have a direct proportionality. The graph illustrates same. The plot above is a linear model satisfied by the equation **Y = 5435x - 0.042** with a coefficient of determination R^2=1. This is a pointer which indicates a direct proportionality of the heat flowing across the compressor of a cascade process to the molar flow rate of propane refrigerant across that same loop.

Refrigerant Molar Flow Rate and Condenser Duty

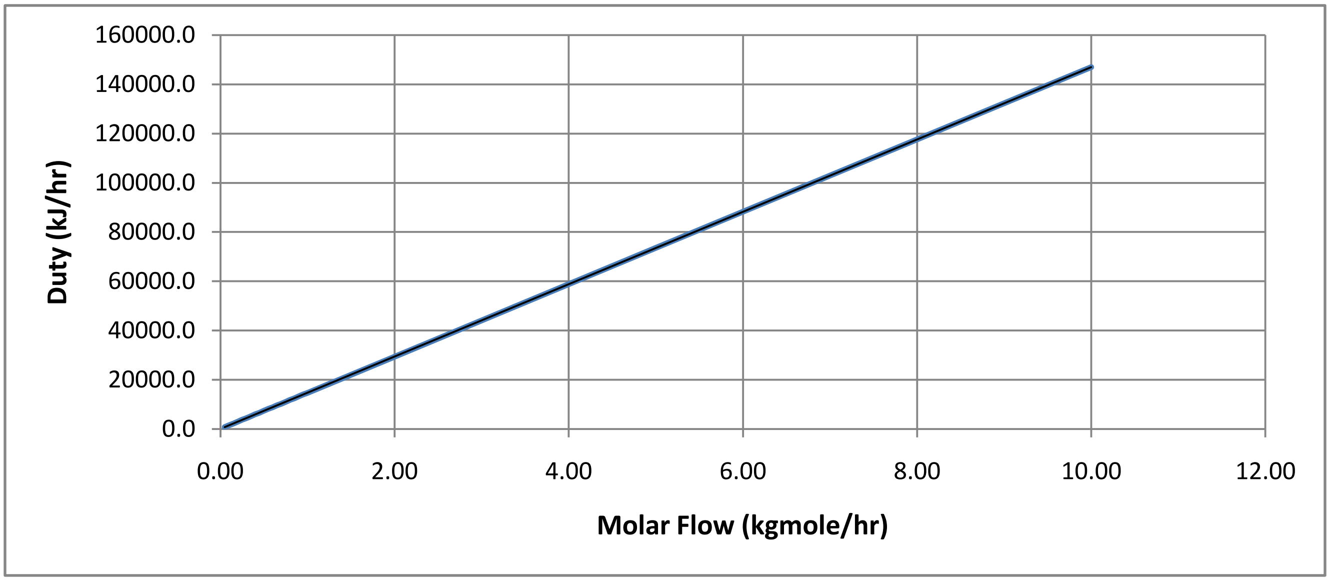


Figure 5.3: Plot of Condenser Duty against Molar Flow of Pure Propane Refrigerant

This is another linear relationship satisfied by the equation **Y = 14700x + 1.073** with a coefficient of regression, $\mathbf{R^2 = 1}$

Effect of the Natural Gas Feed Pressure into LNG Heat Exchanger on the Outlet (Fractionators Inlet) Temperature

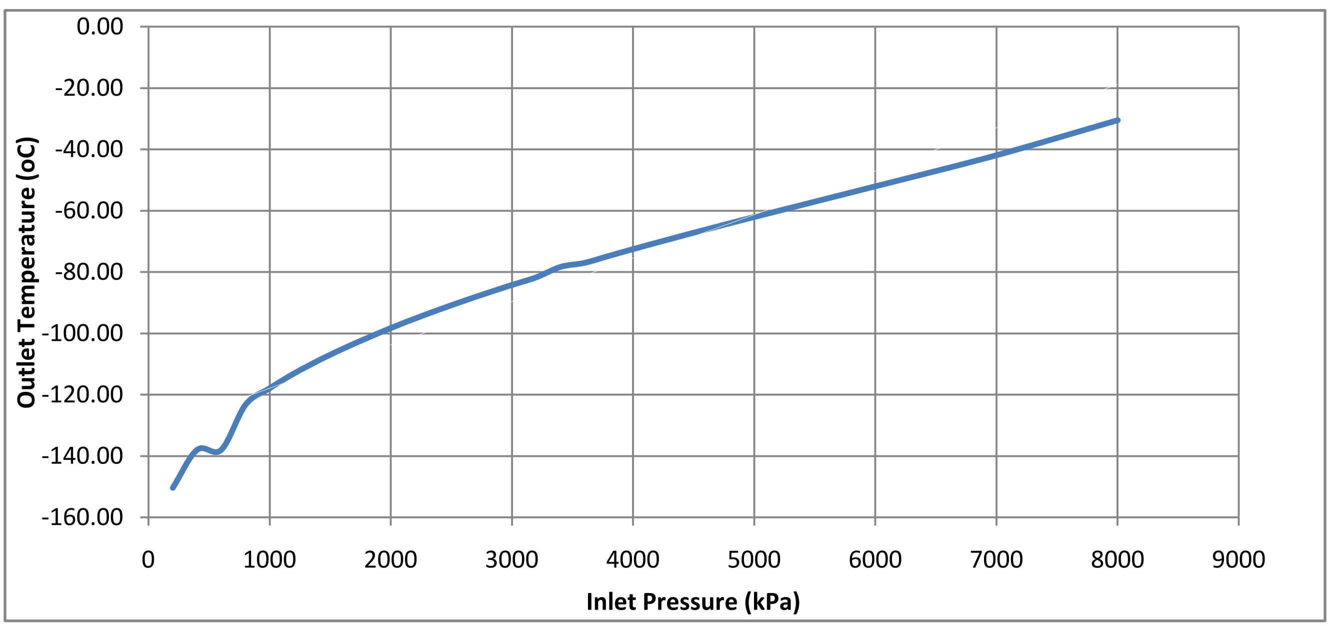


Figure 5.4: NG Feed Outlet Temperature against Inlet Pressure of the LNG Heat Exchanger

The plot on Figure 4.4 above helps in determining the feed pressure necessary to achieve a given temperature into the fractionators which invariably has an effect on the temperature of the top and bottom products from the fractionators. The resulting mathematical model provides a framework for meeting product specification downstream the liquefaction section by varying the feed inlet pressure into LNG heat exchanger, which in this study could be called the main cryogenic heat exchanger (MCHE).

The mathematical model: $\mathbf{Y = 0.014x - 132.1}$ and $\mathbf{R^2 = 0.936}$

Where Y is the outlet Temperature of the LNG heat exchanger and X = the inlet pressure into the LNG heat exchanger and R^2= Coefficient of Determination.

5.2 The Fractionation Segment

The products of fractionators are just okay with yield over 85% and these are similar to those obtained elsewhere. Thus makes it a veritable tool for efficient utilization.

Effect of natural gas feed pressure into LNG heat exchanger on the temperature of fractionator product.

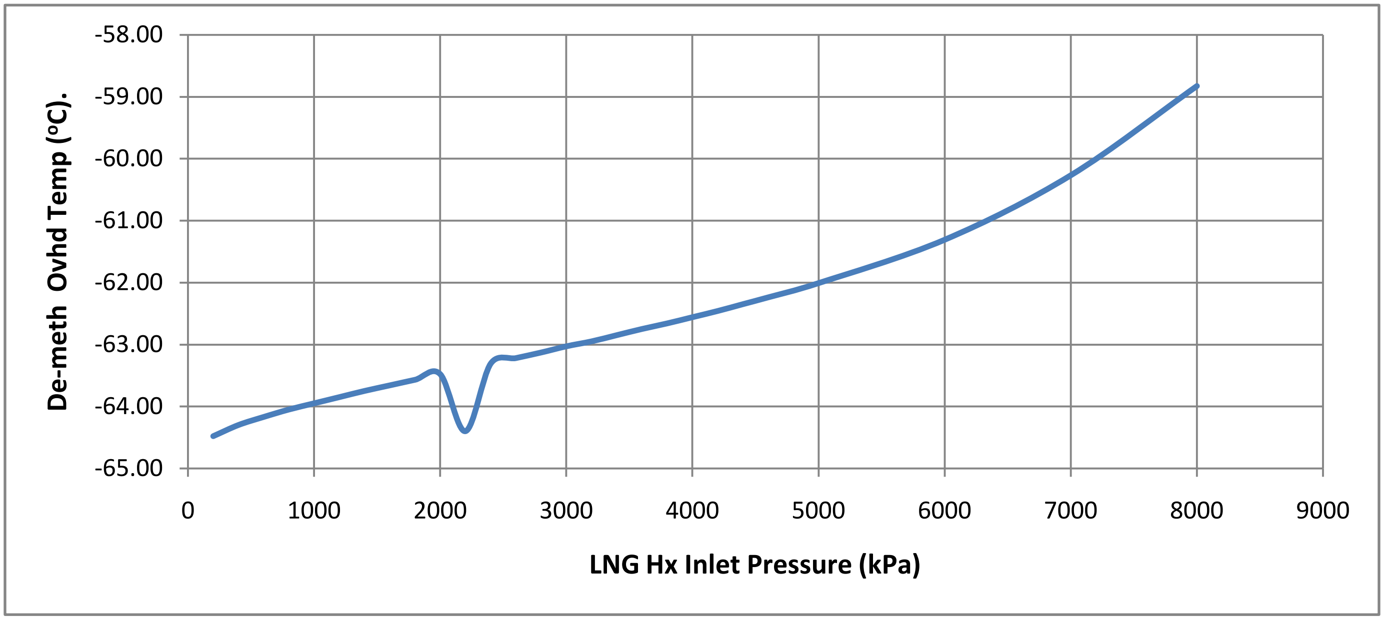


Figure 5.5: Plot of Natural Gas Feed Pressure into LNG Heat Exchanger on the Temperature of the De-Methanizer Overhead Product

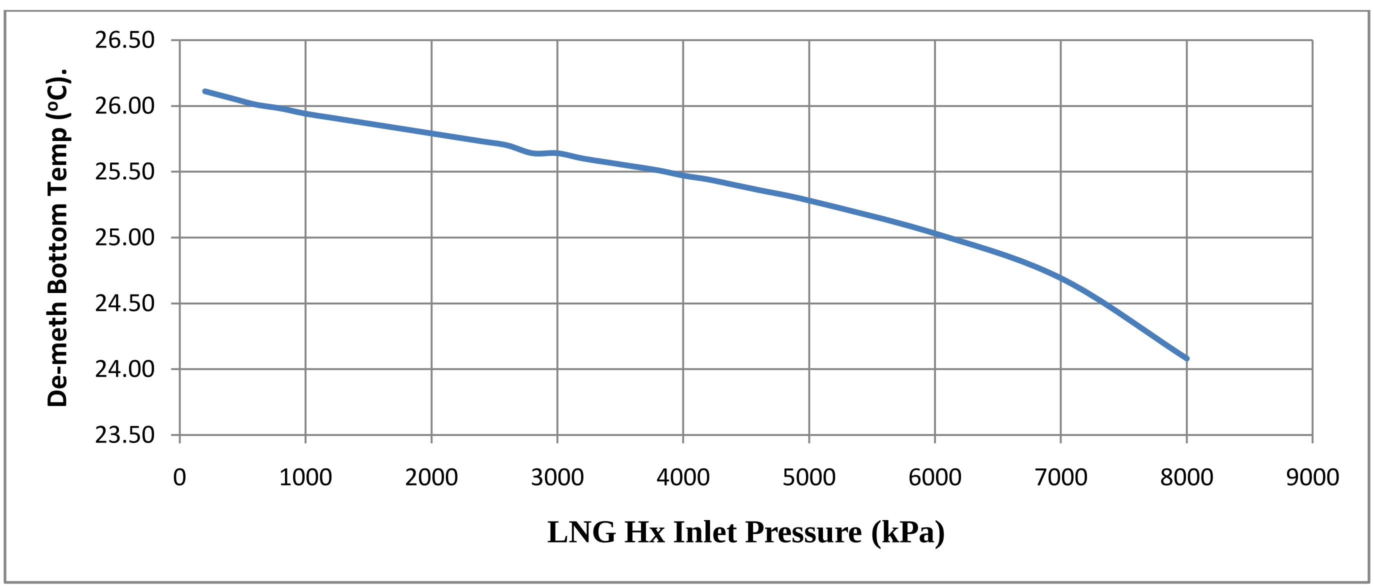


Figure 5.6: Graph of Natural Gas Feed Pressure into LNG Heat Exchanger on the Temperature of the De-Methanizer Bottom Product

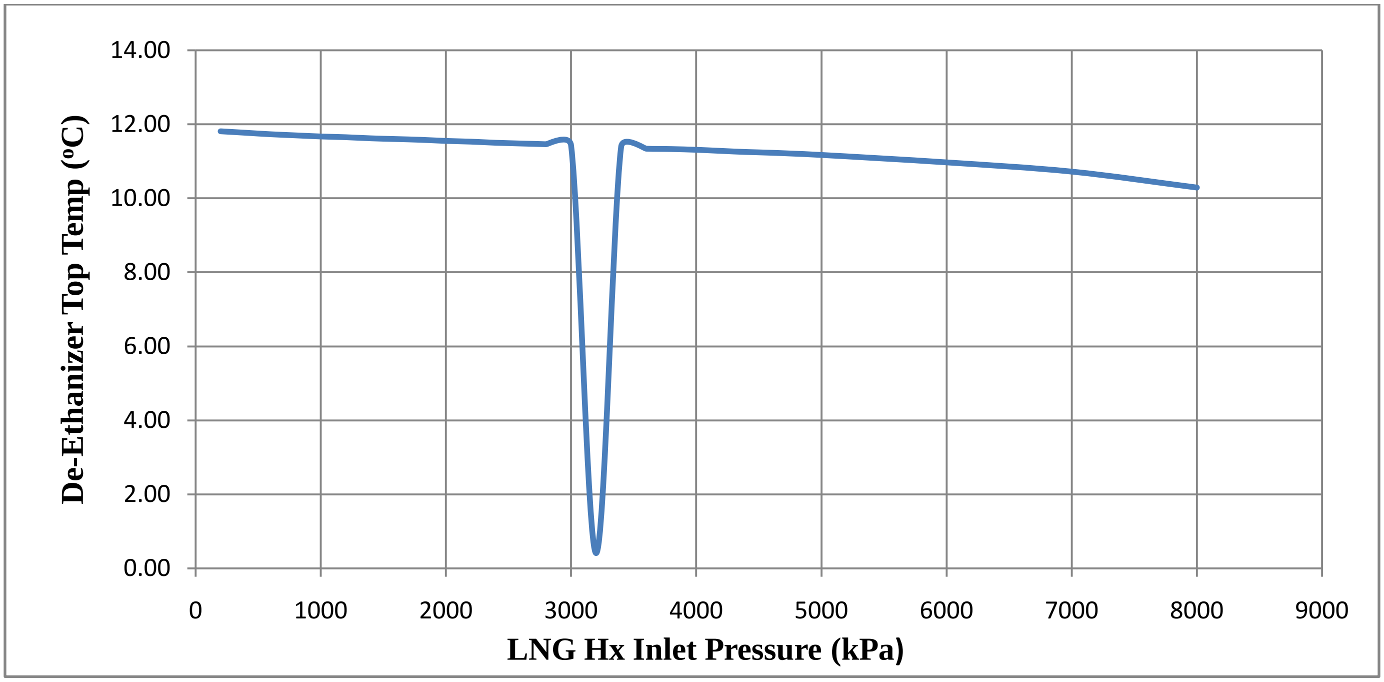


Figure 5.7: Graph of Natural Gas Feed Pressure into LNG Heat Exchanger on the Temperature of the De-Eethanizer Overhead Product

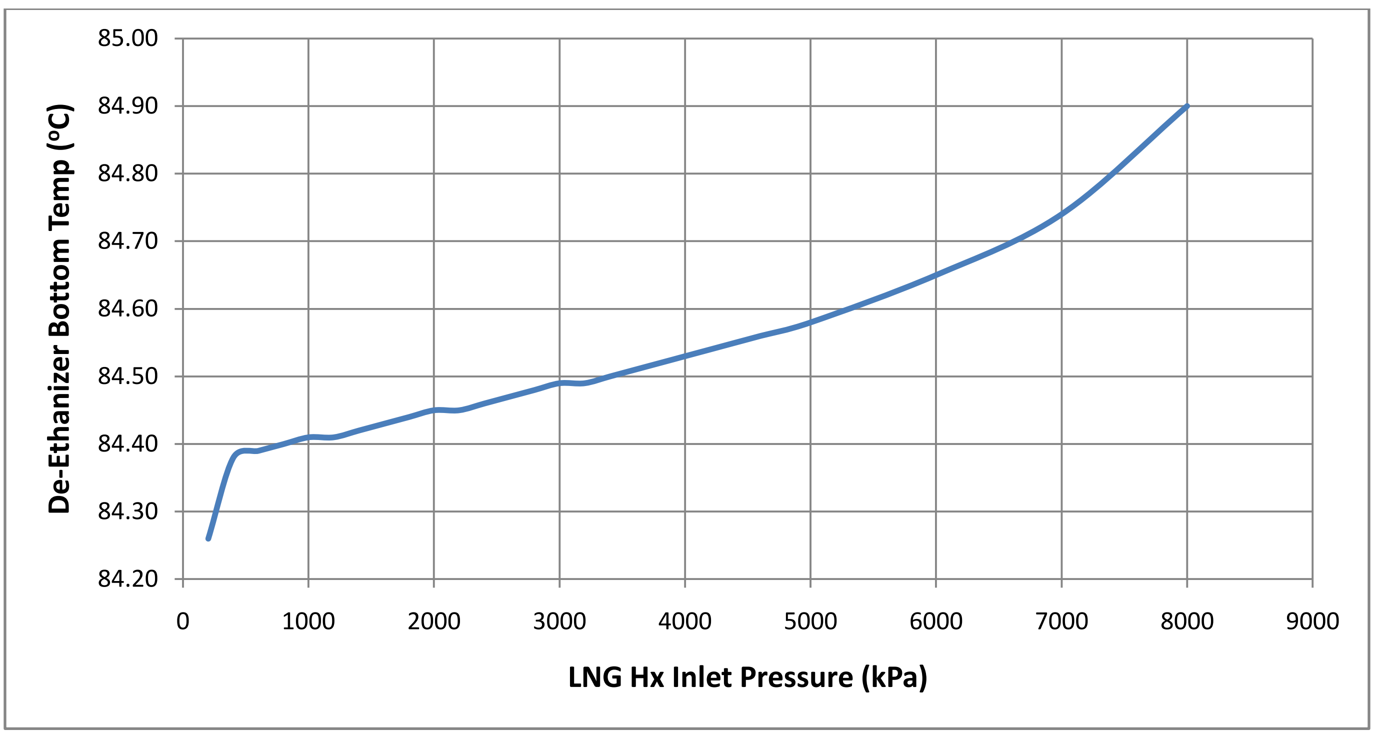


Figure 5.8: Graph of Natural Gas Feed Pressure into LNG Heat Exchanger on the Temperature of the De-Eethanizer Bottom Product

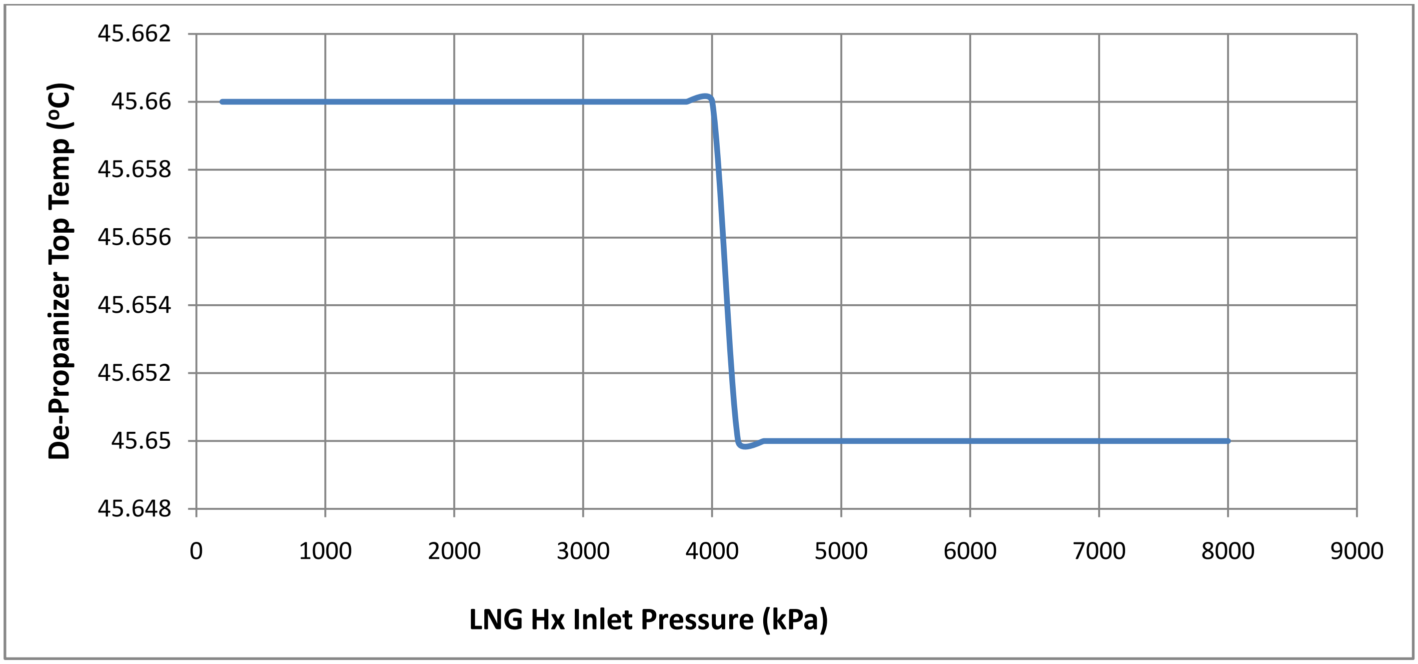


Figure 5.9: Graph of Natural Gas Feed Pressure into LNG Heat Exchanger on the Temperature of the De-Propanizer Overhead Product

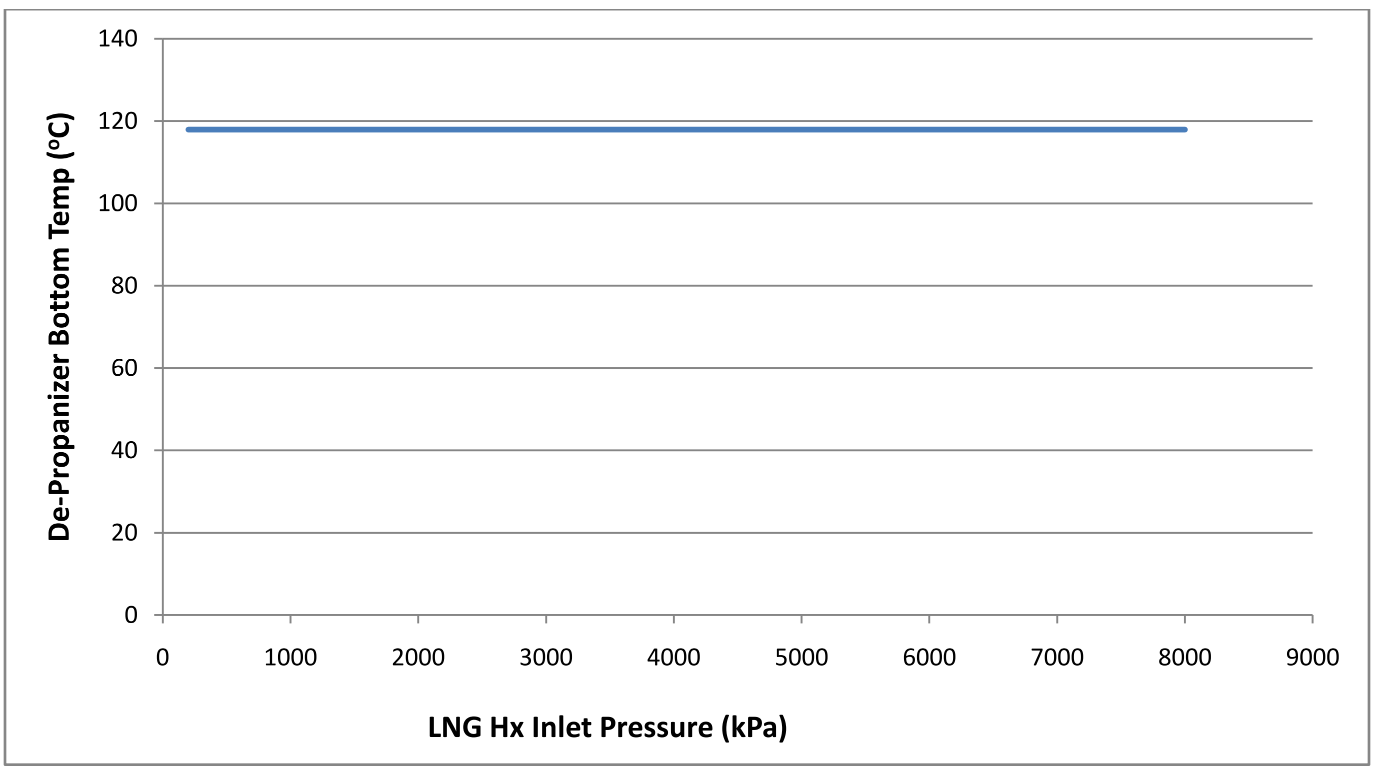


Figure 5.10: Natural Gas Feed Pressure into LNG Heat Exchanger versus Temperature of the De-Propanizer Bottom Product

Figure 5.5 to 5.10 illustrates a linear relationship between the natural gas feed pressure on the fractionators top and bottom product temperatures. It can also be deduced that as the fractionation proceeds, the temperature of the fractionator products is not affected by the feed inlet pressure (or temperature) as observed in Figure 5.9 and 5.10 above. We can also conclude here that from this applies to the de-propanizers, de-butanizers etc. The graph enables the engineer/ researcher to choose an optimal feed pressure to achieve a particular outlet temperature spec for further downstream handling/ utilization.

VI. CONCLUSION

The results show that the molar flow rate of propane in refrigerant in a simple cascade refrigeration model has a direct proportionality to the power consumed by the compressor and the heat flow across the compressor. The molar flow rate of the propane refrigerant is with a linear relationship with the heat duty of the condenser.

The inlet pressure of the natural gas feed has a linear relationship on the downstream products (bottoms and tops), however, at the third fractionator, this effect is seen to diminish. So, we can conclude that after the second fractionators, the increase in feed pressure doesn't affect the downstream products i.e propane, butane and other heavies.

VII. RECOMMENDATIONS

- ✓ This result from this research is a viable tool to designing a plant for that is energy effective for a good yield of hydrocarbon fraction
- ✓ Similar study should be carried out on other refrigeration technology so as to develop similar models as the one presented in this work.

REFERENCES

[1]. Al-Sobhi, Saad Ali. Simulation and Integration of Liquefied Natural Gas Process. Texas: Office of Graduate Studies,Texas A&M University, 2007.
[2]. Barclay, and Delton. "Thermodynamic Cycle Selection for Distributed Natural Gas Liquefaction." Liquefied Natural Gas Journal, 2005: 34-36.
[3]. British Petroleum (B.P). Stastics Review of World Energy. British Petroleum PLC, United Kingdom: Bob Dudly, Group Chief Executive, 2016.
[4]. Hamidreza, Taleshbahrami, and Saffari Hamid. "Optimization of the C3MR Cycle with Genetic Algorithm." Canadian Society of Mechanical Engineering 34, no. 3-4 (2010): 433-448.
[5]. Meiquan, Wang, Zhang Jian, Xu Qiang, and Kuyen Li. "Thermodynamic Analysis-based Engegy Consumption Minimization for Nautral Gas Liquefaction." Industrial and Engineering Chemistry Research (American Chemical Society) 50, no. 22 (2011): 12630-12640.
[6]. Nigerian National Petroleum Corporation. NNPC Business. August 15, 2018. http://www.nnpcgroup.com/nnpcbusiness/midstreamventures/nigeriangasmasterplan.aspx (accessed August 15, 2018).
[7]. Paradowski, Heni, Mohamed Ould Bamba, and Christian Bladanet. "Propane Precooling Cycles for Increased LNG Train Capacity." 14th International Conference and Exhibition on Liquefied Natural Gas. Cedex, Paris: 92973 Paris La Defense, 2004. 107-124.
[8]. Per E, Whal, and W Sigurd. "Formulating the Optimization Problem When Using Sequential Quadratic Programming Applied to a Simple LNG Process." Computers and Chemical Engineering Journal 82 (2015): 1-12.
[9]. Phebe, Asantewa Owusu, and Asumadu-Sarkodie Samuel. "A Review of Renewable Energy Sources, Sustainability Issues and Climate Change Mitigation." Cogent Engineering 3 (2016).
[10]. Roy, Partho S., and Amin M. Ruhul. "Aspen-HYSYS Simulation of Natural Gas Processing Plant." Journal of Chemical Engineering, IEB 26 (2011): 65.
[11]. Salehi, Bahareh. LNG Plant Modeling and Optimization. Chemical and Petroleum Engineering, Curtin University, Western Australia: Curtin University, 2018.
[12]. Shukri, T. "Liquefied Nnatural Gas Technology Selection." Hydrocarbon Engineering 9, no. 2 (2004): 71-76.
[13]. Yoon, Jong. "Characteristics of Cascade and C3-MR Cycle on Natural Gas Liquefaction Process." International Journal of Chemical, Molecular, Nuclear, Materials and Metallurgical Engineering III (2009): 620-624.

APPENDIX A

Table 4.2: De-Methannizer Overhead Composition

	Mole Fractions	Vapour Phase
Methane	0.8613	0.8613
Ethane	0.1258	0.1258
Propane	0.0040	0.0040
i-Butane	0.0000	0.0000
n-Butane	0.0001	0.0001
i-Pentane	0.0000	0.0000
n-Pentane	0.0000	0.0000
n-Hexane	0.0000	0.0000
n-Heptane	0.0000	0.0000
n-Octane	0.0000	0.0000
Nitrogen	0.0030	0.0030
CO2	0.0058	0.0058

Source: ASPEN-HYSYS Excerpt

Table 4.3: De-Methannizer Bottom Composition

	Mole Fractions	Liquid Phase
Methane	0.0000	0.0000
Ethane	0.5184	0.5184
Propane	0.3907	0.3907
i-Butane	0.0062	0.0062
n-Butane	0.0490	0.0490
i-Pentane	0.0206	0.0206
n-Pentane	0.0115	0.0115
n-Hexane	0.0017	0.0017
n-Heptane	0.0011	0.0011
n-Octane	0.0006	0.0006
Nitrogen	0.0000	0.0000
CO2	0.0000	0.0000

Source: ASPEN-HYSYS Excerpt

Table 4.4: De-Ethannizer Overhead Composition

	Mole Fractions	Vapour Phase	Liquid Phase
Methane	0.0000	0.0000	0.0000
Ethane	0.9437	0.9437	0.8714
Propane	0.0563	0.0563	0.1284
i-Butane	0.0000	0.0000	0.0000
n-Butane	0.0000	0.0000	0.0001
i-Pentane	0.0000	0.0000	0.0000
n-Pentane	0.0000	0.0000	0.0000
n-Hexane	0.0000	0.0000	0.0000
n-Heptane	0.0000	0.0000	0.0000
n-Octane	0.0000	0.0000	0.0000
Nitrogen	0.0000	0.0000	0.0000
CO2	0.0000	0.0000	0.0000

Source: ASPEN-HYSYS Excerpt

Table 4.5: De-Ethannizer Bottom Composition

	Mole Fractions	Liquid Phase
Methane	0.0000	0.0000
Ethane	0.0100	0.0100
Propane	0.7906	0.7906
i-Butane	0.0137	0.0137
n-Butane	0.1077	0.1077
i-Pentane	0.0453	0.0453
n-Pentane	0.0252	0.0252
n-Hexane	0.0038	0.0038
n-Heptane	0.0025	0.0025
n-Octane	0.0013	0.0013
Nitrogen	0.0000	0.0000
CO2	0.0000	0.0000

Source: ASPEN-HYSYS Excerpt

Table 4.6: De-Propannizer Overhead Composition

	Mole Fractions	Liquid Phase
Methane	0.0000	0.0000
Ethane	0.0124	0.0124
Propane	0.9726	0.9726
i-Butane	0.0046	0.0046
n-Butane	0.0104	0.0104
i-Pentane	0.0000	0.0000
n-Pentane	0.0000	0.0000
n-Hexane	0.0000	0.0000
n-Heptane	0.0000	0.0000
n-Octane	0.0000	0.0000
Nitrogen	0.0000	0.0000
CO2	0.0000	0.0000

Source: ASPEN-HYSYS Excerpt

Table 4.7: De-Propannizer Bottom Composition

	Mole Fractions	Liquid Phase
Methane	0.0000	0.0000
Ethane	0.0000	0.0000
Propane	0.0200	0.0200
i-Butane	0.0520	0.0520
n-Butane	0.5195	0.5195
i-Pentane	0.2371	0.2371
n-Pentane	0.1318	0.1318
n-Hexane	0.0198	0.0198
n-Heptane	0.0132	0.0132
n-Octane	0.0066	0.0066
Nitrogen	0.0000	0.0000
CO2	0.0000	0.0000

Source: ASPEN-HYSYS Excerpt

APPENDIX B

Table B1: LNG Heat exchanger Inlet Pressure and Outlet Temperature/ De-methannizer Product Temperature Profile

LNG Heat Exchanger		De-Methanizer Temperature (oC)	
Inlet Pressure (kPa)	Outlet Temp (oC)	Top	Bottom
200	-150.40	-64.48	26.11
400	-138.00	-64.30	26.06
600	-138.00	-64.17	26.01
800	-123.30	-64.05	25.98
1000	-117.90	-63.95	25.94
1200	-113.10	-63.85	25.91
1400	-108.90	-63.75	25.88
1600	-105.10	-63.66	25.85
1800	-101.60	-63.57	25.82
2000	-98.26	-63.48	25.79
2200	-95.16	-64.40	25.76
2400	-92.23	-63.31	25.73
2600	-89.44	-63.22	25.70
2800	-86.77	-63.13	25.64
3000	-84.20	-63.03	25.64
3200	-81.72	-62.95	25.60
3400	-78.32	-62.85	25.57
3600	-76.99	-62.75	25.54
3800	-74.73	-62.66	25.51
4000	-72.51	-62.56	25.47
4200	-70.35	-62.46	25.44
4400	-68.22	-62.35	25.40
4600	-66.13	-62.24	25.36
5000	-62.03	-62.01	25.28
6000	-52.03	-61.31	25.03
7000	-41.88	-60.27	24.69
8000	-30.46	-58.83	24.08

Source: During Sensitivity Analysis

Table B2: De-methannizer Product Temperature Profile

De-Ethanizer Temperature (oC)		De-Propanizer Temperature (oC)	
Top	Bottom	Top	Bottom
11.81	84.26	45.66	117.9
11.77	84.38	45.66	117.9
11.73	84.39	45.66	117.9
11.70	84.40	45.66	117.9
11.67	84.41	45.66	117.9
11.65	84.41	45.66	117.9
11.62	84.42	45.66	117.9
11.60	84.43	45.66	117.9
11.58	84.44	45.66	117.9
11.55	84.45	45.66	117.9
11.53	84.45	45.66	117.9
11.50	84.46	45.66	117.9
11.48	84.47	45.66	117.9
11.46	84.48	45.66	117.9
11.43	84.49	45.66	117.9
0.41	84.49	45.66	117.9
11.39	84.50	45.66	117.9
11.34	84.51	45.66	117.9
11.33	84.52	45.66	117.9
11.31	84.53	45.66	117.9
11.28	84.54	45.65	117.9
11.25	84.55	45.65	117.9
11.23	84.56	45.65	117.9
11.17	84.58	45.65	117.9
10.97	84.65	45.65	117.9
10.72	84.74	45.65	117.9
10.29	84.90	45.65	117.9

Source: During Sensitivity Analysis

Table B3: Single Refriggeration Loop Data

SINGLE REFRIGERATION LOOP			
Propane Refrigerant	Compressor		Condenser
Molar Flow (kgmole/hr)	Power (kW)	Heat Flow (kJ/hr)	Duty (kJ/hr)
0.05	0.0755	271.8	735.1
0.10	0.1510	543.6	1470.0
0.15	0.2265	815.3	2205.0
0.20	0.3020	1087.0	2940.0
0.25	0.3775	1359.0	3675.0
0.30	0.4530	1631.0	4410.0
0.35	0.5285	1902.0	5146.0
0.40	0.6040	2174.0	5881.0
0.45	0.6795	2446.0	6616.0
0.50	0.7549	2718.0	7351.0
0.55	0.8304	2990.0	8086.0
0.60	0.9059	3261.0	8821.0
0.65	0.9814	3533.0	9556.0
0.70	1.0570	3805.0	10290.0
0.75	1.1350	4077.0	11030.0
0.80	1.2080	4348.0	11760.0
0.85	1.2830	4620.0	12500.0
0.90	1.3590	4892.0	13230.0
0.95	1.4340	5164.0	13970.0
1.00	1.5100	5436.0	14700.0
2.00	3.0200	10870.0	29400.0
5.00	7.5490	27180.0	73510.0
10.00	15.1000	54360.0	147000.0

Source: During Sensitivity Analysis