B.Tech thesis on

Simulation analysis of a Propane Recovery Plant using ASPEN PLUS

In partial fulfilment of requirements for the degree of

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(Chemical Engineering)

Submitted by

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CERTIFICATE

This is to certify that the thesis entitled “**Simulation analysis of a Propane Recovery Plant from using ASPEN PLUS**”, submitted by **Sanjana Anand Choudhary (111CH0455)**, to National Institute of Technology Rourkela is a bonafide report of the research work carried out by her, in the Department of Chemical Engineering, National Institute of Technology, Rourkela under my supervision.

The thesis submitted is worthy of consideration for the partial fulfilment of award of the degree of “Bachelor of Technology” in accordance with the regulations of this Institute. The results incarnated in the report are authentic to the best of my knowledge

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ABSTRACT

Natural gas is a good source of various hydrocarbon compounds, but, alkanes with three or more number of carbon atoms in there chain are of greater value as fuels as well as pure components. A cryogenic plant is setup to recover propane, butane, and other heavier alkanes with two operating distillation columns; de-methanizer and de-ethanizer. The available software Aspen Plus is used to simulate this cryogenic plant to serve the purpose of recovery .This software facilitates us to infer the impact of the operating variables on the efficiency of the plant. The use of cryogenic plants for this purpose is a latest project carried on just by a few companies, so the idea here is to analyse the operating variables to optimize the recovery.

Keywords: turbo-expander; gas sub-cooled process; cryogenic plant;

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OVERVIEW OF THE PROJECT

Ever since the recognition of natural gas being used as a desirable fuel, the prerequisite for its transportation to industries has led to the enforcement of treating and processing technologies. All over the world, the gas processing market meets a wide variety of economic challenges. These may be for merely fulfilling a gas conveyance arrangements to acquiring merely high ethane recovery for providing feed to other plants. Similar to gas usage, local industries for natural gas liquids (NGL), and infrastructure vary, the contaminants and inlet conditions of the natural gas to be processed may vary, too. Every undergoing project must meet specific norms for cost-effectiveness depending on the markets for gas and NGL products and the combination of characteristics of the available gas. The NGL recovery generally falls in following three parts: ethane-plus recovery, propane-plus recovery, and flexible ethane rejection or recovery.

Propane is a by-product of domestic **natural gas processing**, or by production from crude oil refinement. Due to the advantages of propane, of being a cleaner fuel which can be used for almost all purposes, that is from running vehicles to household activities, it has been on a rising demand. Nearby, 48 million of population in the U.S. use propane on a daily basis. Various Propane Recovery plants have been set up in the last decade to provide the world with a source of cleaner energy.

Objective of this work is to provide an analysis of a propane plus recovery plant from natural gas. Analysing different methods and their outputs, giving review for the most suitable methods. Carrying the simulation of this process, and checking its dependence on various operating parameters, the optimum conditions are found using ASPEN plus software.

CHAPTER 1 – INTRODUCTION AND LITERATURE SURVEY

1.1. Natural gas

Natural gas is formed after the decomposition of dead plants and animals that have been deposited centuries ago beneath the rocks due to some natural calamities. These submerged beds are also a rich source of various other hydrocarbons, coal and petroleum. Natural gas contains organic as well as inorganic compounds. Cleaner than petrol and diesel, it is the cheapest form of fuel that is available to the humankind. In vehicles it is used in compressed form as CNG; used for household purposes, aviation, generation of power, transportation, production of hydrogen and other hydrocarbons.

Inlet gas properties including its pressure, degree of water saturation, heavier liquid content, and other contaminants can affect the selection of the appropriate technology to treat gases. Production objectives such as liquid and gas product specifications, liquid recovery percentages, and type of liquid products to be also produced have significant impact.

1.2. Propane recovery

Natural gas has a maximum composition of methane (89%), and considerable ethane (6-7%). Hence, the rejection of methane and ethane from the natural gas would serve the purpose of propane plus recovery. This is done by employing gas processing plants which are, Absorption plants and Cryogenic plants. As the names suggest, cryogenic plants involve refrigeration; absorption plants requires action of absorbent on the gas. These plants have different efficiencies, advantages and disadvantages which are discussed in the following section.

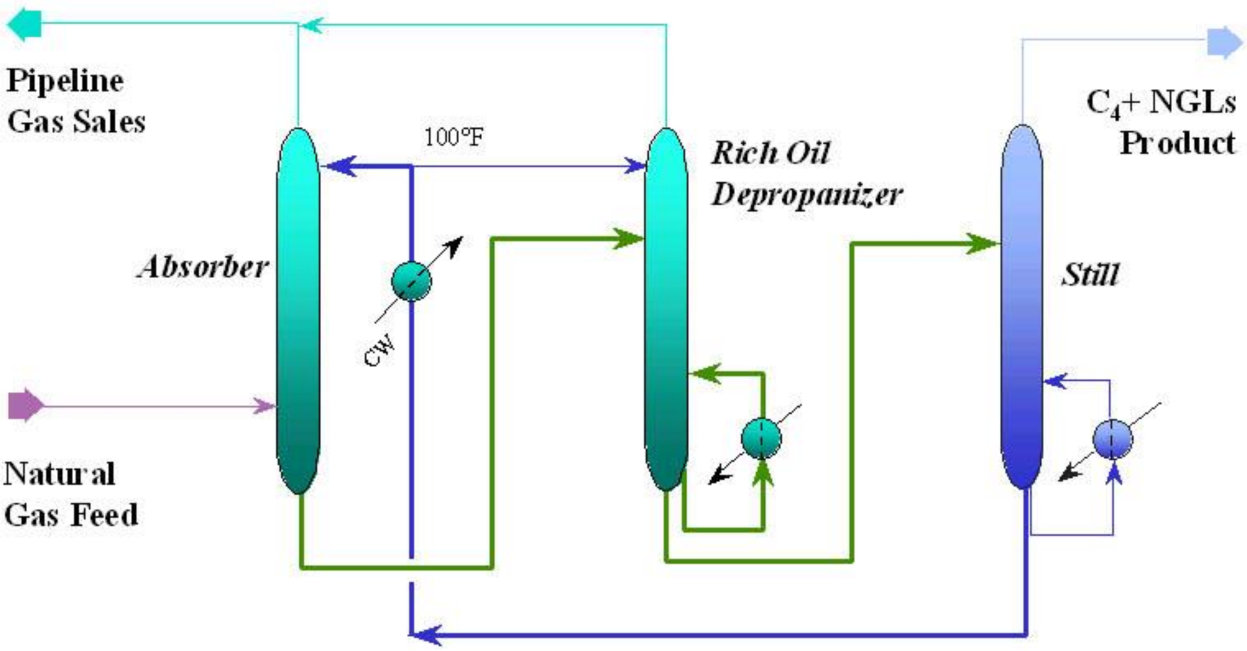

Figure 1.1: Absorption plant (ADVANCED EXTRACTION TECHNOLOGIES, INC)

1.3. Cryogenic Plant Process

In an absorption plant as shown in figure 1.1, two towers are employed. They use an oil absorbent. The first tower is used for absorption of propane plus components in the oil absorbent and thereby recovered, while the second is used for the regeneration of the absorbent to get bottom product as lean solvent and top product as the required propane plus components. This is a continuous process, hence, we have a recycle stream containing lean solvent, to the first tower. This plant can get a recovery of 70-90%. Most of the part of the capital is spent on equipment and still for oil absorber plant.

Cryogenic plants work by refrigerating the gases, liquefying almost all propane and other heavier hydrocarbons. A series of fractionating columns can be used to separate the liquids according to the requirement. They have less process equipment and more mechanical pressure changers. The recovery is in the range of 90-98%. Hence, more profitability is ensured in the cryogenic process plant set up as it gives better recovery and is cheaper than latter. The cryogenic methods employing turbo-expanders were introduced mainly for ethane recovery, the residue-reflux systems improves this system to avail propane-plus recovery in the two tower systems, such developments add to the original GSP process .

1.4. Cryogenic Processing Gas Technologies

Methods given by Ortloff are Improved Overhead Recycle Process, Gas Subcooled Process, Split-flow Reflux Process, and the Single Column Overhead Recycle Process.

a) Gas subcooled process (GSP)

It is one of the methods developed by Ortloff during the 1970s [13] to attain maximum efficiency in propane plus recovery. In this process, the feed is subcooled and condensed, and is flashed to the operating pressure of the fractionating column. It is then introduced to the column as its top feed, this acts as reflux. As the reflux comes in contact with the vapour, it rectifies the vapour by absorbing the propane plus components, another part of the feed gas is

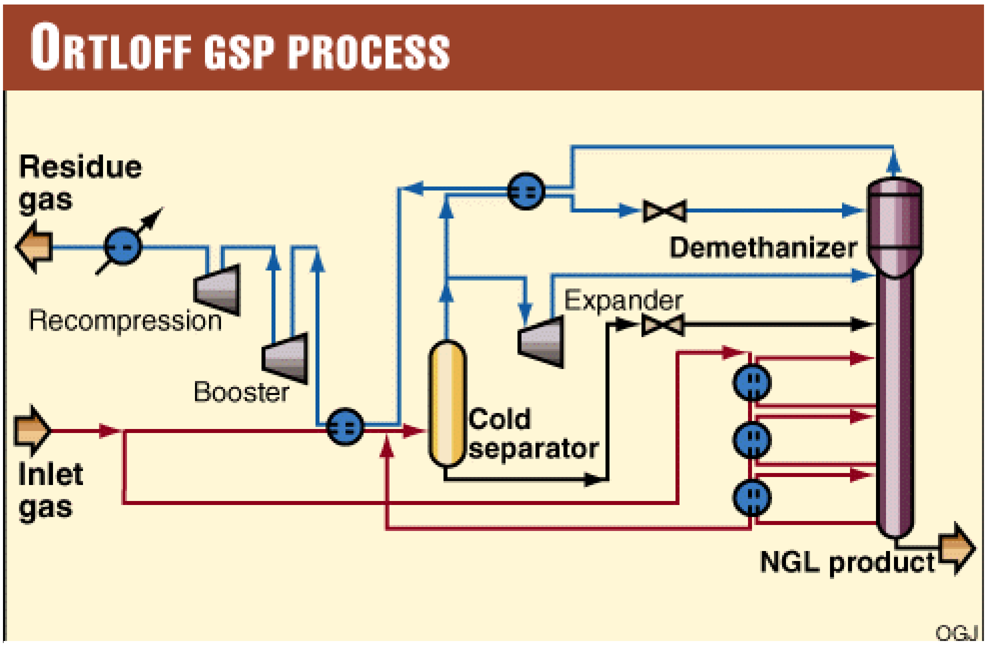

Figure 1.2: GSP PROCESS (Ortloff Engineers ltd)

also expanded and is fed to the tower as intermediate feed. Fig. 2 shows a sketch of a plant based on the Gas Subcooled Process.

b) Single Column Overhead Recycle Process

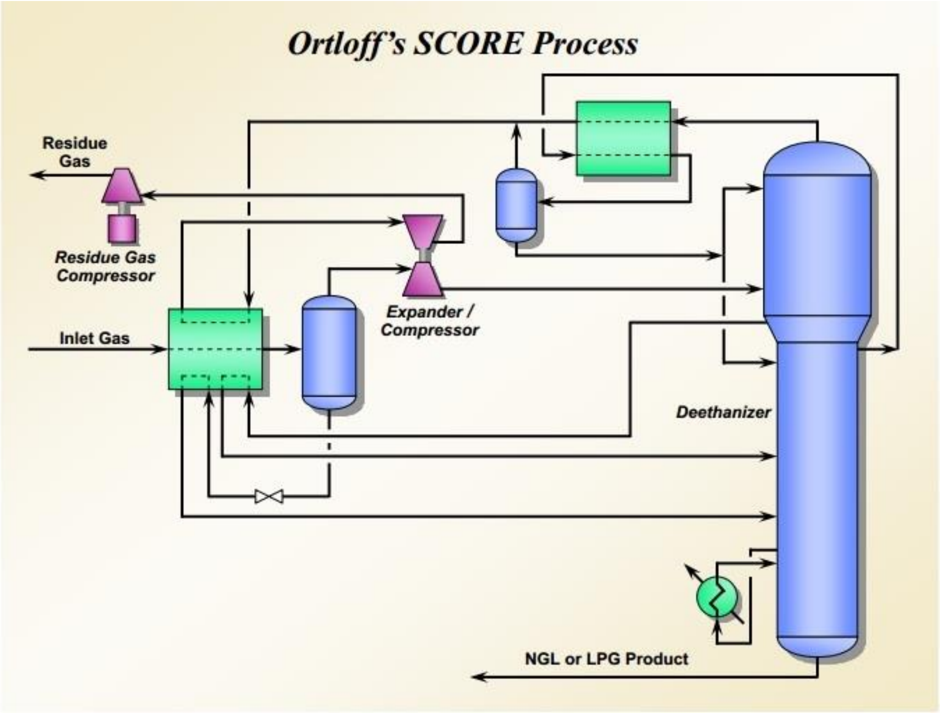

Figure 1.3: SCORE process (Ortloff Engineere ltd.)

A vapour side draw stream is condensed and generated as reflux for the column. To reduce the heat input, a liquid side cut is used for the process cooling. With appropriate design features like adding equipment or increasing piping, a plant using the SCORE process can also be switched to work in ethane recovery mode utilizing Ortloff's Gas Subcooled Process (GSP)[2]. Major advantage is that this process is efficient for propane recovery with minimum compression requirements, while discarding lighter components to meet specifications of the liquid product. Gas processing plants where 40-60% recovery of ethane may be desirable without the addition of any equipment. In normal operation, the product given by SCORE (fig 3) process is a mixed LPG product stream, containing a maximum ethane in propane liquid product specification. When converted to GSP mode, a mixed NGL product stream is produced, contains a maximum methane in ethane liquid product specification.

Overhead recycle process is efficient for high propane recovery, but it is not suitable for high ethane recovery. These methods have a common limitation that they depend on the composition of vapor stream that becomes the reflux for the de-methaniser column.

1.5. Dual Column Propane Plus Recovery

This is a general dual column cryogenic process which gives more than 98 % recovery of propane as well as 90-95 % recovery of ethane. This has been referred from TECHNIP licensor named as CRYOMAX DCP, i.e., Dual Column Recovery [3] Feed gas at 298K, 70 bars cooled to 243K in a series of exchangers. The separator separates liquid and vapor. Expander is used to expand the cooled high pressure gas to 30 bars, and the resulting steam is used to feed demethaniser. Liquid

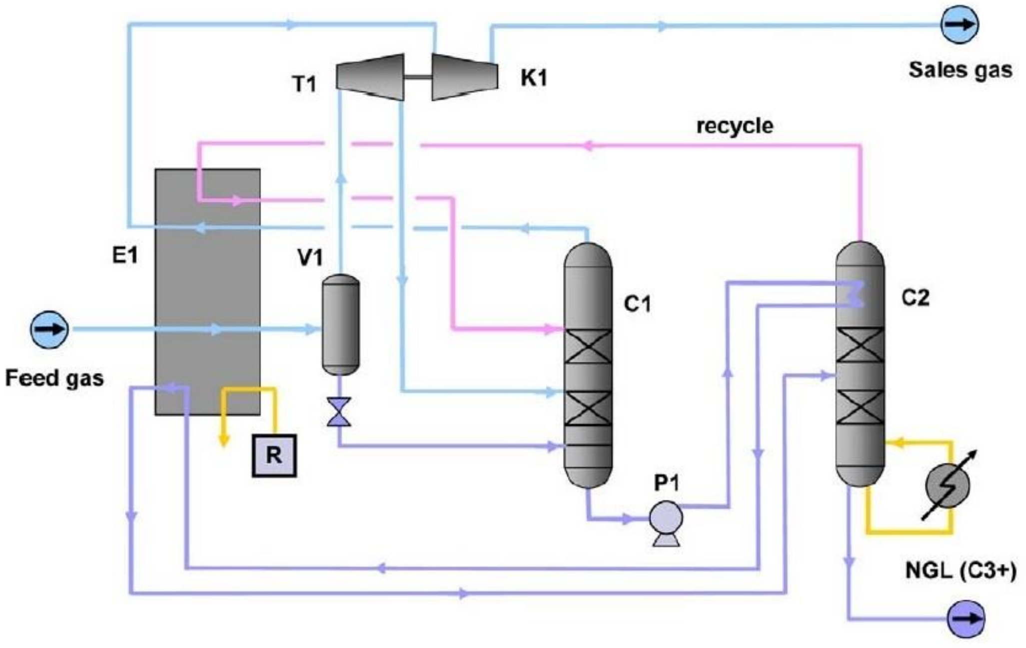

Figure 1.4: DCP CRYOMAX (TECHNIP licensors)

from separator is also treated as feed to demethaniser bottom. Liquid from demethaniser is pumped to 33 bars &is reheated to 293K to feed de-ethanizer. A vapor distillate having high concentration of ethane is given by de-ethanizer as a product, this stream gets liquefied in series of exchangers & sent as reflux. At 30 bars, treated gas is compressed to sales gas in compressor.

1.6. Previous works

D.Fissore et. al. [4] presented the simulation considering a cryogenic plant for a gas subcooled process. They used the separator temperature, split fraction of the overhead output stream from the separator and the pressure condition in the demethaniser and de-ethaniser column, and observed their dependence on the exchanger duties and compressor power to analyse the optimised cost consumption and they found optimised operating conditions ,i.e. , 0.1 split fraction of the overhead stream of the separartor is found optimum for more propane recovery when the temperature of the flash separator is maintained at -50°C and the column pressure is as low as 9.5 bar.

1.7. ASPEN PLUS

ASPEN PLUS allows you to create your process model, starting with the flow sheet, then specifying the chemical components and operating conditions. It helps in understanding the data required for the feasibility of the process, making us aware about various outcomes possible for a system. After the completion of required inputs in the flow sheet, the simulation is done. This software displays the results stream wise and block wise.

CHAPTER 2 – PREREQUISITE INPUT AND FLOW SHEET MODELING

2.1 Natural Gas Composition

Table 2.1: Natural Gas Composition (Cawthorne Channel)

Component	Mole%
Methane	89
Ethane	7
Propane	3.077
i-Butane	0.2
n-Butane	0.2
n-Pentane	0.18
n-Hexane	0.15
n-Heptane	0.1
n-Octane	0.09
n-Nonane	0.002
n-Decane	0.001
Other gases	Traces

Table 2.2: Natural Gas Composition

Component	Mole%
Methane	80
Ethane	5
Propane	3.077
i-Butane	0.2
n-Butane	0.2
n-Pentane	0.18
n-Hexane	0.15
n-Heptane	0.1
n-Octane	0.09
n-Nonane	0.002
n-Decane	0.001
Other gases	Traces

Table 2.2 gives a typical range of values of composition of natural gas. In Table 2.1, we see a more particular set of composition obtained from the Cawthorne Channel- Degema Local Government Area of Rivers State, Nigeria. This is the set of data that we are going to consider for our further study.

2.2 Process Flow sheet

After studying various methods that have been introduced for efficient propane and ethane recovery, a cryogenic plant equipped with dual column is studied to extend further, our objective of optimum recovery of propane plus as well as ethane. This study is in accordance with the study of D,Fissore et. al ,so, considering the flow sheet given by them. The flow sheet is shown in figure 1.4[3].

2.3 Columns and equipment

As per the flow sheet, various types of columns and equipment are used. The number of columns is reduced as compared to an absorption plant but we have more of compressor, expander and pipeline. The selection of each equipment requires some criteria, which help us to understand the stream conditions better. Equipment used here are mentioned as follows from Figure 1.4:-

a. Pressure Changers: three COMPRESSORS, valves;
b. Distillation Column: two RADFRAC columns ;
c. Exchangers: HEATER, three in number and HEATX, two in number ;
d. Separator: one column for FLASH separator.

2.4 Prerequisite data

For such a complex flow sheet of a plant we need a lot of data as per the number of streams and the number of equipment used. The conditions help us in modelling of the plant. Some of these conditions are considered [4] in order to fulfil the input requirements for the aspen plus flow sheet. These data are considered constant for all the simulations we have done.

TABLE 2.3: Specified Data for Distillation Column(D. Fissore et al)

Demethaniser : T-001	
Number of equilibrium stages	15
Pressure	13 bar
Feed streams	
14	Stage 1
13	Stage 1
6	Stage 7
22	Stage 15
De-ethaniser : T-002	
Number of equilibrium stages	6
Pressure	13 bar
Feed stream	
21	Stage 1

Table 2.4: Specified Parameters of Exchangers(D. Fissore et al)

Exchanger ID	**Pressure Drop (bar)**	**Outlet Temperature (deg. C)**
E001	0.7	200
E002	0.3	-
E003	0.3	-50
E004	0.3	-50
E005	0.3	-110

Table 2.5: Specified Parameters of Compressors(D. Fissore et al)

Compressor ID	**Discharge Pressure (bar)**
C001	13
C002	55

2.5 Description of process

In the figure 1.4, initially, the feed is introduced at a temperature of 310K and 55 bar to the splitter. The split fraction of the stream is not specified, assuming it is 0.5 equal flow is recorded in stream 2 and 7. Stream 2 is used to provide a heating medium for the reboiler E1. It is then cooled, through E1, to feed the separator V1. Stream 7 is also cooled, due to heat exchange with stream 15 to the inlet temperature of V1, through E1. E1, is a series of exhangers used to as heating and cooling mediums, V1 is a flash separator, which gives stream 6, liquid from the bottom, and stream 9 overhead gaseous stream from the top, as products. The observed distillation columns C1, C2 are used to reject methane and ethane respectively. So, according to their purpose, they are named as de-methaniser (C1) and de-ethaniser (C2). C1 not only rejects methane but a good amount of ethane too. Stream 6 is fed to C1 whereas stream 9 is again introduced in a splitter. Again, it is assumed that the split fraction is 0.5.

The turbo-expander T1-K1 is fed with the stream 10 from splitter, here a fraction of the gas is liquefied due to the removal of heat of the stream. Reflux condenser (E1) has the key purpose to subcool the expanded (from the valve) Stream 11 to C1 pressure and fed to C1 column.

The product stream from K1, stream 14, is a gas–liquid mixture which is fed to C1. Stream 21 is a liquid product which is free from methane, charged in the de-ethaniser (C2), while the vapour from de-ethaniser (stream 22) is fed back to the C1 bottoms. Stream 25 gives the final product as a de-ethanised liquid product from C2. The overhead vapour stream of C1, provides cooling medium in E1, after which it is fed to T2, where it is given a pressure boost and is driven by the expander K1. Sales gas compressor can be added that provides a pressure boost to the stream 17 to sales gas line pressure.

2.6 Challenges

The plant discussed above is a very complex plant set up for an aspen simulation. The project of recovery of propane plus at large scale is a budding and a very competitive one, at present in the oil and gas industry. Hence, many prerequisite data are not yet been publicised. This problem had been a major one in running the simulation. More than 250 simulations have been run to finalise the values of the remaining unavailable parameters with the help of references. The details of the process in the description helps in finding suitable values to the variables to run this plant to find the optimum range. Input Data after the approximations are specified in following table-

Table 2.6: Assumed data for other equipment of the plant

Equipment	**Parameter**	**Value**
Splitter ; S1	Split Fraction	0.5
Splitter ; S2	Split Fraction	0.5
Separator ; V1	Temperature	223 K
	Pressure	61 atm
Demethaniser; T1	Condenser	Partial vapour
	Distillate Rate	6000 kmol/h
	Reflux Ratio	4
	Condenser Pressure	15 atm
De-ethaniser; T2	Condenser	Partial vapour
	Distillate Rate	5000 kmol/h
	Reflux Ratio	4
	Condenser pressure	14 atm
Compressor; C003	Discharge pressure	55 atm
Exchanger; E3	Pressure	19 atm
Exchanger; E5	Pressure	25 atm

Minimum temperature approach of the exchanger E2 is set equal to 2 degrees and for other exchangers it is set to 1 degree. The convergence of tear stream is set equal to 10^{-6}.

2.7 Aspen plus flow sheet

Finally, modelling of the plant is done to obtain the required results and verified if not in agreement.

The product should contain more than 95% propane plus hydrocarbons[6] . The overhead stream contains rejected methane and ethane. Hence, the overhead stream of Demethaniser should contain 99% of methane[4] and 10-15 % of ethane and the overhead stream of de-ethaniser should contain 95-98% of ethane[8] . The sole purpose of the flow sheet is to maximise the recovery of propane and heavier hydrocarbons as well as ethane and methane.

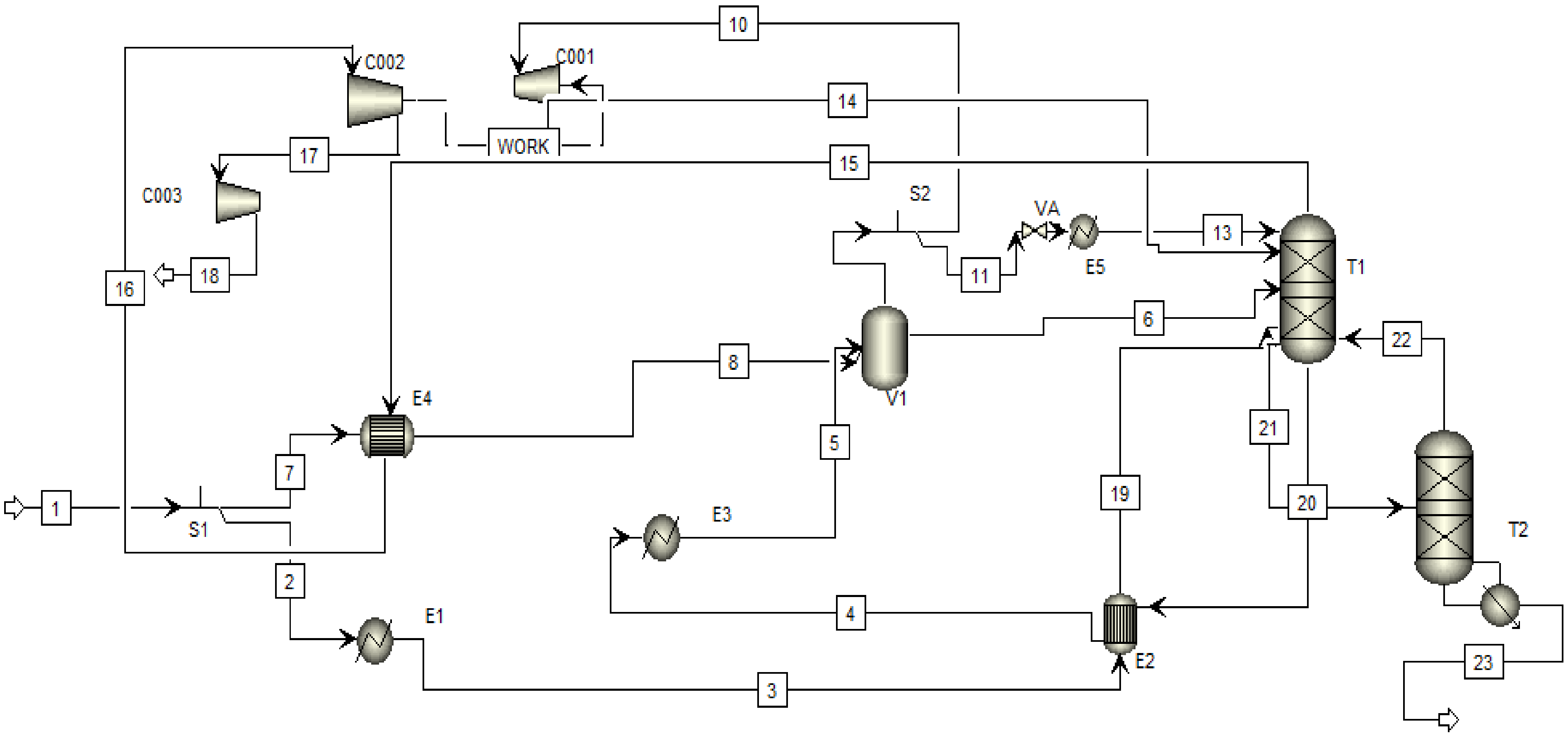

Figure 2.2: ASPEN PLUS flow sheet

CHAPTER 3 - RESULTS AND DISCUSSION

3.1 Stream Results

Properties	**Stream 1**	**Stream 23**
Mole Flow kmol/hr	**MIXED**	**VAPOR**
METHANE	8900	85.465
PROPANE	307.7	17199.46
ETHANE	700	8.22E+05
iso-BUTANE	20	16528.64
n-BUTANE	20	1654.875
n-PENTANE	18	1854.658
n-HEXANE	15	0
n-HEPANE	10	0
n-OCTANE	9	0
n-NONANE	0.2	0
n-DECANE	0.1	0
WATER	0	0
Total Flow kmol/hr	10000	8.60E+05
Total Flow kg/hr	1.84E+05	3.80E+07
Total Flow l/min	77813.93	6.81E+07
Temperature K	310	289.7056
Pressure atm	54.28078	12.7465
Vapour Fraction	0.9950546	1
Liquid Fraction	4.95E-03	0

Table 3.1: Feed- Product Stream Results

These stream results are in accordance with the expected product output. Amount of propane in stream 23 is 97% of the total stream and ethane is 0.0154.

3.2 Plant Performance Analysis

These are software generated results and can be considered for the further analysis of operating variables and dependence of heat duty, cost optimization and plant output on these variables. Three variables are considered for the purpose of investigation of the plant performance to maximise profitability. These variables have been chosen on the basis of their effects on other streams and to what extent it is affecting the output, also these are the parameters that can be changed using controllers in the plant. These variables are:-

- Separator (V1) Temperature
- Split fraction of splitter S2
- Pressure of the Demethaniser (T1) and de-ethaniser (T2)

When these parameters are varied, we need to record variables that will let us understand the impact of the variations on the plant and its yield. After analysing the changes in the data and their variations, it is been understood that following variables should be recorded:-

- Percentage of recovered propane,

- Exchanger duty for E1, E3 and E5.

At the operating conditions at which stream results are calculated these variables are found to be as follows:-

Table 3.2: Output of considered parameters

Property	**BHP C003**	**Exchanger Duty E1**	**Exchanger Duty E3**	**Exchanger Duty E5**
Result	344 MW	132 MW	-82 MW	-49.5 MW

These results are rounded to the closest integer. The negative sign signifies the requirement of refrigeration in these exchangers.

BHP C003 signifies the brake horse power of the Compressor as it consumes most of the electricity. It is the amount of power generated by the compressor or motor without considering its auxiliary parts which might slow down the actual speed of the compressor.

a. Effect of split fraction S2

As we increase the split fraction of stream 9, flow rate in stream 11 to the demethaniser column is increased which increases the concentration of the liquid in the tower. This leads to increase in the molar fraction of ethane in the product stream, thereby becoming less efficient for propane recovery. It makes the column T1 and T2 less efficient. Increased flow in stream 10 also leads to increased duty of E5. The exchanger duty of reboiler E2 is increased as well, in order to overcome the heat duty required the flowrate in stream 2 had to be increased which leads to an additional heat duty of E1. As flow rate in stream 10 is decreased due to increased flow in stream 11, this decreases the amount of compression.

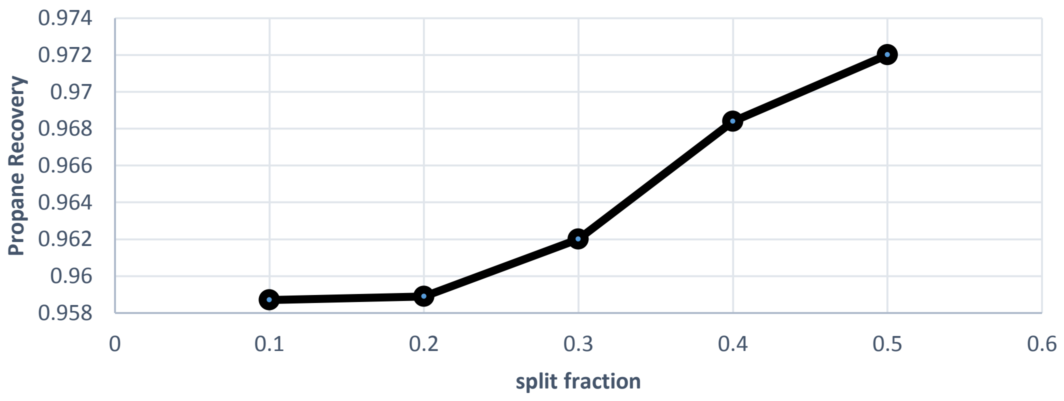

Figure 3.1: Propane Recovery vs Split fraction of S2

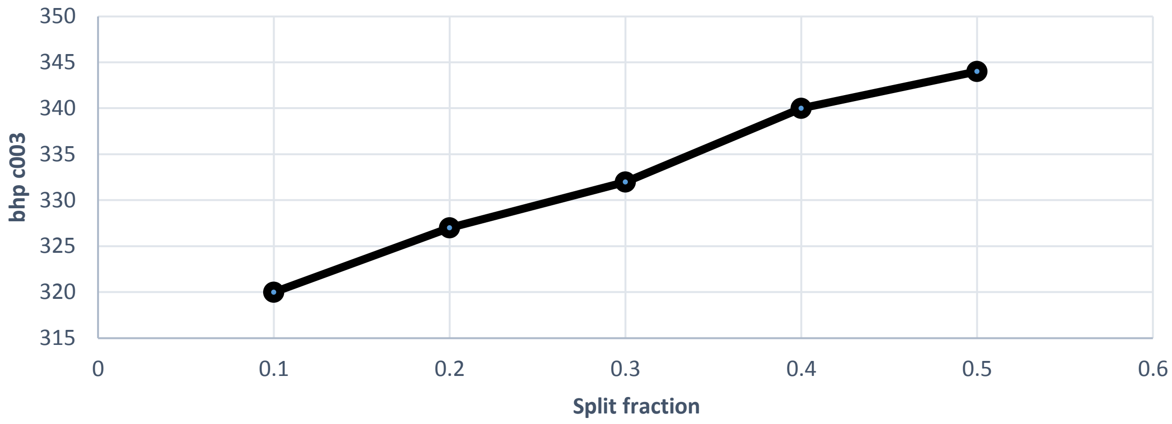

Figure 3.2: BHP C003 vs Split fraction of S2

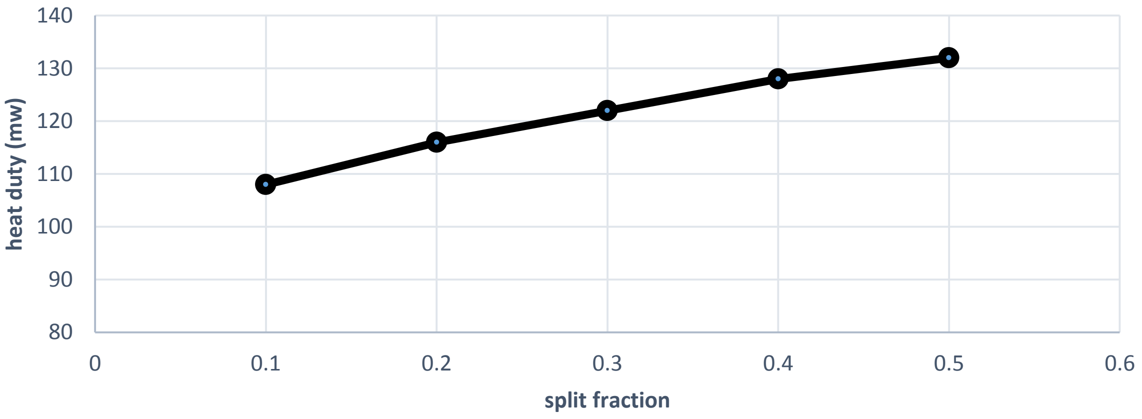

Figure 3.3: Heat Duty of E-001 vs Split fraction of S2

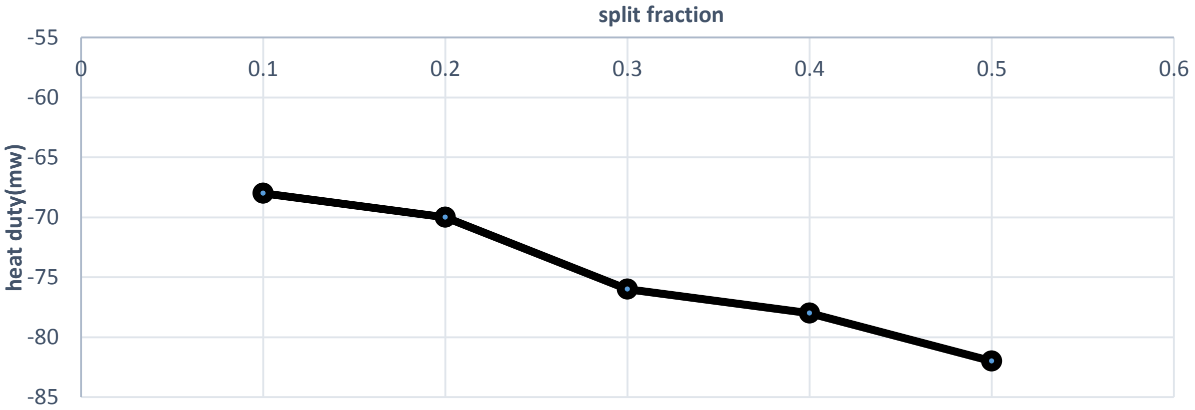

Figure 3.4: Heat Duty of E003 vs Split fraction S2

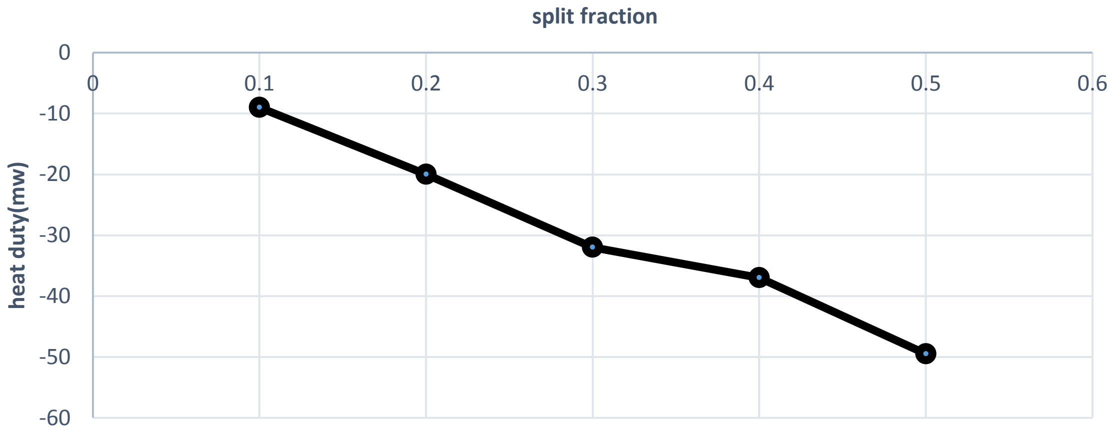

Figure 3.5: Heat Duty of E-005 vs Split fraction of S2

The graphs shown above are plotted between mole fraction of propane recovered and S2 split fraction, brake horse power of C003 (MW) and S2 split fraction, Heat duty of E1, E3 and E5 (MW) with S2 split fraction.

b. Effect of temperature of V1

When we increase the temperature of V1, the temperature of both the streams, stream 6 and 9 are increased. As the reboiler provides heating medium, the exchanger duty of E2 is reduced, thus cooling medium required is also reduced. Stream 3 provides the cooling medium. But for reduced heating medium the flowrate in stream 3 is decreased and so does the flow rate in stream 2, so heat duty of is also reduced.

When the stream 2 flow rate is reduced, the flow rate in stream 7 is increased. This leads to a decrease in temperature of stream 16, thereby lowering the temperature of stream 17. Thus

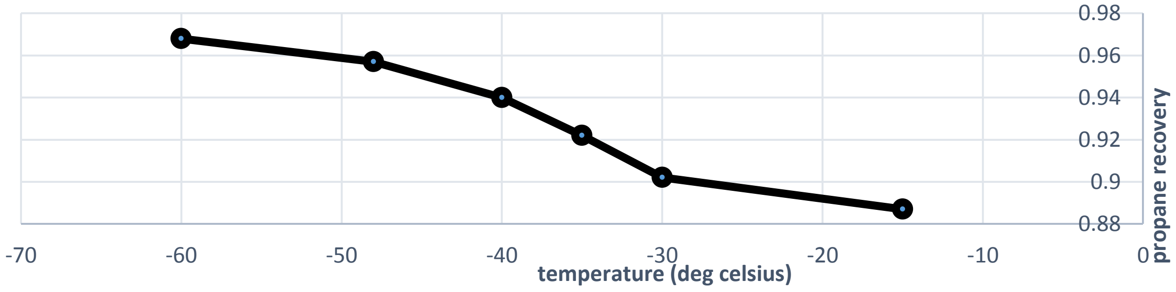

Figure 3.6: C3 Recovery vs V1 temperature

reducing the BHP of C003. Heat to be withdrawn from E3 is also reduced due to the decreased flowrate in stream 2.

If we vary the temperature from −60 °C to −15 °C, there is an observed reduction in the duty of E-001 and E-003 that of 30% and 65% respectively. At higher temperature, the recovery of propane is reduced as it tends to retain in gaseous state at higher temperatures.

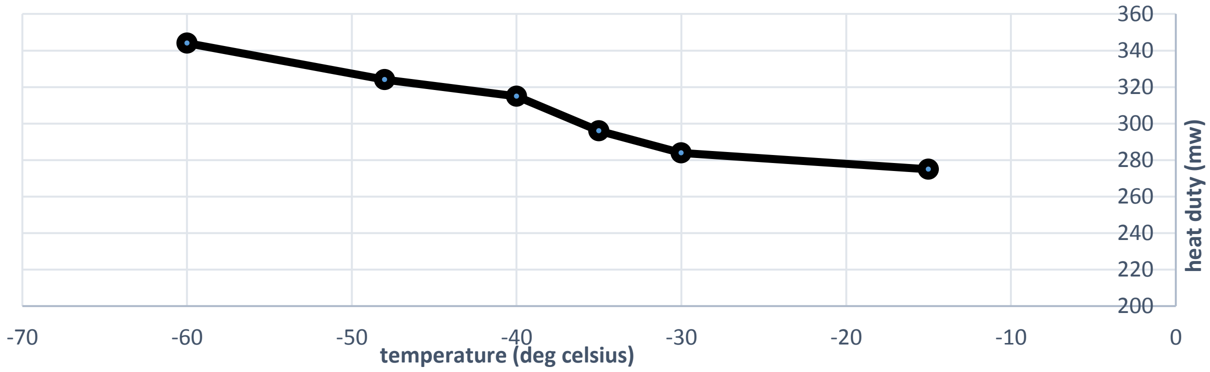

Figure 3.7: BHP C003 vs Temperature of V1

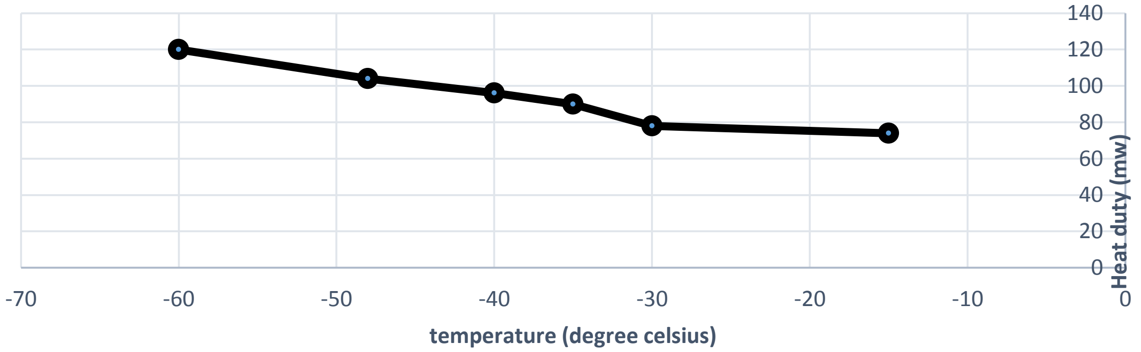

Figure 3.8:Heat duty of E1 vs temperature of V1

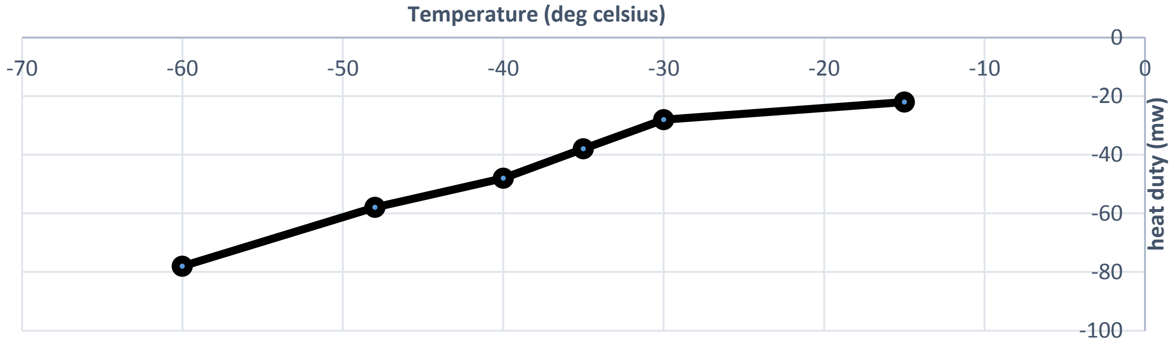

Figure 3.9:Heat Duty of E3 vs temperature of V1

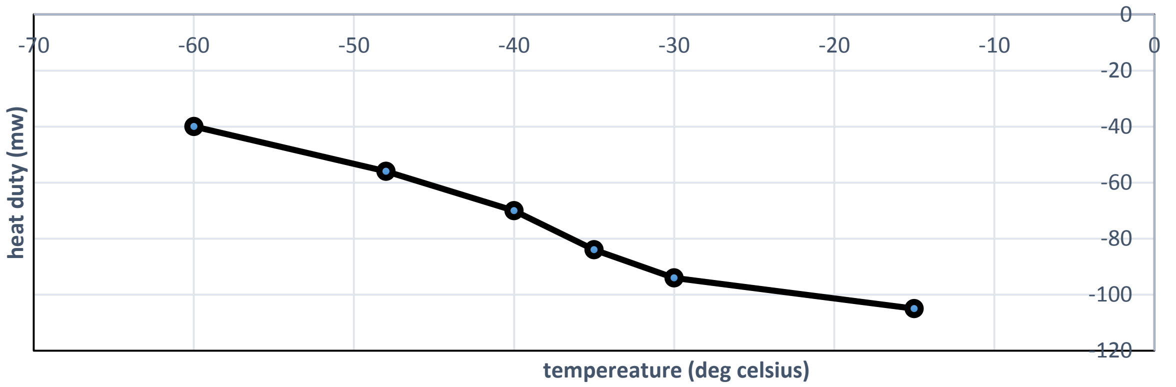

Figure 3.10: Heat Duty of E5 vs temperature of V1

The graphs shown above are plotted between mole fraction of propane recovered (C3 Recovery) and temperature of V1, brake horse power of C003 (MW) and temperature of V1 , Heat duty of E1, E3 and E5 (MW) with temperature of V1.

c. Effect of column pressure

It is observed that the duty of reboiler E2 increases when the pressure of the column is increased. Hence the product stream is slightly affected. Flow in stream 2 is increased when the reboiler duty is increased in order to provide cooling medium in the reboiler. As we increase the flowrate of stream 2, there is a noticeable increase in the heat duty of E1 and E3, thereby causing more expenditure. From the graph we observe that increase in BHP of C003 takes place on decreasing the pre ssure in the column. This increase is around 25%.

The graphs shown above are plotted between mole fraction of propane recovered (C3 Recovery) and distillation column pressure, brake horse power of C003 (MW) and distillation column pressure, Heat duty of E1, E3 and E5 (MW) with distillation column pressure.

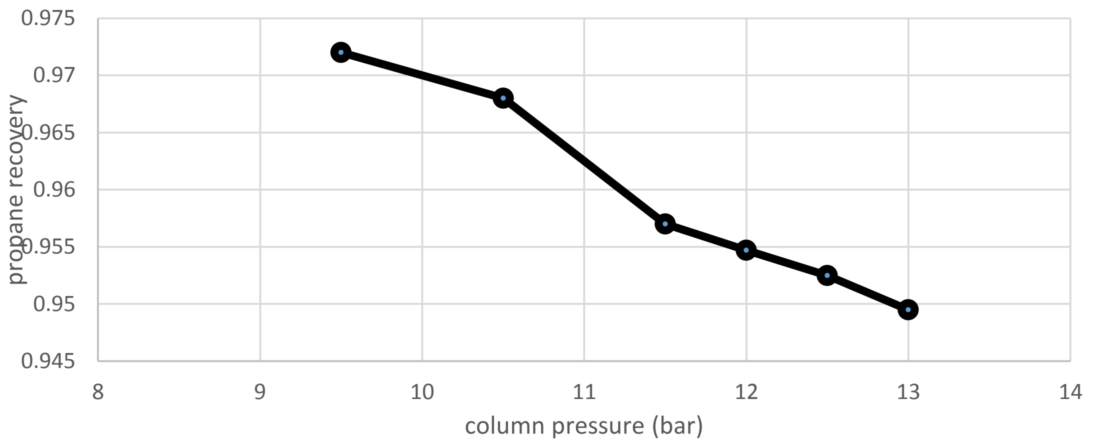

Figure 3.11: Propane Recovery vs Column Pressure

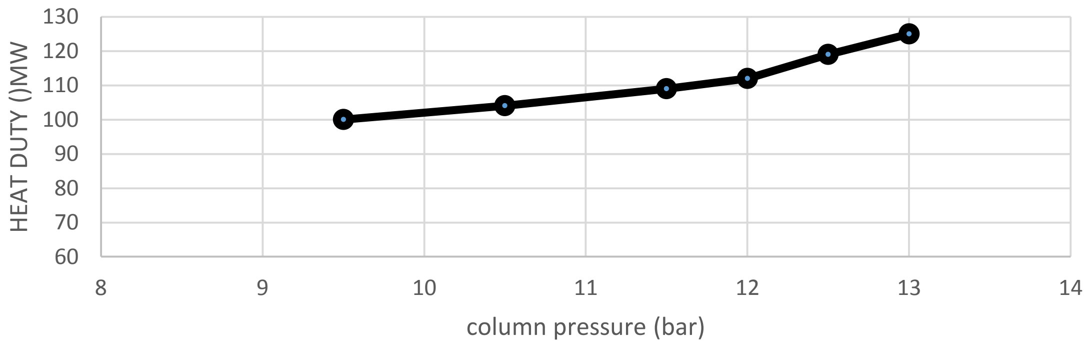

Figure 3.12: Heat Duty of E-001 vs Column Pressure

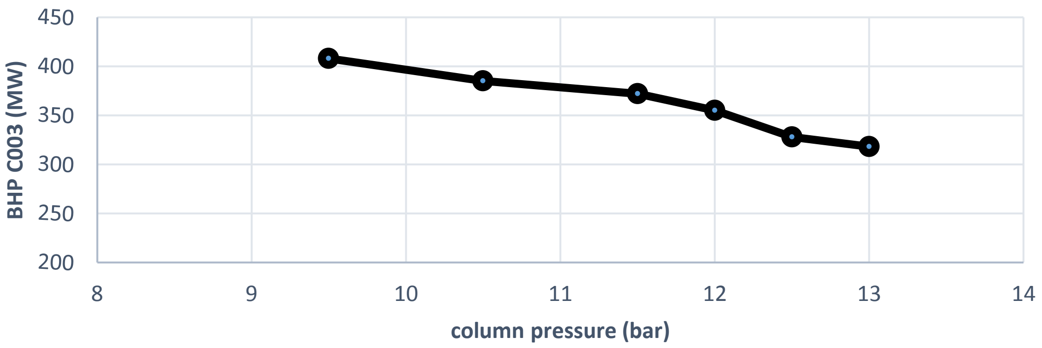

Figure 3.13: BHP C003 vs Column pressure

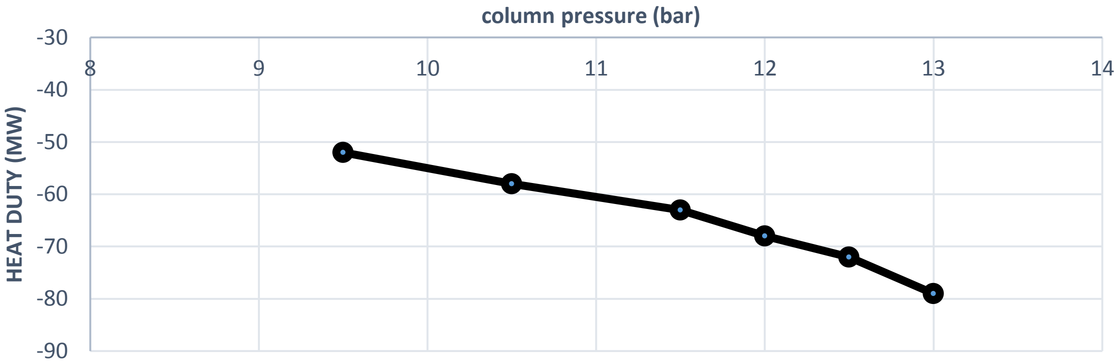

Figure 3.14: Heat Duty of E-003 vs Column Pressure

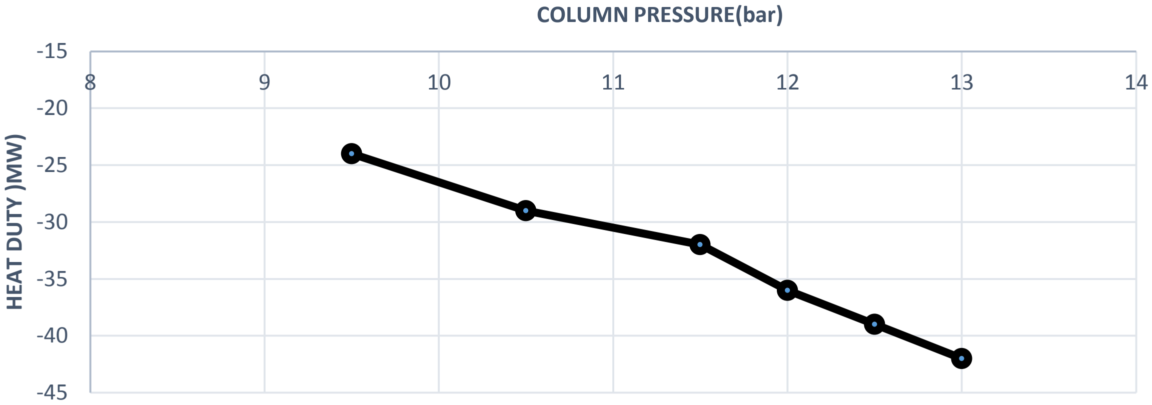

Figure 3.15: Heat Duty of E-005 vs Column Pressure

CHAPTER 4 - CONCLUSIONS

Further, the variation of the parameters that are considered, have been evaluated for one at a time. A brief study on the correlation of these parameters have to be some to finalise the design criteria of the plant. According to the observation and the previous works, we reach to the following design criteria conclusions, the split fraction of the outlet of separator stream should be decreased to the least, choosing this we not only increase the propane recovery but merely reduce the exchanger duties. The variation according to the separator temperature has shown that on lowering the temperature the consumption of energy is reduced to a great extent but this also triggers less propane recovery. So, to fix a certain temperature, observations are needed to be taken at other operating parameters in order to attain maximum recovery relative to other values of operating parameters. Similar is the case for the variation of distillation column pressure on other variables. On lowering the pressure in the columns, sufficient reduction in the energy consumption is noticed but, compression energy increases many folds.

These results are independent of variation in one another, so further study can be carried out by checking their dependence on each other recording observations for maximum recovery of propane, ethane and less energy consumption.

The temperature of the separator, the split fraction of the overhead stream of outlet of the separator and the column pressure significantly affect the refrigeration requirement of exchangers, cost of compression, and the heat required in various heat exchangers. Maximum reduction of the heat required is up to 25% and the refrigeration required is up to 60% with respect to the initial choice of the operating parameters, without affecting significantly C3+ recovery.

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