Process Simulation of a Back-up Condensate Stabilization Unit

by

Ilmi Bin Ilias

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CERTIFICATION OF APPROVAL

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Approved by,

____

(DR Nejat Rahmanian)

UNIVERSITI TEKNOLOGI PETRONAS

TRONOH, PERAK

SEPTEMBER 2012

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

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ILMI BIN ILIAS

ABSTRACT

Hydrocarbon condensate recovered from natural gas may be shipped without further processing but is stabilized often for blending into the crude oil stream and thereby sold as crude oil. In the case of raw condensate, there are no particular specifications for the product other than the process requirements. The process of increasing the amount of intermediates (C3 to C5) and heavy (C+6) components in the condensate is called “condensate stabilization”. The purpose of this work aims to investigate Reid Vapor Pressures (RVP) values in a back-up condensate stabilization unit with a given feed of condensate and obtaining the best actual operating parameter for each of equipment. On the basis specified target for stabilized in this unit, two properties of product should stabilize before storing in storage tanks and export which for RVP of maximum 10 psia for summer season and 12 psia for winter season. Based on the research, it is found some techniques of condensate stabilization which are flash vaporization and fractionation. The separation of the feed is using flash vaporization in back-up unit which does not have any distillation column and just uses heating and flashing processes as we want to have simple process in case of plant shut down. In back-up CSU, salt and sulfur content are not affect the process as there are no any distillation in column and it operate only for shut-down plant as well as not a continuous process. Results show that CSU’s RVP and sulfur content is 7.932 psai and 2408.52 ppm which is the optimum condition for the process.

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ABBREVIATIONS AND NOMENCLATURES

RVP	Reid Vapor Pressure
CSU	Condensate Stabilization Unit
LPG	Liquefied Petroleum Gas
MEG	Ethylene Glycol
PPM	PETRONAS Penapisan Melaka

CHAPTER 1

INTRODUCTION

1.1 Project Background

Nowadays, the consumers of condensate require a stable and sweet product and the gasoline produced by modern plant processes must meet established pipeline and marketing standards. So, stabilization of condensate refers to the stripping of the light ends content (methane - ethane) from the raw liquids and the removal of all acidic constituents to produce a suitable product for the market.

The stabilization operations involved are simple and the principles are similar to the ones used in LPG fractionation systems. In general, condensate stabilization accomplishes several goals, the foremost of which are:

a) To increase the recovery of methane-ethane and LPG products.
b) To lower the vapor pressure of the condensate, therefore making it more suitable for blending and reducing the evaporation losses while the product is in storage or shipment.
c) To sweeten the raw liquids entering the plant by removing the hydrogen sulphide and carbon dioxide contents, in order to meet the required specifications.
d) To maintain the purity and molecular weight of the lean absorption oil, free of certain components like pentanes and heavier hydrocarbons.

1.2 Problem Statement

Natural gas condensate is a low-density mixture of hydrocarbon liquids that are present as gaseous components in the raw natural gas produced from many natural gas fields. It condenses out of the raw gas if the temperature is reduced to below the hydrocarbon dew point temperature of the raw gas in operating pressure.

The raw condensate may include these components; straight-chain alkanes having from 1 to 12 carbon atoms like paraffins, hydrogen sulfide (H_2S), thiols traditionally also called mercaptans (denoted as RSH, where R is an organic group such as methyl, ethyl, etc.) carbon dioxide (CO_2), nitrogen (N_2), cyclohexane and perhaps other naphthenes, aromatics (benzene, toluene, xylenes and ethylbenzene). There are some hydrocarbon condensates are lighter component present in the mixture when a condition has lower pressure will flash off. When this happen, it can cause hazardous conditon for the storage and also transportation of condensate will lose as they evaporate into the atmosphere. Hence, it should be stabilized before transfering to the storage tanks.

In oder to stabilize the hydrocarbon condensate, a condensate stabilization unit with back-up unit as the back-up unit is used only plant failure time. The vapour pressure is called as Reid Vapour Pressure (RVP) and the final product is different according the customers' desired. The reason to build a back-up unit is to operate the condensate stabilization unit although it in failure to ensure the production the condensate for the export.

1.3 Objectives

This project's objective is to simulate a back-up Condensate Stabilization Unit (CSU) that is able to bring down the Reid Vapor Pressure (RVP) of the Summer Rich Condensate of maximum 10 psia for summer season and 12 psia for winter season**.**

Besides that, this project is to find the best operating parameters for each of the equipment in a back-up condensate stabilization unit.

1.4 Scope of Study

This project will focus on researches and findings related to Reid Vapor Pressure for the operating parameter in order to understand the effects on the condensate stabilization unit via HYSYS software.

1.5 Relevancy of project

In terms of the relevancy of this project, it poses a great deal of significance to the oil and gas industry. This process which is condensate stabilization unit is performed primarily in order to reduce the vapor pressure of the condensate liquids so that a vapor phase is not produced upon flashing the liquid to atmospheric storage tanks. In other word, the scope of this process is to separate the very light hydrocarbon gases, methane and ethane in particular, from the heavier hydrocarbon components ($C+_3$). Stabilized liquid, however, generally has a vapor pressure specification, as the product will be injected into a pipeline or transport pressure vessel, which has definite pressure limitations. Condensates may contain a relatively high percentage of intermediate components and can be separated easily from entrained water due to its lower viscosity and greater density difference with water. Thus, some sort of condensate stabilization should be considered for each gas well production facility.

1.6 Feasibility of project

All the objectives stated earlier are achievable and feasible in terms of this project duration and time frame. The whole project is schedule to be completed in 2 semesters

- 1st semester
  - Understanding build up
  - Data collection
  - Familiarization of software
  - Documentation for the whole idea of the project

- 2nd semester
  - Input data to HYSYS software
  - Tuning of operating parameters so that RVP value can be achieved.
  - See the result on RVP value before entering the storage and also final compositions.
  - Analyses the result

CHAPTER 2

LITERATURE REVIEW

Hydrocarbon condensate recovered from the natural gas may be not transferred for further processing but they will be stabilized first in order to bending with crude oil stream and then sold as crude oil. For the case of raw condensate, there are no any specific requirement for the product other than the process specification. So, the process of increasing the amount of intermediates (C_3 to C_5) and heavy ($C+_6$) components in the condensate is called "condensate stabilization" **[1]**. Hence, the hydrocarbon condensate stabilization is required to minimize the hyrocarbon losses from the storage tank **[5]**. This process is needed to be done because a vapour phase will not produce upon flashing to the atmospheric storage tank in order to reduce the vapor pressure of the condensate liquid. Besides that, the purpose of this process is to separate light hydrocarbon gases like methane and ehane from heavier hydrocarbon components such as ethane and others. Heavier components can be used for oil refinery cracking processes which allow the production of light production such as liquefied petroleum gas (LPG) and gasoline **[6]**. Nevertheless, stabilized liquid has vapor pressure specifications as, the product will be transferred into pipelines which have limitation of pressure **[1]**.

In order to measure the vapor pressure of the condensate is by measuring the Reid Vapor Pressure (RVP). Reid vapor pressure (RVP) is a way to measure how quickly fuels evaporate; it's often used in determining gasoline and other petroleum product blends **[2]**. It means that higher RVP of a fuel, the more it quickly evaporates indicating the loss of the product. RVP represents the fuel's evaporation at 100 degrees Fahrenheit (37.8 degrees Celsius), and is measured in pounds per square inch, or PSIs **[2]**. Hence, the property that RVP measures often is referred to as the gasoline's volatility. RVP can be estimated without performing the actual test by using algorithm **[7]**.

2.1 Natural Gas-Processing

Figure 1 shows the flow of condensate to be stabilized before transferring to the storage tank which is starting from the natural gas well.

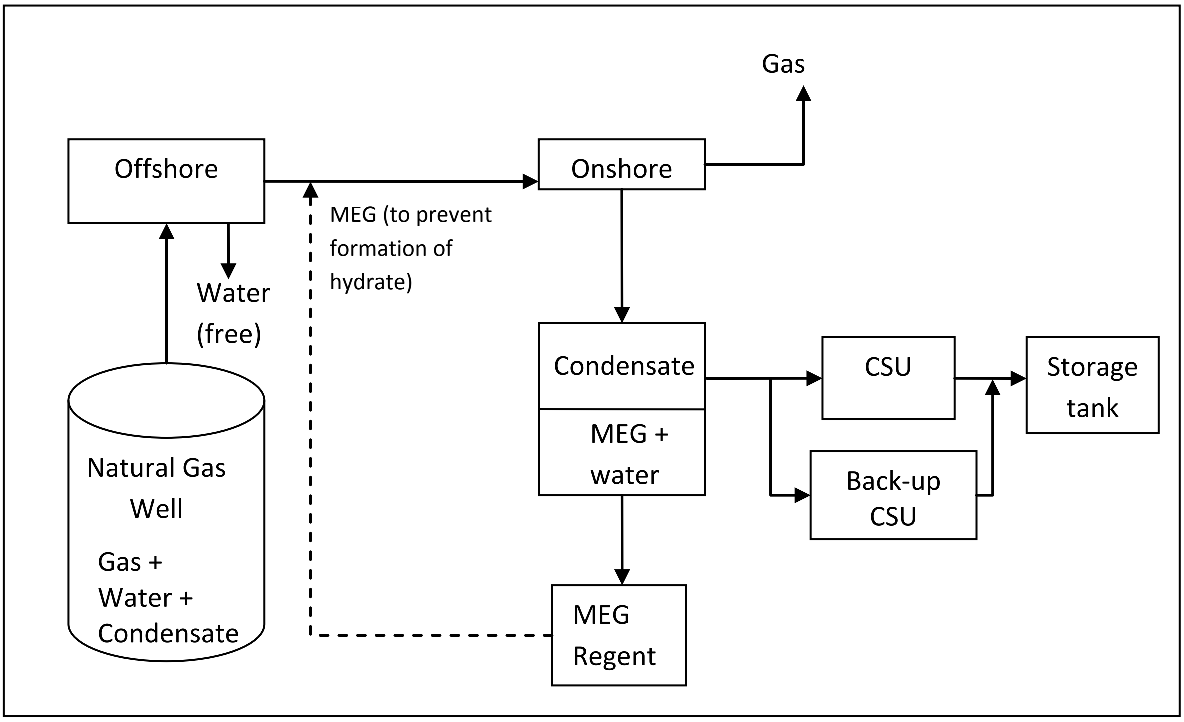


Figure 1: Flow Diagram of Condensate Stabilization
(Dr. Nejat, 2012)

Firstly, at natural gas well, a mixture of natural gas which consists of gas, water and condensate will be extracted before it is being transferred to the offshore plant (oil rig). Then, some water will be removed out from the mixture and transported to the onshore plant. The transportation of the treated gas will be done through a pipeline about 120km from offshore plant to onshore plant. As the result, the gas mixture will dehydrate and form a blockage which the flow of gas will not go smoothly. Hence, monoethylene glycol (MEG) is channeled to the pipeline in order to prevent the formation of gas hydration.

Once gas mixture reaching in onshore plant, it will be separated into two stream; gas stream and liquid stream. The gas stream will be transferred to gas plant and the liquid stream that consists of condensate, MEG and water is further separated which form a condensate stream and mixture of MEG and water stream. The mixture MEG and water will be treated in MEG regeneration unit which MEG will be recycled to the pipeline. Then condensate stream will send to the condensate stabilization unit (CSU) with a back-up unit to run the plant during failure. After treated in CSU, the condensate will be stored in the storage tanks.

2.2 Condensate Stabilization

Stabilization of condensate streams can be accomplished through either flash vaporization or fractionation.

2.2.1 Flash Vaporization

Stabilization by flash vaporization is a simple operation employing only two or three flash tanks **[1]**. This process is similar to stage separation utilizing the equilibrium principles between vapor and condensate phases. Equilibrium vaporization occurs when the vapor and condensate phases are in equilibrium at the temperature and pressure of separation **[1]**.

Figure 2 shows the typical of flash vaporization process for the condensate stabilization. Based on the Figure 2, the main feed which is condensate coming from the inlet separator is passing through a heat exchanger entering the high-pressure flash tank where the pressure is maintained at 600 psai. A pressure drop which costly 300 psai help the flashing of large amounts of lighter ends which they will be discharge to sour vapor stream after recompression. The discharged ones can be sent to the further units or recycled into the reservoir. After that, the bottom liquid from the high-pressure tank will enter the middle pressure flash tank where the additional mehtane and ethane will be released. Then, the bottom the product will enter again to the low-pressure tank and they will enter the condensate stripper for the purification before sending to the storage tank.

To ensure efficient separation, condensate is degassed in the stripper vessel at the lowest possible pressure prior to storage **[1]**. This reduces excess flashing of condensate in the storage tank and reduces the inert gas blanket pressure required in it. Multistage flashing is based on the principle of progressively lowing the pressure of condensate during each stage **[5]**. This will enhance for the flashing of lighter components from the condensate.

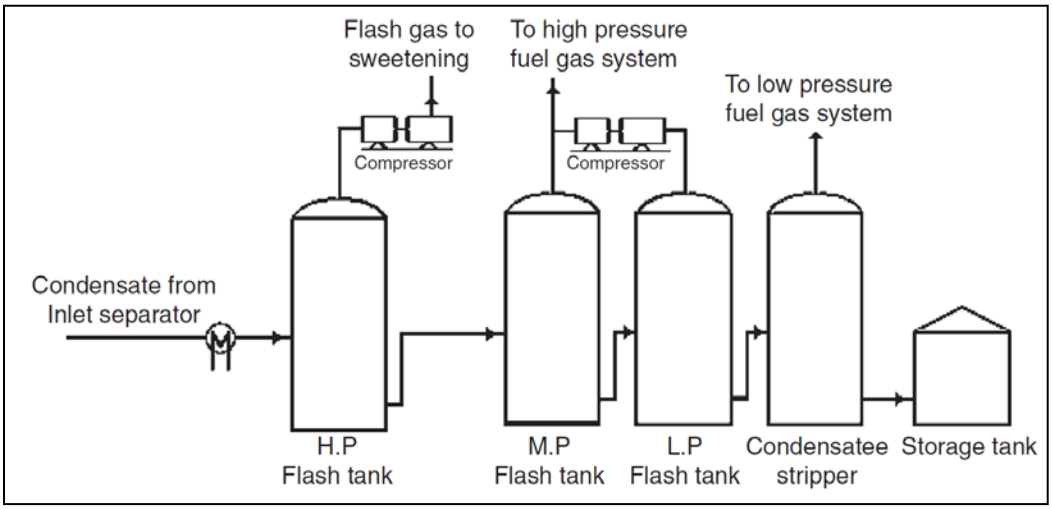


Figure 2: Flash Vaporization Method
(Mokhatab, S., Poe, W.A. and Speight, J.G., 2006)

2.2.2 Fractionation

During the condensate stabilization unit, the light compoent like methane, ethane, propane and butane are removed and recovered. Hence, the desired product form the bottom column consits of pentane, heavier components and also small amount of butane. Actually, the porcess make a cut between the lightest liquid component (pentane) and the heaviest gas (butane) **[1].** The final product is liquid free from the all gaesous components and can be stored in the storage tank safely.

Figure 3 shows a typical fractionation of condensate stabilization process. Firstly, the liquid hydrocarbon (condensate) is sent into the system from the inlet separator and heated in the stabilizer feed/bottoms exchanger before entering the stabilizer feed drum.

In the condensate stabilizer, it reduces the vapor pressure of the condensate by removing the lighter components. Typically, fractionation method required the process in a rebouled absorber. However, if a better separation is required, typically the column is changed from a top feed reboiled absorber to a refluxed distillation tower **[1]**.

At the bottom of the stabilizer, some of the liquid is circulated through a reboiler to increase the tower. The heavy ends can get stripped out of the gas at each try as the gas goes up from tray to tray. So, the gas is in rich of light ends and leaner in the heavy ends. Overhead gas from the column will then send to the low-pressure fuel gas system through a back-pressure control valve to maintain the pressure of the stabilizer because it seldom meet the requirement the market demand. For the bottom product, they will undergo a series of the stage flahes at ever-increasing temperatures to remove off the remainding light components. They must be cooled to sufficiently at lower temperature to prevent the flashing to atmosphere in the condensate storage tank.

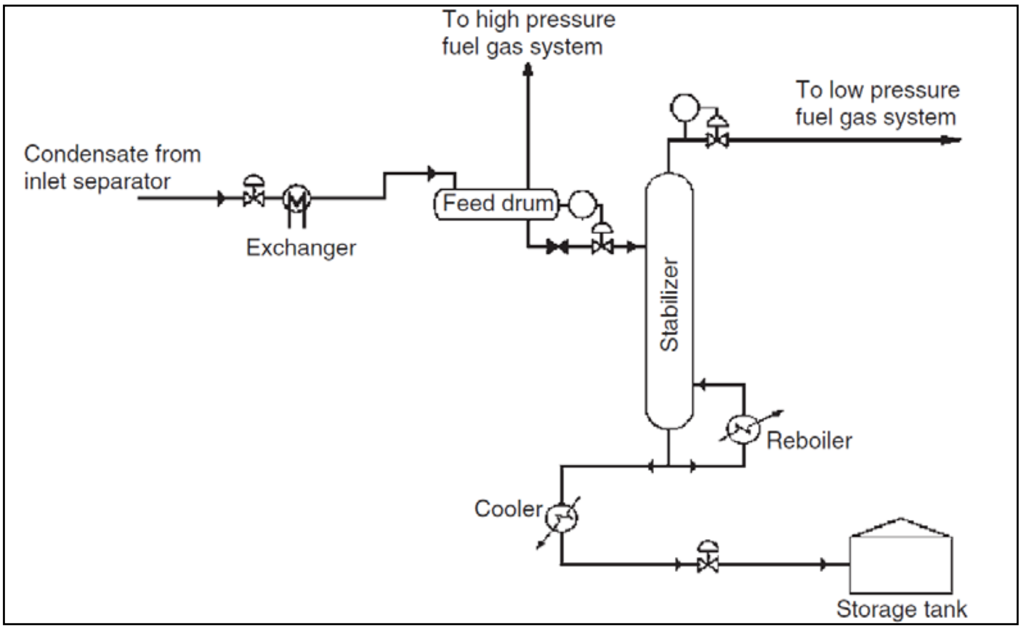


Figure 3: Fractionation Method
(Mokhatab, S., Poe, W.A. and Speight, J.G., 2006)

Selection of the stabilization technique shall be governed by parameters like reservoir conditions, fluid compositions and specification of export condensate vapor pressure **[5]**. For the back-up unit, it is found that it just use only simple heating and cooling process as we want to reduce the cost as well as the it is not in continuous process. Hence, back-up unit prefers to use flash vaporization method to run its operation. This method just uses only some pressure to stabilize the condensate before sending to the storage tank.

2.3 3-Phase Separator

For the separation of condensate from the mixture, 3-phase separator is used as there is a large amount of gas to be separated from the liquid **[8]**. This separator is a pressure vessel that is usually used to remove and separate the water from the mixture of crude oil. However, in the oil refinery plant, the 3-phase separator is designed to separate the gas that flashes from the liquid and also separate oil and water because flow normally enters these vessels comes from onshore plant at higher pressure. Hence, proper selection of the separator type is important.

For the 3-phase separator, a horizontal separator is more effective than a vertical separator **[9]**. This is because, in a horizontal separator, the area of the vapor space is reduced and the possibility of liquid entrainment increase as the liquid level is increase. So, the separation will be effective because it can separate water and unwanted gas at large portion. On other hand, the liquid entrainment should not be concerned at high liquid level and the vapor-flow area remains constant in the vertical separator. The advantages and disadvantages of horizontal and vertical 3-phase separator are show below:

	3-phase Separator	
	Horizontal	Vertical
Advantages	1. It has high separation efficiency in comparison with a vertical separator 2. It is the only choice for a single inlet and two vapor outlets 3. It is easy to design 4. It is more suitable for handling large liquid volume	1. The liquid surface area does not change with the liquid height, hence liquid entrainment is reasonably constant. 2. It requires smaller footprint area. 3. It is easier to install level instruments, and others 4. It is usually more efficient for vapor-liquid ratios.
Disadvantages	1. It is required a larger footprint area. 2. The liquid entrainment rate increases with the increase in liquid level.	1. It is not suitable for 3 phase separation. 2. It is less suitable for vapor-liquid ratio.

Table 1: Advantage and Disadvantage of Type 3-Phase Separator

Based on the comparison between type of 3 phase separator which are horizontal and vertical, for the back-up condensate stabilization unit, horizontal 3-phase separator will be used as it will separate gas, oil and water at higher efficiency separator and very suitable used for handling large liquid volume. These ensure that the product from this unit will have high quality and meet the customer demand.

2.4 Impact of Salt and Water on Back-up CSU

Apart from crude oil in the mixture of condensate, there are also presence of salty, acidic water and solid particulate which cause various problems in the stabilization plant. Separation of water phase from the condensate can be problematic as many fields or plants from various regions experience it. Although the condensate viscosity is very low and the difference of density with water is high, other impurities tent to create stable condensate/water emulsion that are difficult to separate efficiently **[10]**.

There are many consequences on the impact of impurities in the condensate stabilization plant. Many plants in worldwide have reported that several following consequences may arise due to water carry over that contains dissolved salt like:

1. Plant upsets and stability of the plant is reduced.
2. Quality issues of the final products for example gasoline and LPG.
3. Excessive corrosion and deposits inside the stabilizer and re-boiler.
4. Power consumption is increased due to the ingression of excessive levels of water and loss of heat transfer caused by the contaminants.
5. There will be frequent shutdown of the stabilization train for the cleaning purposes, causing a drop in production and hence loss of revenue if the flow rate cannot be compensated by the other stabilization trains.
6. The corrosion products will be created in the export condensate storage tank and in the export pipeline also referred to as ‘black powder’.

Figure 4: Salt deposits in the de-ethanizer reboiler top tube sheet before cleaning (Crew Energy Inc., 2011)

Figure 5: Salt deposits in the de-ethanizer reboiler top tube sheet after cleaning (Crew Energy Inc., 2011)

Figure 6: Deposits collected from reboiler tubes at Middle East Plant (Crew Energy Inc., 2011)

Normally, contaminants found in the unstabilized condensate include free, emulsified and dissolved water, salts acidic components (Sodium Chloride, Magnesium Chloride, etc); corrosion inhibitors, hydrate inhibitor (Mono Ethylene Glycol (MEG), methanol, and Kinetic Hydrate Inhibitors), and solid particles (corrosion products , sand) and solid-like particles (waxes,gels) **[10]**. Hence, water, salts and particle should be removed from the stabilizer operation and also the export pipeline.

These contaminants that mostly affect the water separation from the condensate usually are the corrosion inhibitor, MEG or methanol as they act as surfactants lowering the Interfacial Tension (IFT) and creating stable emulsions that cause water carryover. Many results show that water carryover issue is the common problem from various types of the separators.

Water in condensate downstream of inlet separator is typically present in concentrations varying from few hundreds ppmw (parts per million by weight) up to 5% **[10**]. The salinity of the water contamination is measured by the formation water and varies from hundred water ppm to few hundred thousand water ppm. Quality specifications of the condensate prior to the stabilizer an export pipeline is free water concentration ranging from less than 10 ppmv (parts per million by volume) to less than 100 ppmv **[10]**.

In order to separate impurities from the condensate, we need to have desalter/dehydrators in the plant. Mostly, desalter is electrostatic precipitators and utilizes new technologies which are three grid-grid electrode system and horizontal emulsion distribution for better separation performance **[11]**. This equipment should be installed in the stabilization plant. However, in back-up condensate stabilization unit, desalter is not included it is an expensive equipment and also only used after the plant shutdown. This is save a lot of money as well as it can get more profit from selling the product. Besides that, we can see the effect of the impurities on the stabilizer which the distillation column and the impurities affect the reboiler performance. As the result, we cannot get desired product and the desalter should be installed in the main condensate stabilization unit.

Figure 7 : Desalter
(Cameron Inc., 2010)

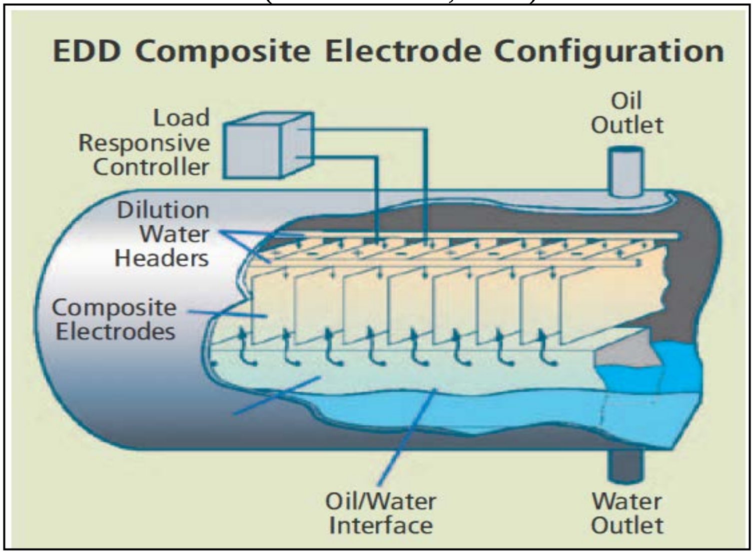


Figure 8: New Technology of Desalter
(Cameron Inc., 2010)

2.5 Impact of Sulfur Concentration on Final Product

Elemental sulfur is a powerful oxidant. It means that the strong oxidizing property In the oil and gas industry, sulfur is recognized as aggressive corrosion accelerators, particularly for pitting and other forms of localized corrosion **[12]**. Normally, sulfur is formed in sour oil and gas systems from some of the following mechanisms; differential solubility of sulfur in high pressure sous gas, destabilization of hydrogen polysulfide presents in sour gas an others. If there is more than 2.5% sulfur present in crude, they are called sour crude **[13]**.

According to McConomy curve, measure corrosion rates of carbon, low allow, and stainless steels are significantly high where significant concentrations of Mercaptans which containing sulfur element are present in crude oils and hydrocarbon condensate **[14]**. It suggested that sulfur element that containing hydrocarbon condensate cause higher corrosion rate than sulfur species in general. Besides that, there more species of Mercaptans in the condensate, the higher corrosion rate will occur. Thus, we need to concern about presence of Mercaptans in the final product in the back-up condensate stabilization unit.

In addition to that, Mercaptans will also give smell on the condensate. This will affect the quality of condensate before selling to the customer. Nevertheless, Mercaptans are added to odorless natural gas for safety reason which in normal operations, gas companies add it to deliver to the city gas stations and commercial usage **[15]**. This is because Mercaptans will prevent the potential underground water contamination which natural gas will be not in good condition.

2.6 Flare System

Some of the plant will have gas waste to dispose. Among of the techniques that to dispose gas waste is by burning in the flare system. This is because flare are used in the hydrocarbon and petrochemical industries as a way to achieve safe and reliable vapor release during a plant upset or emergency situation **[16]**. The waste will send to flare stack, where the gaseous such as propane and propene are flared at a safe height above the process area. A schematic diagram of a flare system is shown in Figure 9 while the detail drawing of a flare tip is shown in Figure 10.

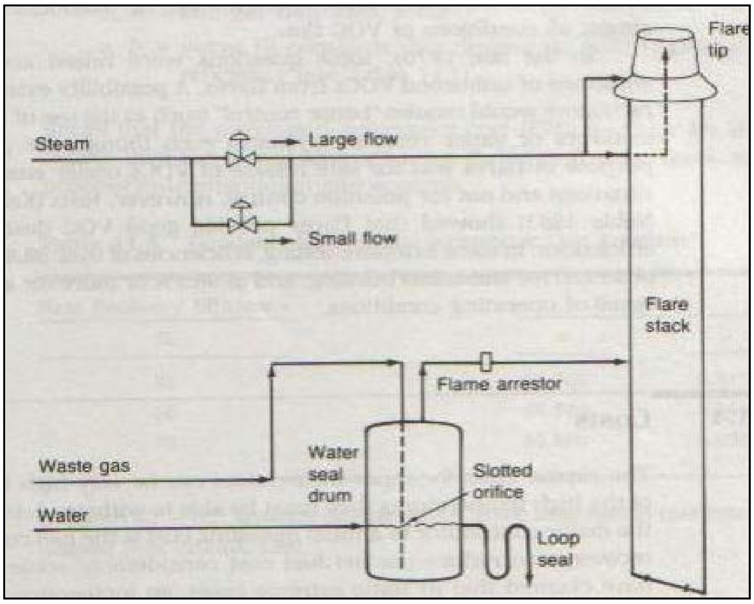


Figure 9: Process Flow of Flaring System
(Fluor Daniel, 2000)

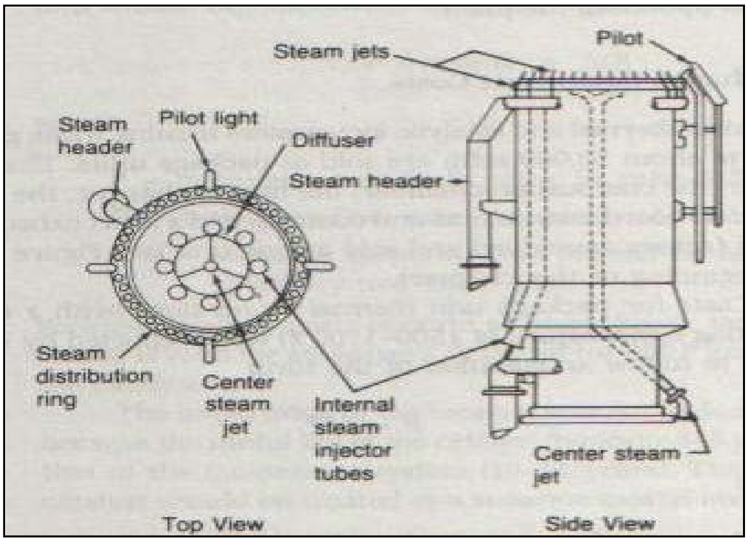


Figure 10: Detailed Drawing of Flare Tip
(Fluor Daniel, 2000)

Flare tips use steam to create a turbulent mixing between air and the stack gas at the top. It also provides some cooling of the flare tip and stack. The flammable gas is ignited at the top by a continuous pilot. The main control that needs to be maintaining along the flaring process is the control of proper steam flow. This is because with proper steam flow, smokeless operation can be maintained at all conditions of gas flow, which provide an almost complete combustion of gaseous.

The flaring process may results in some smoke emissions to the atmosphere. In order to ensure that the little amount of smoke emission is complying with Malaysian Environmental Quality (Clean Air) Regulations 1978, a filter could be installed on top of the stack gas tip before the gaseous is released to the air **[17]**. Gas quality monitoring system need to be installed to ensure the quality of gas that being released into the environment is within the acceptable range of Clean Air Regulation 1978. For the Clean Air Regulation 1978 standard refers to the Appendix B.

2.6 Malaysian Hydrocarbon Condensate

Mostly, Malaysian condensates mainly come from Bintulu Condensate and Terangganu Condensate which is local condensates **[18]**. This is because Malaysia has many gas wells which can produce a lot of quality condensate for the local market demand. The composition of the condensate that usually used in the plant mostly in Malaysia as in Table 2:

Product	**Vol%**
Light Petroleum Gas (LPG) + iC5	9.97
Light Naphthalene (LN)	27.59
Chemical Feed Naphthalene (CFN)	49.74
Straight Run Kerosene (SRK)	9.35
Diesel	3.34

Table 2: Product Yield of Bintulu Condensate Based on Total Boiling Point Cut points (Fatin Nadiah, 2012)

Based on the Table 2, we can see that CFN which has the highest value of volume in the condensate. This shows that CFN has the highest demand in the market.

Besides that, in Malaysia, PETRONAS Penapisan Melaka (PPM) has the specified requirement of the condensate as in the Table 3.

Properties	**Limit**			**Source**
Whole Condensate	**Units**			
Specific Gravity (Dry)	N/A	0.86	max	
Basic Sedimentation and Water (BSW)	vol%	0	max	Design Feed
Total Sulfur	Wt%	0.05	max	Diesel sulfur limit
Salt Content	PTB	0	max	Design Feed
Total Acid No	MgKOH/g	0.5	max	Metallurgy Limit
Pour Point	◦C	45	max	Design Feed
Mercury	ppb	25	max	Design Mercury Removal Unit
Viscosity	cP	3.02	max	Pump Design
Overhead distillate	m^3/hr	90	max	Pump Design

Table 3: Condensate Specification of PPM
(Fatin Nadiah, 2012)

Table 3 shows the condensate specification of PPM required operating in their plant which we can see that the most important part is the total sulfur in the condensate which only 0.05% maximum in the condensate. This is because it can affect the whole process in the plant where the product will not meet the requirement of the customer. Thus, the condensate should be treated in term of sulfur content to be low as possible. For salt content in the condensate, PPM required is 0 PTB which is nearly the zero. In order to meet this requirement, the CSU should consider this factor and eliminate the salt content as high as possible.

CHAPTER 3

METHODOLOGY/PROJECT WORK

This project is develop in two main phase which are construction of plant simulation and analysis of the effects of process parameters. This section covers on the detail of the two main phases, especially on the project structure to give more clear description and understanding about the project itself. Methodology is covered later in this chapter after the project work writing.

3.1 Project work

3.1.1 Overview

In analyzing CSU system performance, plant simulation is modeled first by using HYSYS simulation. It is essential to have a model that reliable in representing CSU system as some of the data is unavailable from the plant and only available from the estimation from HYSYS model. To achieve this objective, the plant simulation is using the actual operating value, gained from data available in real plant. Plant simulation that is using plant actual operating value will able to represent the real simulation of current plant operations. To increase the reliability and confidence in the plant simulation, the estimated data from the simulation will be compared with the actual data plant.

Most of the CSU in the world are using 3-phase separator to separate the water content, oil and gas in the condensate. It is essential to meet the customer’s demand condensate specification as the composition should be same to produce the quality product. The CSU system performance is analyzed in several essential areas such as steam temperature, steam pressure, feed flow rate, feed temperature and feed pressure. By performing such analysis, operator is able to know more and can strategize based on current operating CSU system.

This project is conducted based on three separate components. First is the construction of CSU simulation model in HYSYS. It is constructed based on available design cases that cater most extreme condition such as maximum steam pressure and temperature. Secondly, the results are mapping for data collections. Last component are the RVP and sulfur content analysis based from the available parameter and estimation from simulation model.

3.1.2 Plant Simulation

The model is constructed based on reference CSU plant operation. In the CSU, it consists of one main stream which coming from onshore plant. The condensate is then, passed through three 3-phase separators before sending to the storage tank which to achieve low RVP and also sulfur content. The removed gas will be sent to others unit like gas processing plant. For the heavy liquid, like MEG will be sent to others unit like MEG Regeneration plant. Later in the result, the process description will be discussed in the result and discussion description. Plant simulation model is constructed for the whole CSU plant. However, for initial model construction is based on design basis.

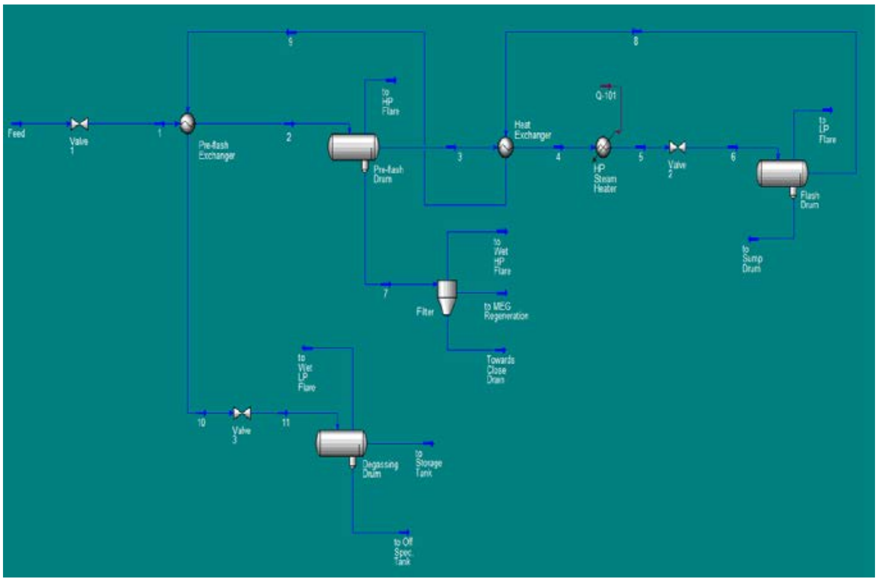


Figure 11: HYSYS Simulation Model

3.1.3 Mapping the Result

The earlier constructed model is based on design basis which cater design cases such as maximum and minimum steam temperature. In operating plant, rarely plant operations are up to maximum condition. Instead of using design variable, the simulation is integrated with process parameter. Using process parameter, the simulation simulates current plant operations. Estimation from simulation model can be compared with the actual data plant of condensate composition to show the reliability of simulation model.

In mapping the result, available process parameters are needed to be identified. With the process parameter input, estimated RVP and sulfur content are generated. With lots of process parameter involved, organized results mapping is a practical use. As in Figure 3.2, process parameters data will be entered in the HYSYS simulation. After finishing input the data, the result will be stored at different spreadsheets which are sulfur content and RVP. It is organized and easily to distinguish between two different database.

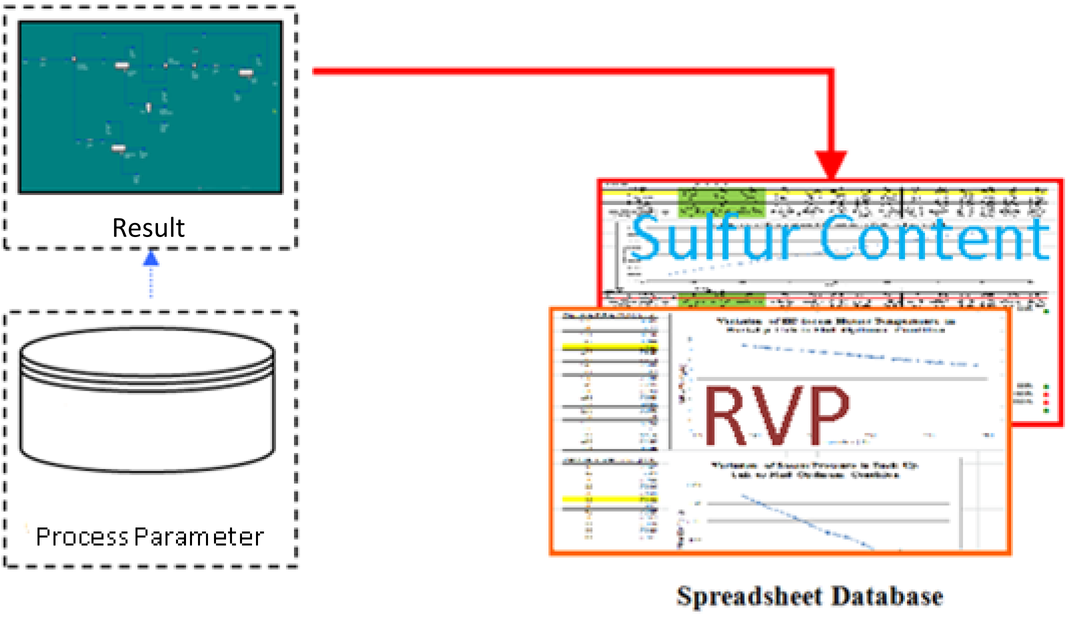


Figure 12: Process Parameter Input and Result Mapping

3.2 Methodology

3.2.1 Project Methodology

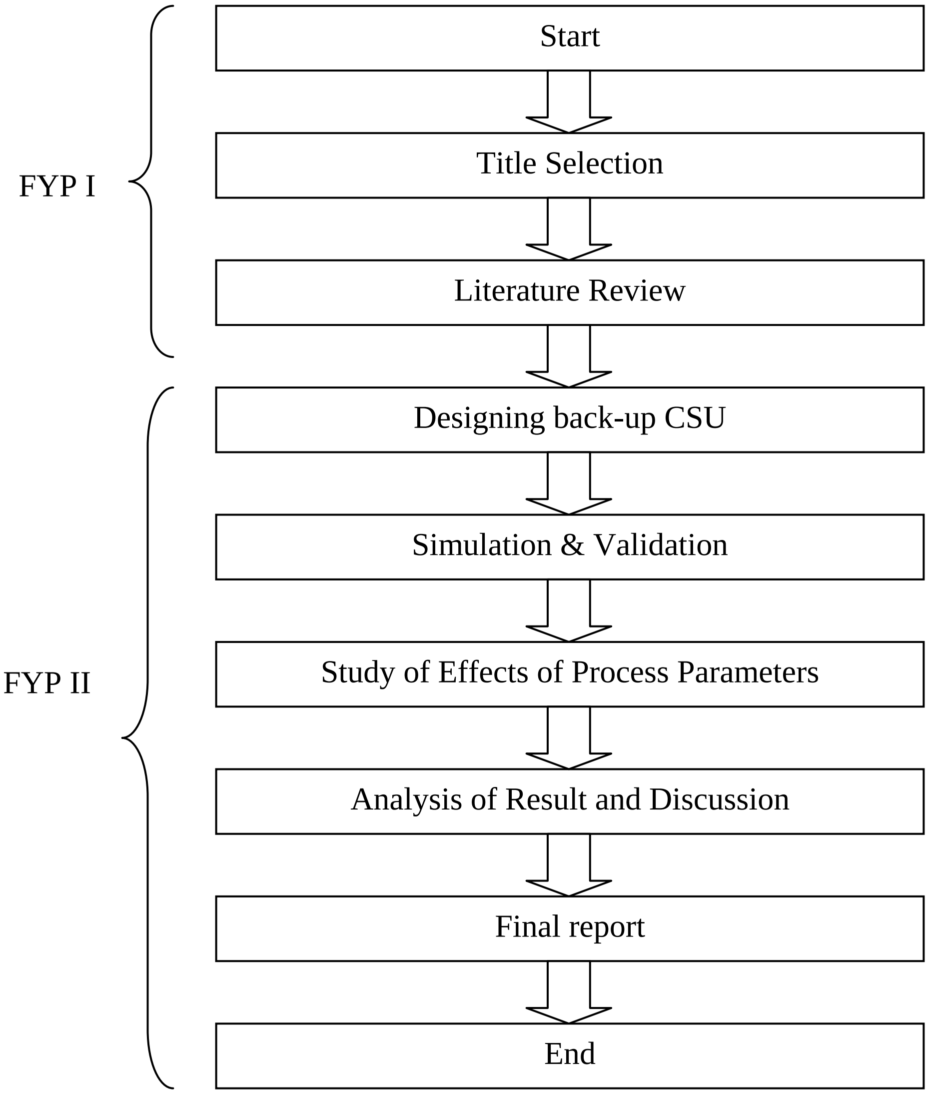


Figure 13: Project Activities Flow

The project is a design base project. Specifically, it is a design of a back-up Condensate Stabilization Unit. First and for most, the project will begin with the research on several issues which had been mention in the research methodology below. With the collective information, the project will proceed with the literature review on the condensate stabilization unit. Besides, the author will discuss a basic knowledge of typical method of condensate stabilization unit which are flash vaporization and fractionation.

After completing the literature review, the further studies will move on to design the back-up condensate stabilization unit. Besides, the author needs to identify the parameters that involved in the condensate stabilization unit such as RVP and temperature. Based on the literature review, it is found that it just uses simple heating and cooling process which does not need a distillation as to reduce the capital cost. Then, the simulation of back-up condensate stabilization unit will be done by using HYSYS. A study of effects of process parameter which are steam temperature, steam pressure, feed flow rate, feed temperature and feed pressure. After completing the simulation, the result and discussion will be done to know the effect of summer and winter season in the plant.

Lastly, all the studies and discussion will be compiled in the final report. Apart from that, the new design of plant elements features can be further explain and justifies. The operational and safety requirements can also be developed from the study.

3.2.2 Research Methodology

Research is a method taken in order to gain information regarding the major scope of the project. The sources of the research cover the handbook of condensate stabilization unit, e-journal, e-thesis and several trusted link.

The steps of research:

1. Gain information of the condensate stabilization unit and comparison of the method is been used.
2. List down the design and parameters of condensate stabilization unit.
3. Finalize the design and parameters that will be used in the simulation.

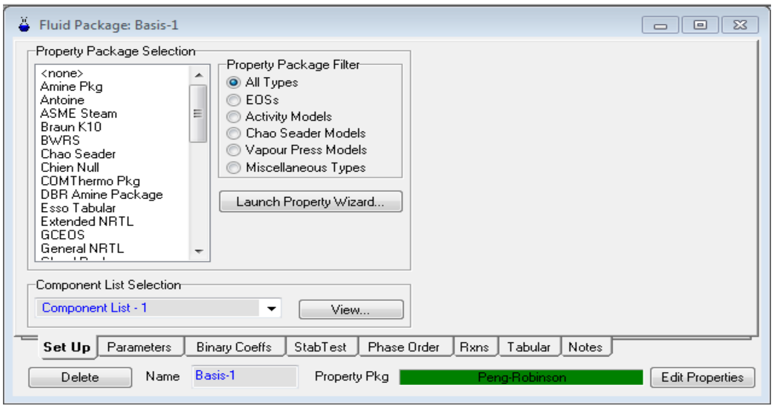
3.2.3 Project Simulation

Aspen HYSYS is process simulation software that enables plant operations simulation in mostly on process area. The software a powerful simulation tools especially in material and heat balance, flow estimation and unit operations. Besides that, HYSYS is also a process modeling took which can be used for conceptual design, optimization and performance for oil and gas production and others.

3.2.4 Process Design

The simulation of the back-up Condensate Stabilization Unit is conducted by using Aspen HYSYS software. The main equipment that are used are, 3-phase separator, heat exchanger, simple solid filter, and heater.

A gas stream composition and conditions are first added for a case study and suitable HYSYS fluid package is chosen. In this case, Peng-Robinson Package is used based on the polarity, electrolyte and pressure of the components.



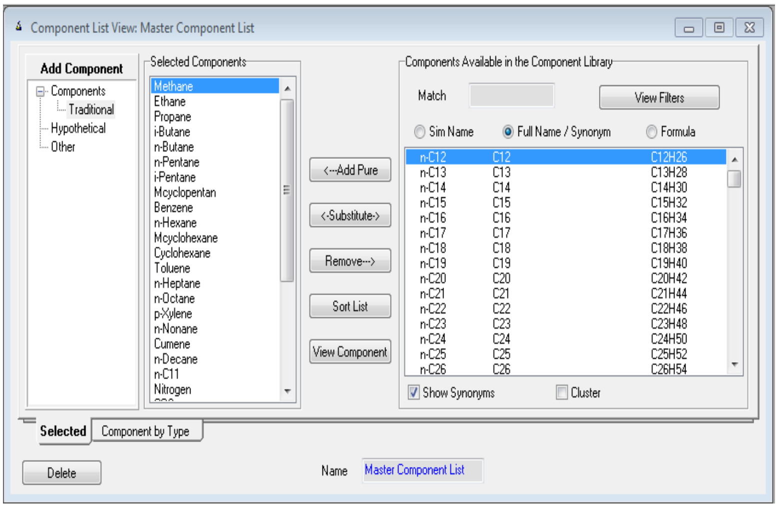
Component List Selection: Component List - 1

Name: Basis-1

Property Pkg: Peng-Robinson

Figure 14: HYSYS fluid package window

The component of the fluid is selected from the component lists provided in HYSYS simulator. Then, the simulation environment is entered and proceeds with the construction of other required equipment.



Selected Components:
Methane
Ethane
Propane
i-Butane
n-Butane
n-Pentane
i-Pentane
Mcyclopentan
Benzene
n-Hexane
Mcyclohexane
Cyclohexane
Toluene
n-Heptane
n-Octane
p-Xylene
n-Nonane
Cumene
n-Decane
n-C11
Nitrogen

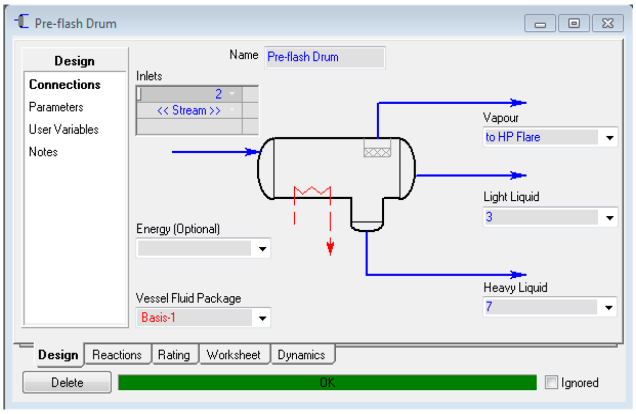
Components Available in the Component Library:

n-C12	C12	C12H26
n-C13	C13	C13H28
n-C14	C14	C14H30
n-C15	C15	C15H32
n-C16	C16	C16H34
n-C17	C17	C17H36
n-C18	C18	C18H38
n-C19	C19	C19H40
n-C20	C20	C20H42
n-C21	C21	C21H44
n-C22	C22	C22H46
n-C23	C23	C23H48
n-C24	C24	C24H50
n-C25	C25	C25H52
n-C26	C26	C26H54

Name: Master Component List

Figure 15: Components Selection Window

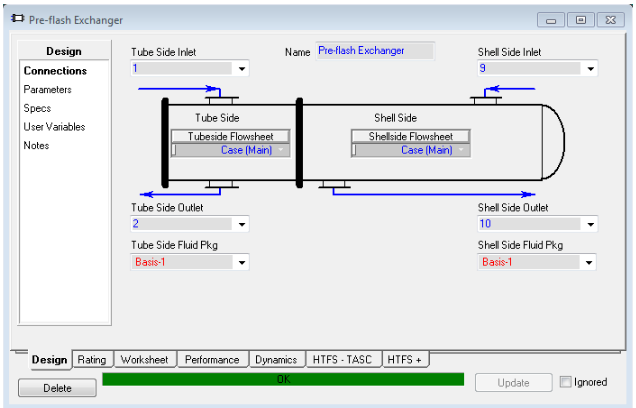
For the 3-phase separator, it is needed to define 1 input stream and 3 output stream. Then, the author need to enter the temperature and pressure required. Data input for Pre-flash drum, the temperature and pressure are 39◦C and 1151 kPa respectively, for flash drum is 128.4◦C and 401.3 kPa respectively and for degassing drum is 40.99◦C and 151.3 kPa respectively.



Name: Pre-flash Drum
Inlets: 2
Vapour: to HP Flare
Light Liquid: 3
Heavy Liquid: 7
Vessel Fluid Package: Basis-1

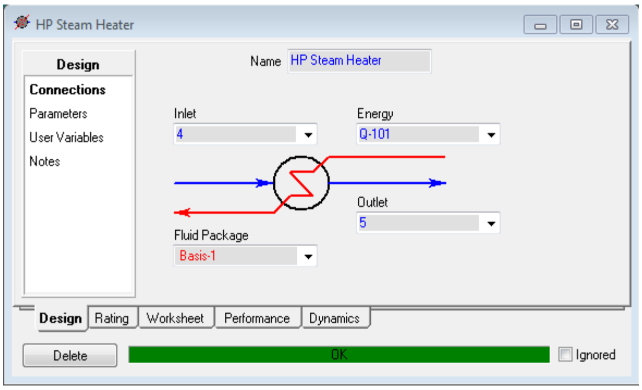
Figure 16: 3-Phase Separator Data Input Window

For heat exchanger, it is needed to define 2 input streams and 2 output streams whereas the heater only needed to define 1 input stream and 1 output stream. Data input for Pre-flash exchanger is 2 input streams' temperature and pressure are 17.27◦C, 39◦C, 1151 kPa respectively and 2 output streams' temperature and pressure are 79.10◦C, 40.93◦C, 331.3 kPa, and 261.3 kPa, for Heat Exchanger are input data's are 39◦C, 80◦C, 1151 kPa and output data's are 128.4◦C, 79.10◦C, and 1655 kPa respectively. For heater input data's 80◦C, 143◦C, and 1151 kPa respectively.



Tube Side Inlet: 1
Name: Pre-flash Exchanger
Shell Side Inlet: 9
Tubeside Flowsheet: Case (Main)
Shellside Flowsheet: Case (Main)
Tube Side Outlet: 2
Shell Side Outlet: 10
Tube Side Fluid Pkg: Basis-1
Shell Side Fluid Pkg: Basis-1

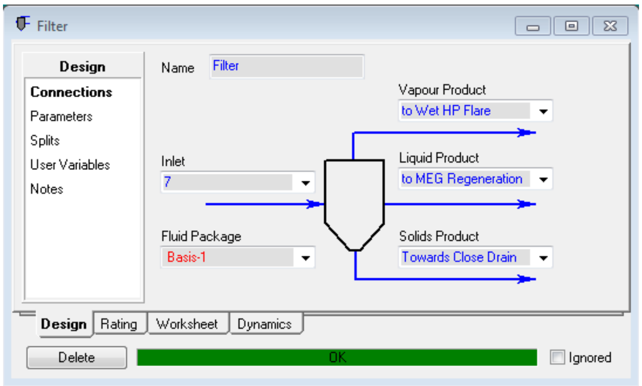
Figure 17: Heat Exchanger Data Input Window



Name: HP Steam Heater
Inlet: 4
Energy: Q-101
Outlet: 5
Fluid Package: Basis-1

Figure 18: Heater Data Input Window

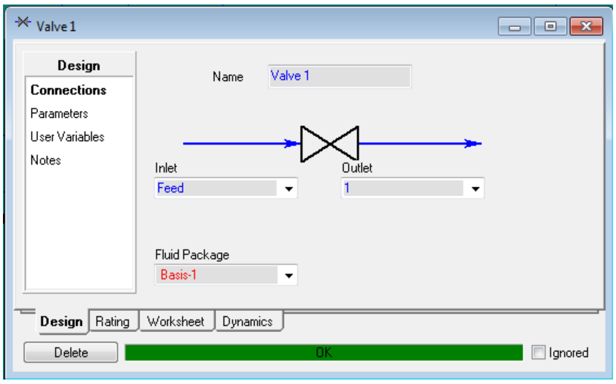
For simple solid filter, it needed to define 1 input stream and 3 output streams. Data input for the filter is by defining the steam fractions in term of mole fraction which solids in vapor, solids in liquid and liquid in bottoms is 0 mole fraction.



Name: Filter
Vapour Product: to Wet HP Flare
Inlet: 7
Liquid Product: to MEG Regeneration
Fluid Package: Basis-1
Solids Product: Towards Close Drain

Figure 19: Simple Solid Filter Data Input Window

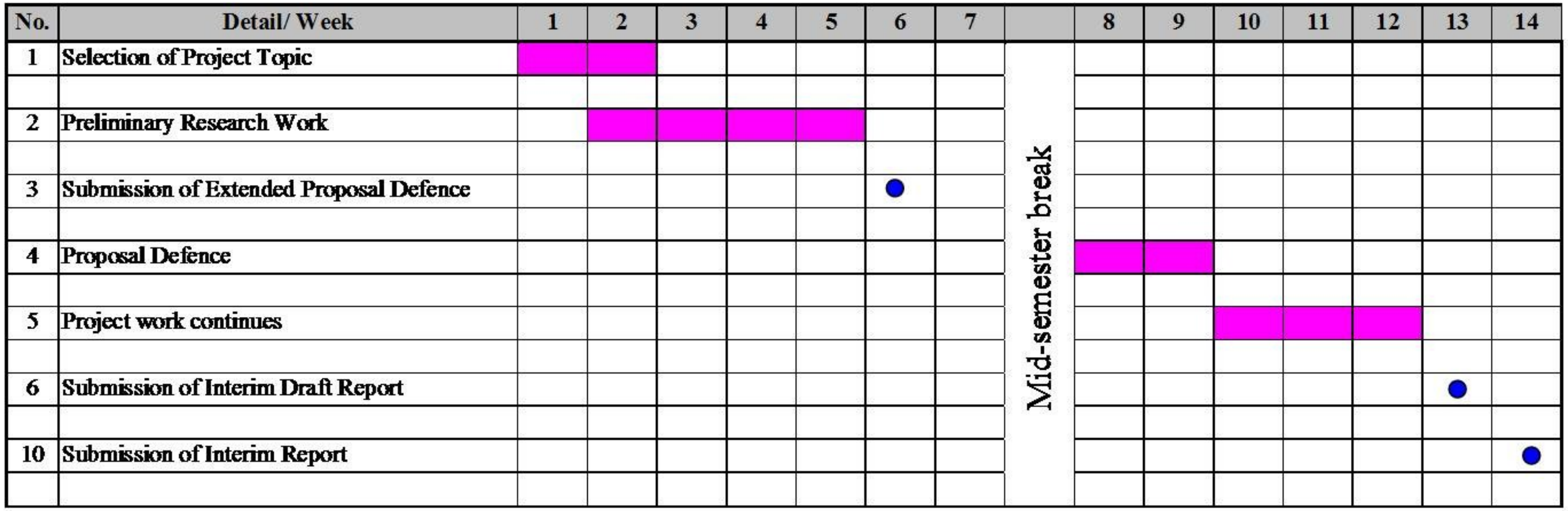
For valve, it is needed to define 1 input stream and 1 output stream. Data input for Valve 1 of temperature and pressure are 17.70◦C and 1251 kPa respectively, for Valve 2 are 143◦C and 1151 kPa and Valve 3 are 40.93◦C and 261.3 kPa.



Name: Valve 1
Inlet: Feed
Outlet: 1
Fluid Package: Basis-1

Figure 20: Valve Data Input Window

3.4 Activities/Gantt Chart and Milestone



No.	Detail/ Week	1	2	3	4	5	6	7		8	9	10	11	12	13	14
1	**Selection of Project Topic**	■	■						Mid-semester break							
2	**Preliminary Research Work**		■	■	■	■										
3	**Submission of Extended Proposal Defence**						●									
4	**Proposal Defence**									■	■					
5	**Project work continues**											■	■	■		
6	**Submission of Interim Draft Report**														●	
10	**Submission of Interim Report**															●

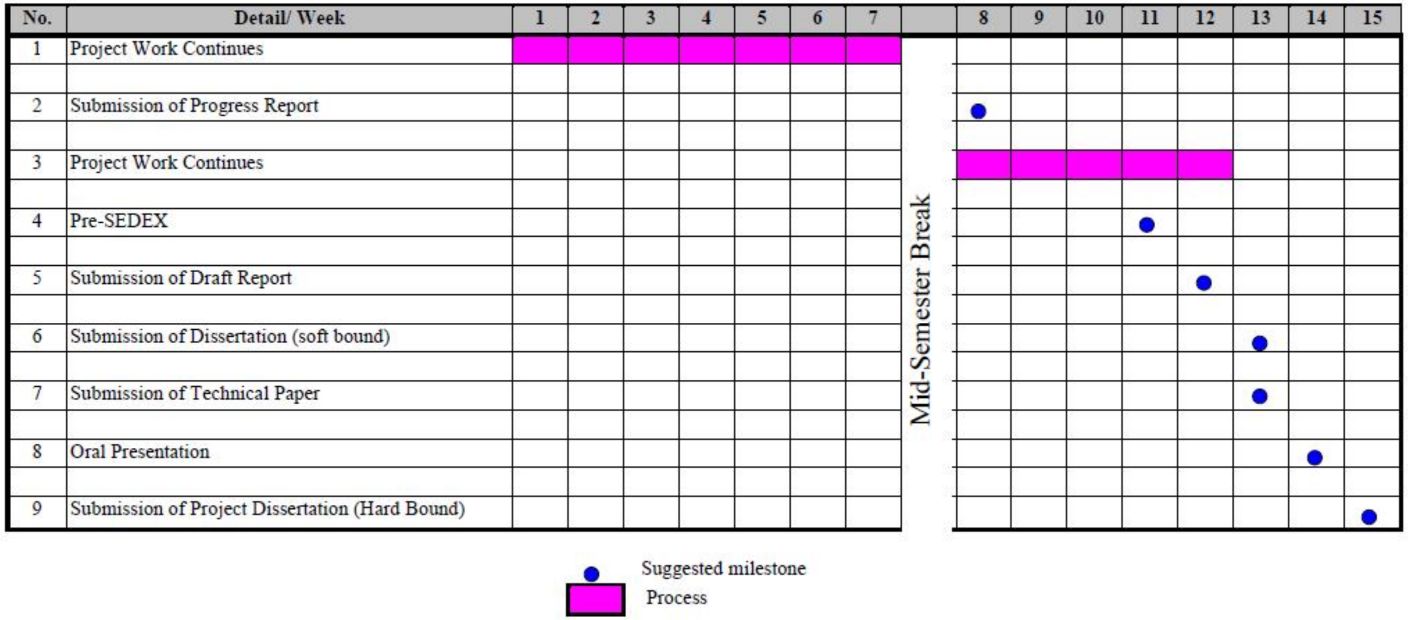
● **Suggested milestone**
■ **Process**

Table 4: Gantt chart FYP I

The tree main tasks to be completed for FYP I are:

a. Extended Proposal
b. Proposal Defense
c. Interim Report

For the second semester (FYP II), the project flow is to be carried out as in the Gantt chart below.



No.	Detail/ Week	1	2	3	4	5	6	7		8	9	10	11	12	13	14	15
1	Project Work Continues	🟪	🟪	🟪	🟪	🟪	🟪	🟪	Mid-Semester Break								
2	Submission of Progress Report									●							
3	Project Work Continues									🟪	🟪	🟪	🟪	🟪			
4	Pre-SEDEX												●				
5	Submission of Draft Report													●			
6	Submission of Dissertation (soft bound)														●		
7	Submission of Technical Paper														●		
8	Oral Presentation															●	
9	Submission of Project Dissertation (Hard Bound)																●

● Suggested milestone
🟪 Process

Table 5: Gantt chart FYP II

The main tasks for FYP II are:

a. Progress Report
b. Pre-SEDEX
c. Technical Paper
d. Oral Presentation
e. Dissertation

CHAPTER 4

RESULT AND DISCUSSION

4.1 Feed for the Process

Based on Figure 21, the envelope curve shows that the feed consists of 0.57 Liquid phases, 0.26 vapor phases and 0.18 aqueous phases. This shows that the feed has 3 phases which consists of gas, oil and water. Hence, in the process, we need to put the 3-phase separator to separate the feed to get the desired product. For the feed compositions refer to the Appendix A.

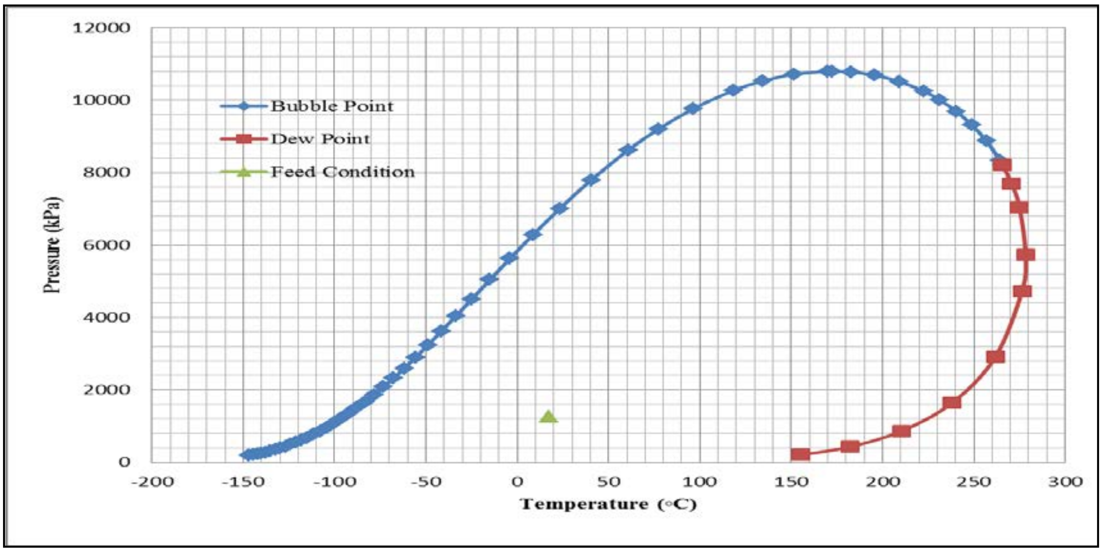


Figure 21: Envelope Curve of the Feed

The inlet of condition of the feed as follows:

Table 6: Total Properties of Feed

Properties	Value
Normal Flow, kmol/h	4645
Normal Flow, kg/h	325604
Heat Flow, kW	4009
Molecular Weight	70.1
Pressure, barg	11.5
Temperature, °C	17.7

Table 7: Total Properties in Vapor Phase of Feed

Properties	**Value**
Molar Flow, MMSCFD	24
Normal Flow, kg/h	25957
Density, kg/cu m @P,T	11.7

Table 8: Total Properties in Liquid Phase of Feed

Properties	**Value**
Standard Liq Vol Flow, SBPD	61349
Normal Flow, kg/h	299647
Actual cu m/h @P,T	389
S.G Liauid @P,T	0.770

4.2 Process Description

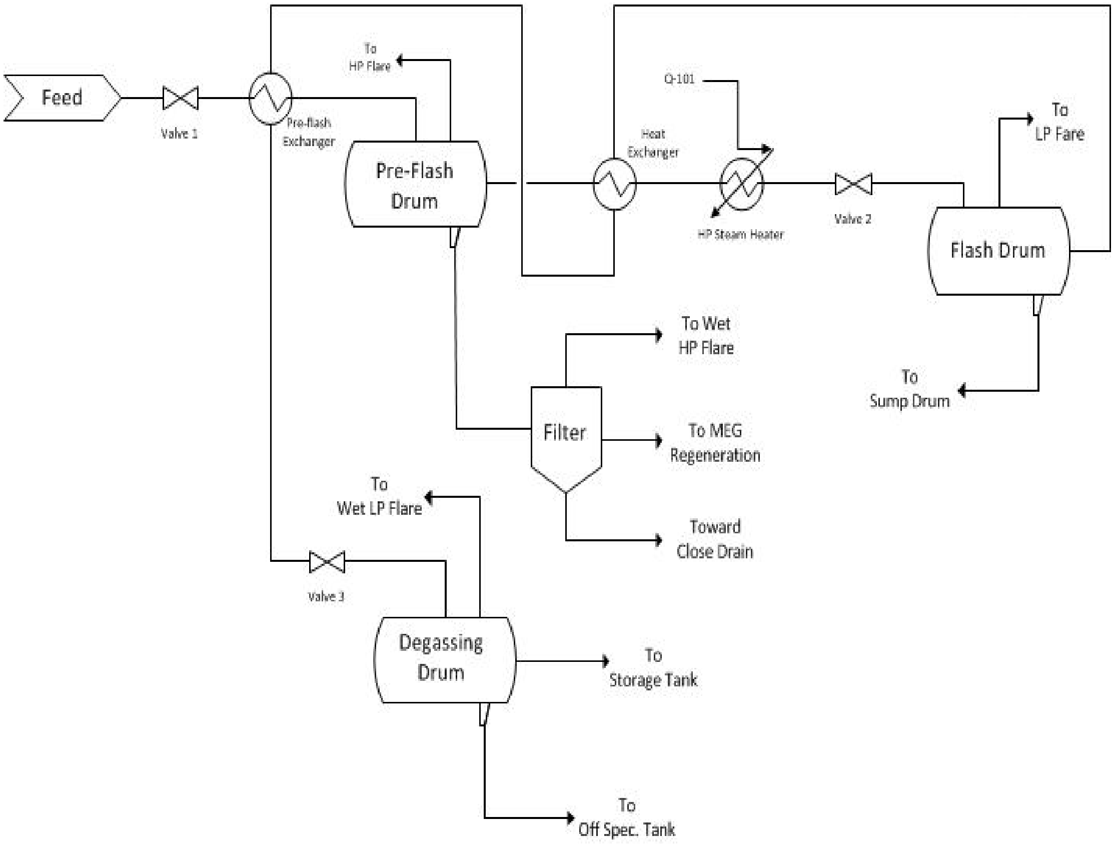


Figure 22: Process Flow Diagram of Simulated Process

The purpose of this process is separation of aqueous phase and gaseous hydrocarbon from the condensate and then to stabilize it for the export by adjusting Reid Vapor Pressure (RVP) which indicating the volatility of the condensate. This is because the quality of the product depends on composition and also RVP before selling to the customers.

Firstly, main feed from the onshore plant is entered to pre-flash drum to remove light hydrocarbons, most value of acid gases and lighter paraffin's will be excited in this step. Next, condensate temperature is increased in two sequential heat exchanger and High Pressure (HP) heater up to 80^{o}C and 143^{o}C respectively. Lastly, this fluid with crossing from of two first shell tube exchanger and degassing in the last flash drum is stored in storage tanks.

The off-gas for example light hydrocarbon like methane, ethane and propane, sulfur components like hydrogen sulfide, and others will be burnt in the appropriate flare system. For aqueous phase like MEG and others are sent to further processing in the suitable units for instance MEG regeneration unit. Besides that, components that have sulfur element like Mercaptans and also water will be sent to off specification tank and then will be transferred to the waste treatment.

4.3 Comparison of Actual Plant Data, Pro/II Software and HYSYS Software of Condensate Composition of Final Product at Normal Condition

For validation of data of final product, the obtained data have been compared with actual plant data in South Pars gas field (Assaluyeh, Iran), Pro/II software version 7.1 and HYSYS Software version 2006. This is because the author wants to see the composition which is valid for this simulation to build in the future. The result has shown below:

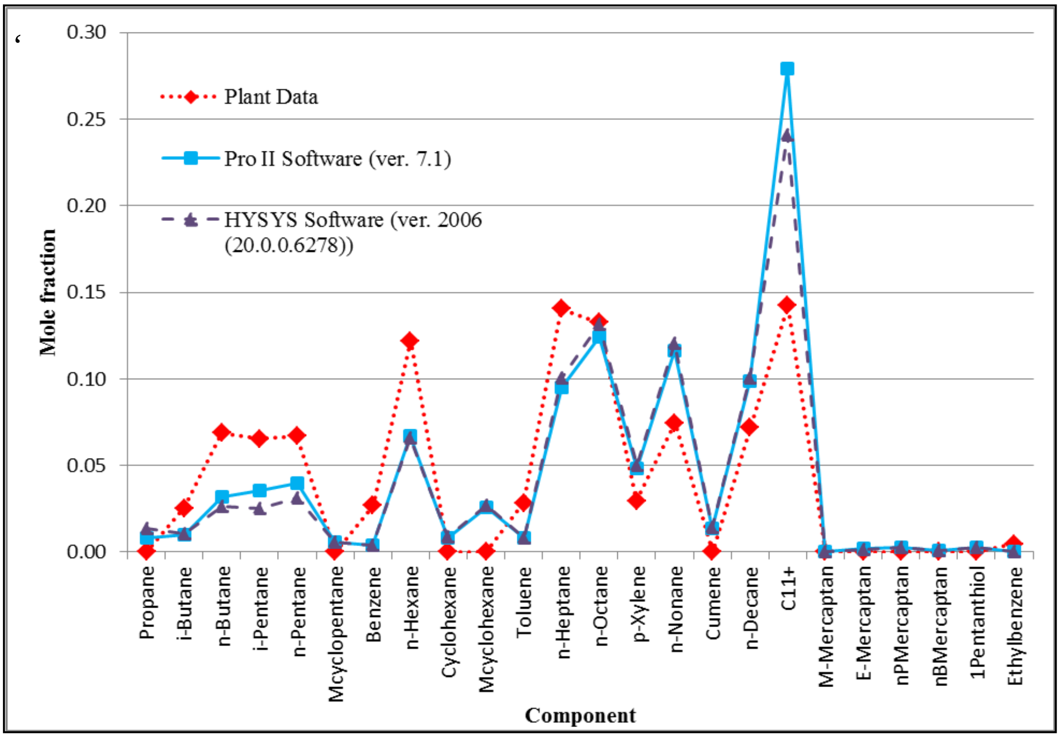


Figure 23: Overall Comparison of Plant Data, Pro/II Data and HYSYS Data

Based on the Figure 23, it can be seen that the simulation of the process is nearly same with the plant data. Hence, this HYSYS data will be validated to the real plant. Besides that, the Pro/II Software looks also the same data with the real plant. Overall; data of final product should be valid for simulation software in order to validate the result for this process.

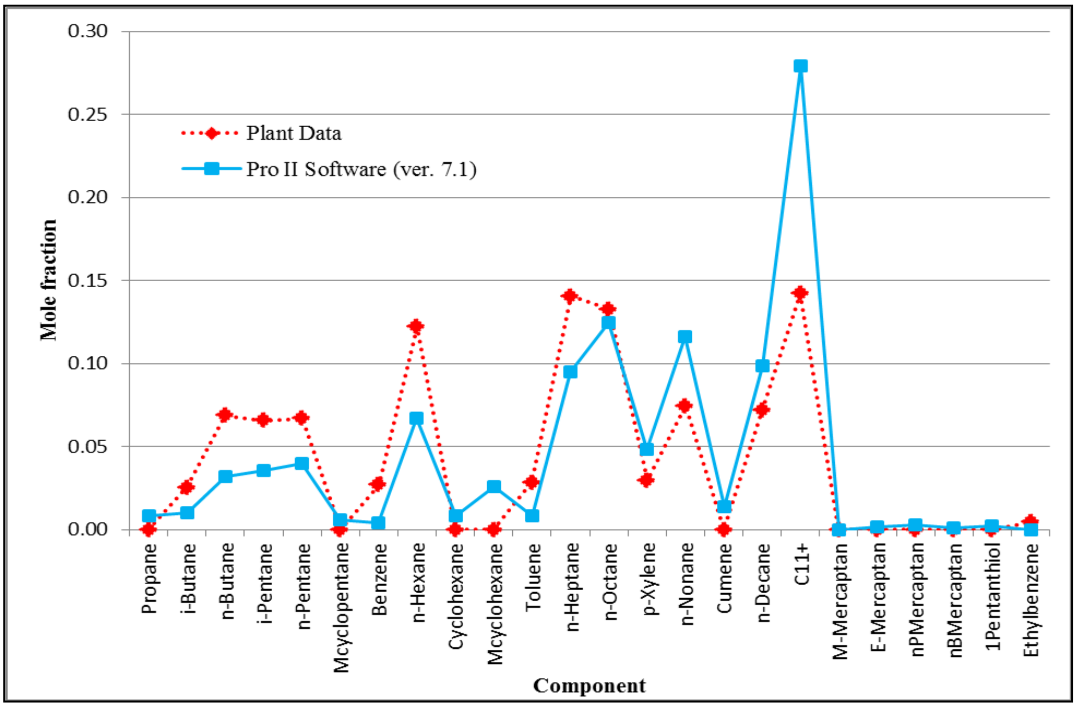


Figure 24: Comparison of Plant Data and Pro/II Data

Figure 24 illustrates the comparison of plant data and Pro/II Software data of final product which their data is slightly different. Light hydrocarbon like n-Butane, n-Pentane and Benzene, the mole fraction is lower than the plant data. It means that light hydrocarbon is flashed at higher rate before sending to the storage tank. This result shows that the process does not want to have light hydrocarbon which indicate the higher quality of the product.

Besides that, heavy hydrocarbons' mole fraction like Benzene, Cyclohexane and others show higher value in Pro/II Software. This means that the quality is higher as we want to have more mole fraction of heavy hydrocarbon in the final product which the customers' demand. Hence, it will increase the marketability of our product.

In addition to that, hazardous components that have sulfur element which are M-mercaptan, n-Pmercaptan and others is very small in mole fraction and also plant data and Pro/II Software data is nearly same. It shows that these component will not affect the quality of the final product and very safe to the process. It justify why this unit does not require desalter.

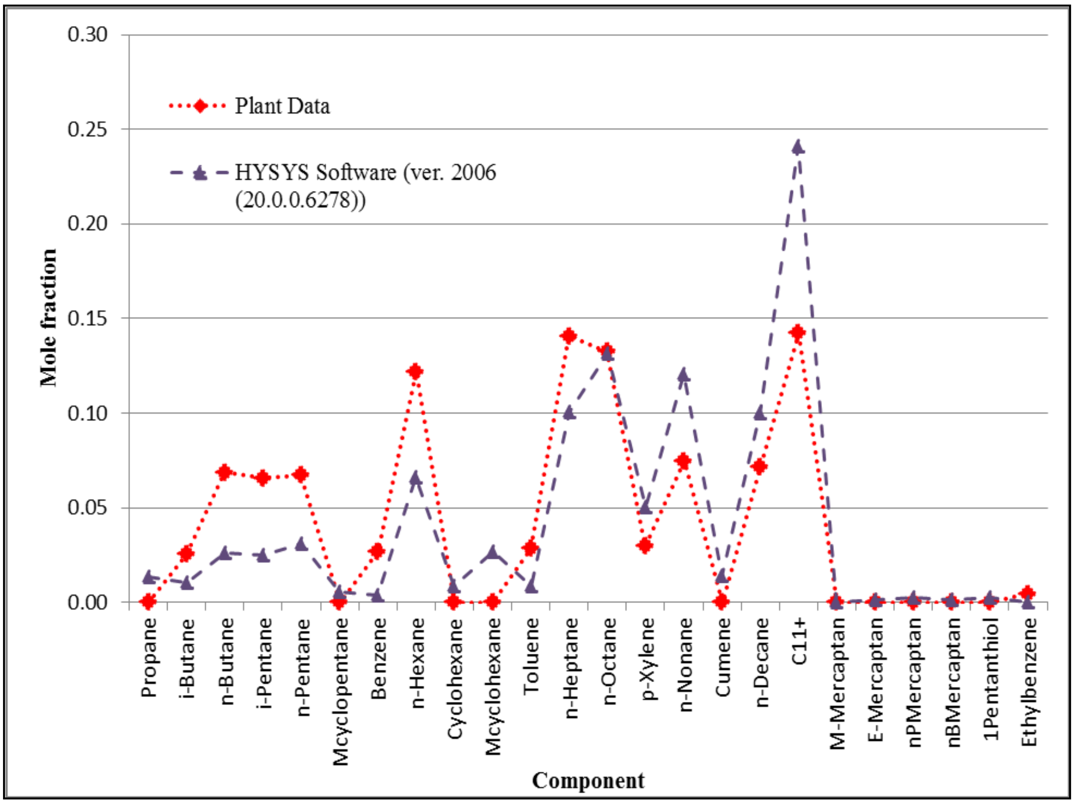


Figure 25: Comparison of Plant Data and HYSYS Data

Figure 25 shows the comparison of plant data and HYSYS data of final product of this process. It can be seen that their data is nearly same like the comparison between plant data and Pro/II Software data. Light hydrocarbon components shows in HYSYS data shows that their mole fraction is lower than the plant data which indicating the unwanted hydrocarbon is already flashed before sending to the storage tank. This will increase the quality of the product.

Furthermore, heavy hydrocarbon in the final product of HYSYS data shows that it is the nearly the same with the plant data. Although the plant data is slightly higher, we can consider that the quality of the product is the same as the plant data because it their differences are not affecting the overall data.

Besides, sulfur element which is contained in M-mercaptan, n-Pmercaptans also same with the plant which are very in small quantities. This shows that our final product should be safe to send to the customer and also the profit should be increased.

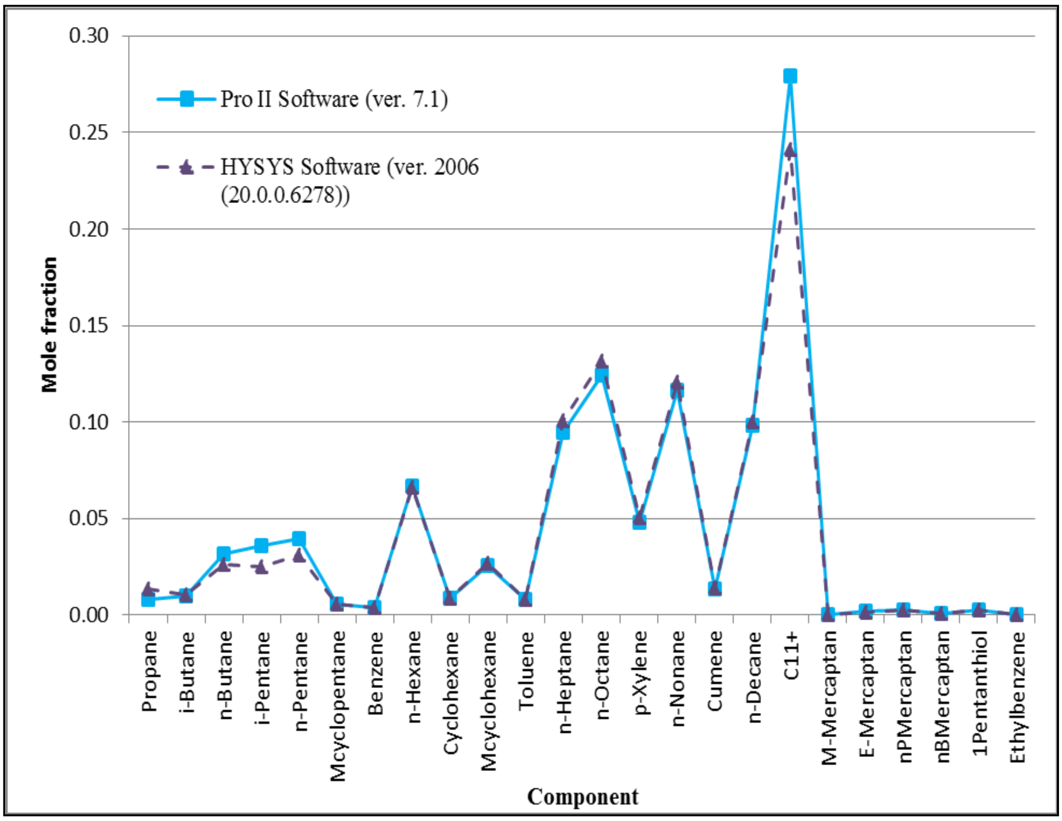


Figure 26: Comparison of Provision Data and HYSYS Data

Figure 26 gives information about the comparison of Pro/II Software Data and HYSYS data which overall look similar to each other's. For instance, light hydrocarbon like Propane, n-Butane and others in the HYSYS data have lower mole fraction compared with Pro/II Software Data. It indicates that the light hydrocarbons have been flashed out from the process in the HYSYS simulation compared to Pro/II Software data. It is essential that HYSYS is more reliable software in simulating the process.

Besides that, for heavy hydrocarbons like Benzene, Cyclohexane and others in HYSYS give the same with the Pro/II Software data. It shows that the condensate that we want have is nearly the same in the simulation and thus validate the process to get customer's desired product.

In addition to that, components that have the sulfur element in the HYSYS and Pro/II Software data had nearly mole fraction in both simulators. This shows that our final products have higher quality and the customer will be satisfied with the service.

4.4 Adjusting Process Parameter

For the simulation, the author wants to see the effect of different process parameter which in the reality, the process is not always in steady state. This is because many factors that can affect the process especially the quality of the product like surroundings condition, breakdown of equipment and others. Hence, it is essential that to know how much the effect of the operating parameters on the final product and also the best optimum conditions that process will be achieve in order to have process optimization.

Therefore, the author has recognized a few of process parameters that will change the final product specification which are changing steam temperature, steam pressure, feed flow rate, feed temperature and feed pressure. These parameters are simulated in one dimensional condition where others parameter is kept constant at a time. The product specification will be monitored by Reid Vapor Pressure (RVP), sulfur content and dominant component that has highest value in sulfur content against the operating parameter.

4.4.1 Effect of Steam Temperature

For this operating parameter, the author has maintained constant variables which are heat duty of heater, pressure of inlet and outlet of the stream 4 and stream 5 and temperature of inlet stream 4. The author has only changed the temperature of outlet stream 5 ranging from 139°C to 159°C. This is because we want to see the effect of the temperature before entering the flash drum on the final product. The result has shown in Figure 27:

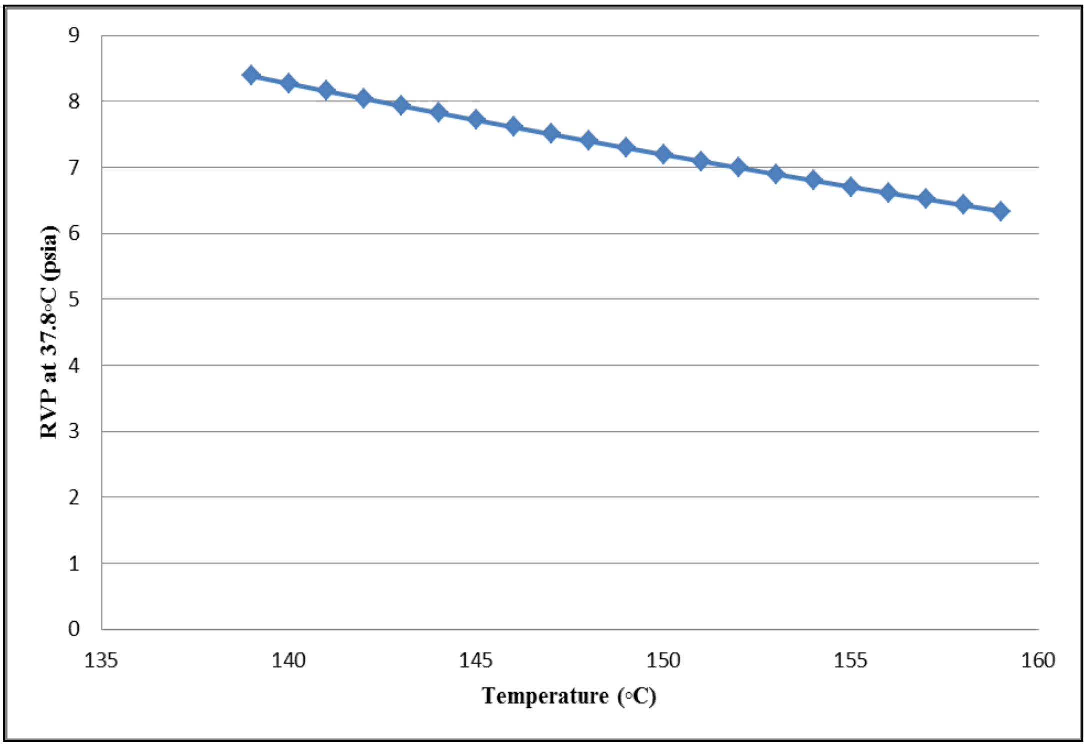


Figure 27: Variation of Steam Temperature in Back-Up Unit
to Find Optimum Condition

Based on the Figure 27, it can be seen that higher temperature gives lower RVP value. This means that higher temperature will remove more acid gases and light hydrocarbon which RVP changing between 8.385 psai and 6.336 psai. From this range, the best temperature for this process to avoid more loss of Propane and Butane as well as stripping corrosive and sour components to promote value of the product is 143°C which causes RVP is 7.932 psia.

For the sulfur content, the result has shown in Figure 18 and Figure 19.

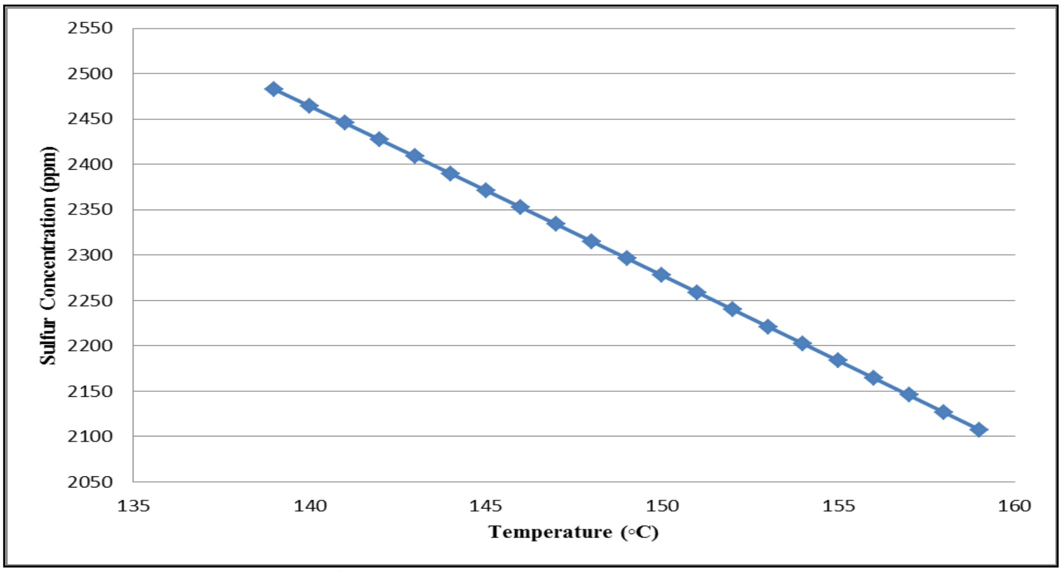


Figure 28: Variation of Steam Temperature against Sulfur Concentration in Back-up Unit

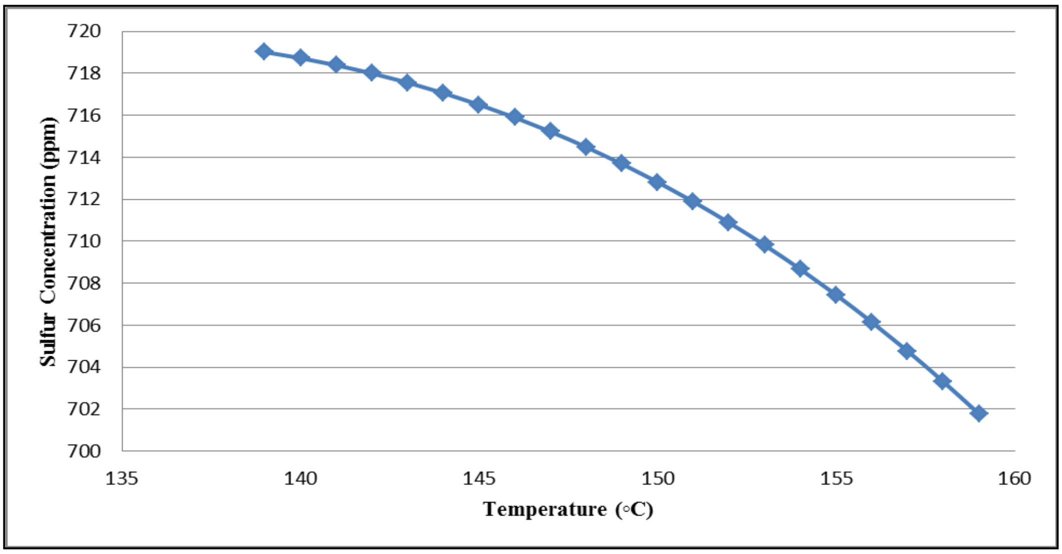


Figure 29: Effect of Dominant of Sulfur Element against Steam Temperature in Back-up Unit

From Figure 28 and Figure 29, it can be seen that the concentration of sulfur decrease as temperature of steam HP heater increase. This is because the components which contain sulfur elemenet will be removed rapidly as higher temperature and it will flash the acidic component. The highest sulfur concentration is 2500 ppm which is very high at low temperature and should be removed in this stage. For dominant of component which contain sulfur element in this operating parameter is 1Pentanthiol and it show that the sulfur concentration is decreasing as the temperature is increasing which it should be removed as high as possible as it can affect the quality of the product.

4.4.2 Effect of Steam Pressure

For this operating parameter, the author has maintained some others parameter which are the heat duty of the heat exchanger, the temperature of inlet and outlet for both cold and hot stream in steam HP heater, steam flow rate, and pressure of outlet steam stream. To obtain the results, the author has only changed the pressure of inlet steam ranging from 10 kPa to 65kPa. This operating parameter should affect on RVP and also sulfur content.

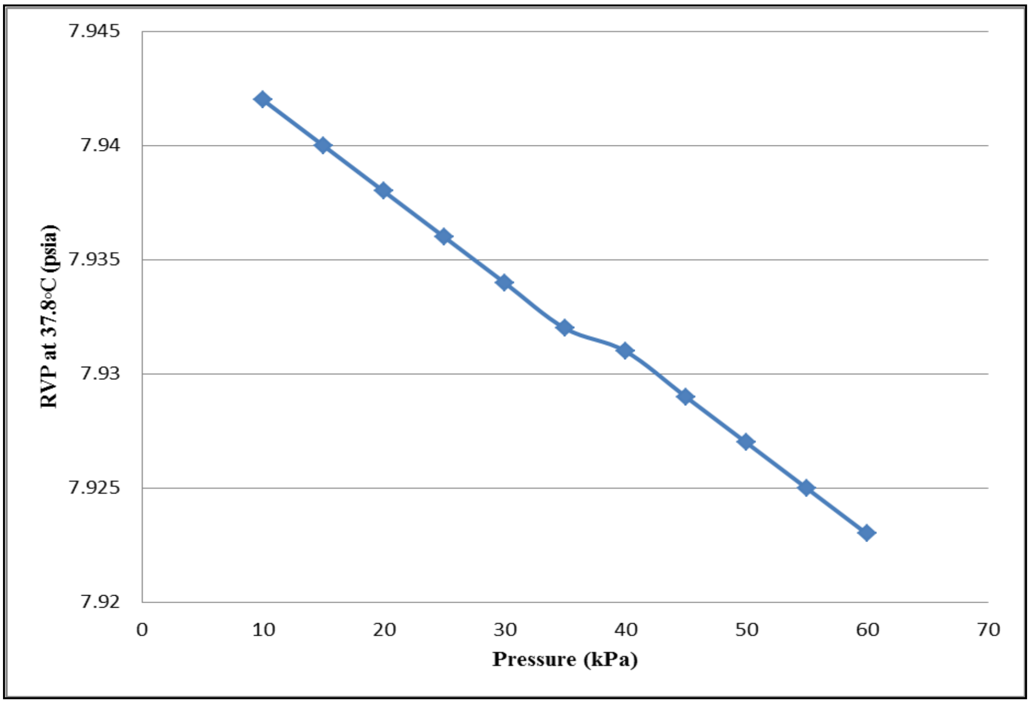


Figure 30: Variation of Steam Pressure in Back-Up Unit to Find Optimum Condition

Based on the Figure 30, it shows that RVP is decreasing as steam pressure is increasing. The lowest of pressure is 10kPa and the highest pressure is 65kPa as lower pressure and higher pressure in this range will give temperature cross in the heat exchanger which is not valid for this process. From this range of the steam pressure, it will cause the RVP changes from 7.942 to 7.921 psia which is the best optimum condition is 35 kPa to remove the unwanted hydrocarbon and also stripping sour component which cause RVP is 7.932 psia based on changing steam temperature. It means that higher steam pressure will increase the steam heat duty. As the result of higher steam heat duty, there are more flashing of acidic gases.

For sulfur content, the result has shown in Figure 23 and Figure 24.

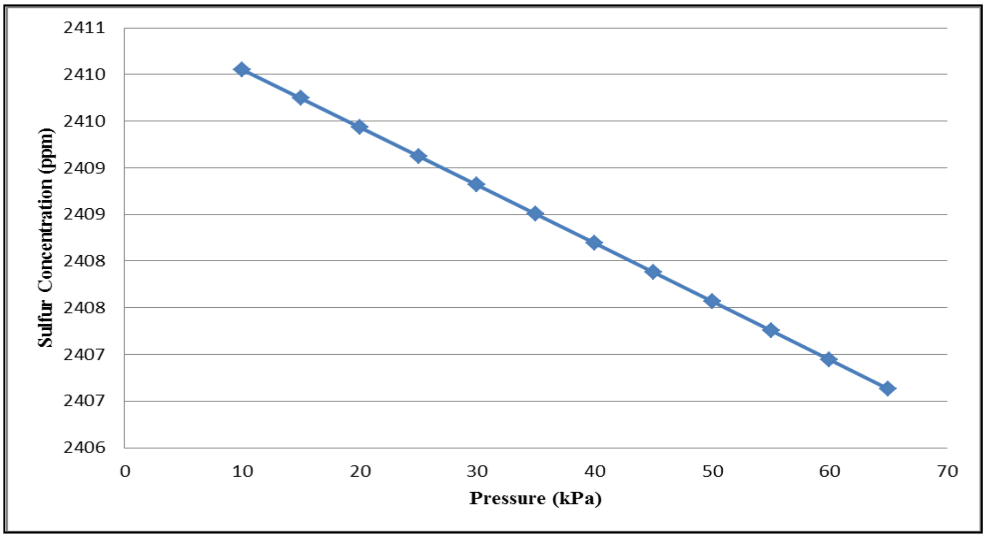


Figure 31: Variation of Steam Pressure against Sulfur Concentration in Back-up Unit

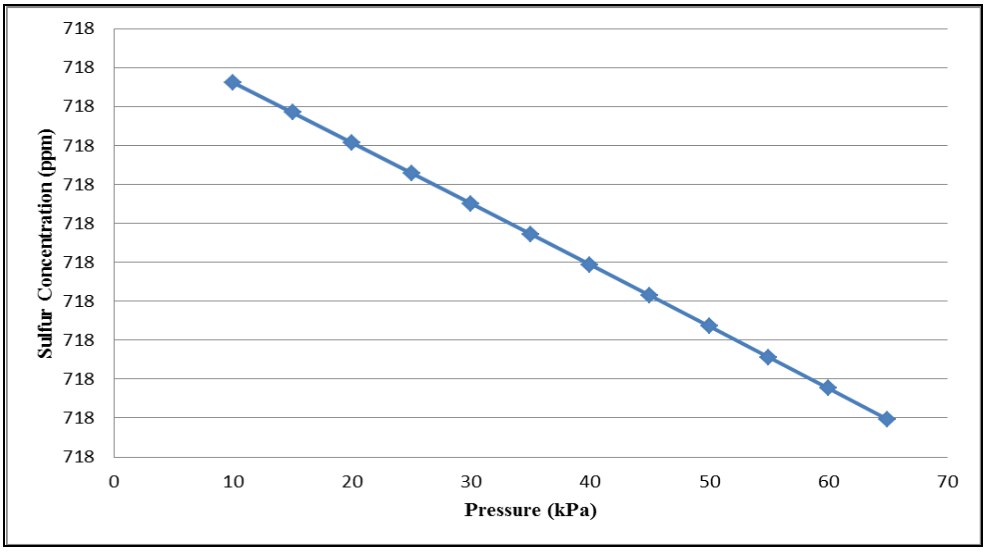


Figure 32: Effect of Dominant of Sulfur Element against Steam Pressure in Back-up Unit

From Figure 31 and Figure 32, it shows that the sulfur concentration is decreasing as steam pressure is increasing. From this trend, it can be seen that higher pressure will remove the components which contain sulfur element faster in the in separator. The highest sulfur concentration is 2410.05 ppm and needed to reduce as low as possible by increasing the steam pressure. For dominant of component which contain sulfur element in this operating parameter is 1Pentanthiol and its concentration is decreasing as pressure is increasing. This is good condition to remove the sulfur as high as possible.

4.4.3 Effect of Feed Flow Rate

For this operating parameter, the author has maintained some constant variables like heat duty of heat exchanger, feed temperature, feed pressure and steam flow rate. The author has only changed the feed flow rate ranging from 1858 kmole/hr to 6038.5 kmole/hr which is in term of percentage 40% to 130% and the original one is 4645 kmole/hr. This is because we want to see the changes when the plant will turndown or overflow of feed flow rate. The result has shown in Figure 25.

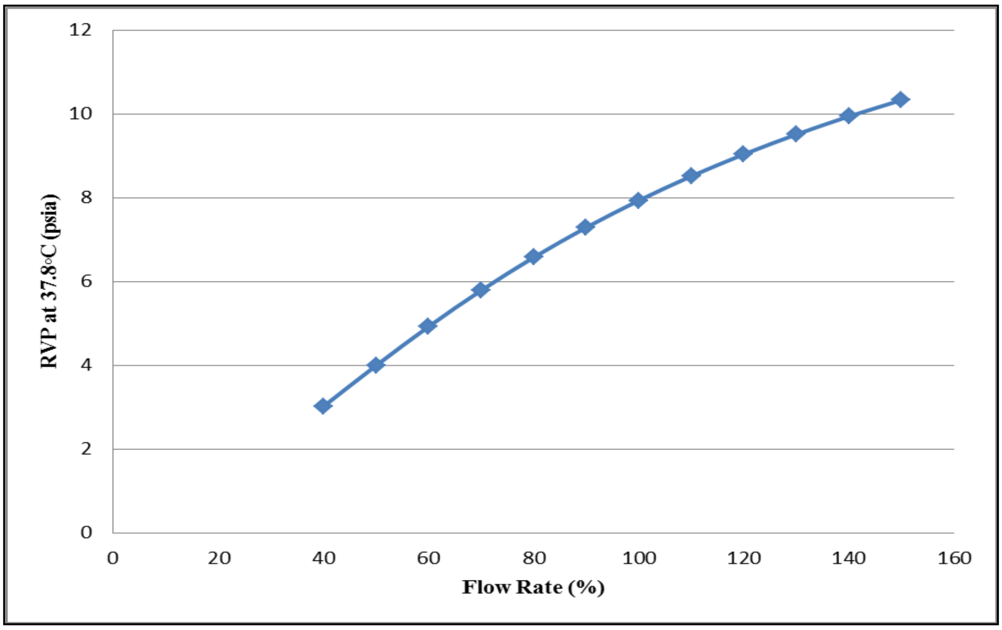


Figure 33: Variation of Feed Flow Rate in Back-Up Unit to Find Optimum Condition

Based on the Figure 33, it can be seen that RVP is increasing as feed flow rate is increasing**.** This is because there are a lot feed to be separated in the separator which cause higher heat required for the heater to supply the heat to the separator. As the result, RVP will increase as insufficient heat to maintain the operation of the separator. From this trend, at 1848 kmole/hr which is 40% from the original one, the plant will turn down as there will be a temperature cross in the heat exchanger. Furthermore, at 5574 kmole/hr (120%), the feed will be overflowed because temperature cross also occurred in the heat exchanger. Therefore, the optimum condition for feed flow rate is ranging from 50% to 110%.

For sulfur content, the result has shown in Figure 26 and Figure 27.

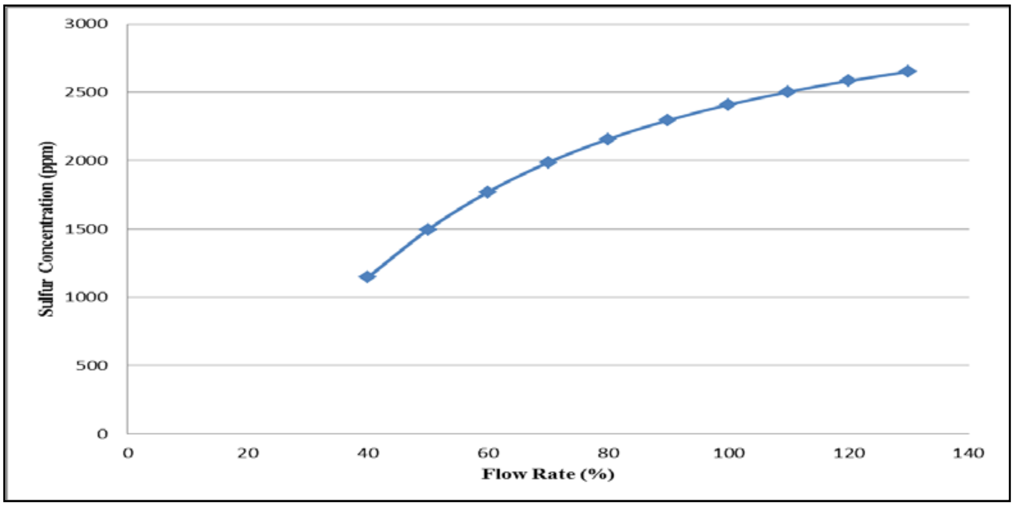


Figure 34: Variation of Feed Flow Rate against Sulfur Concentration in Back-up Unit

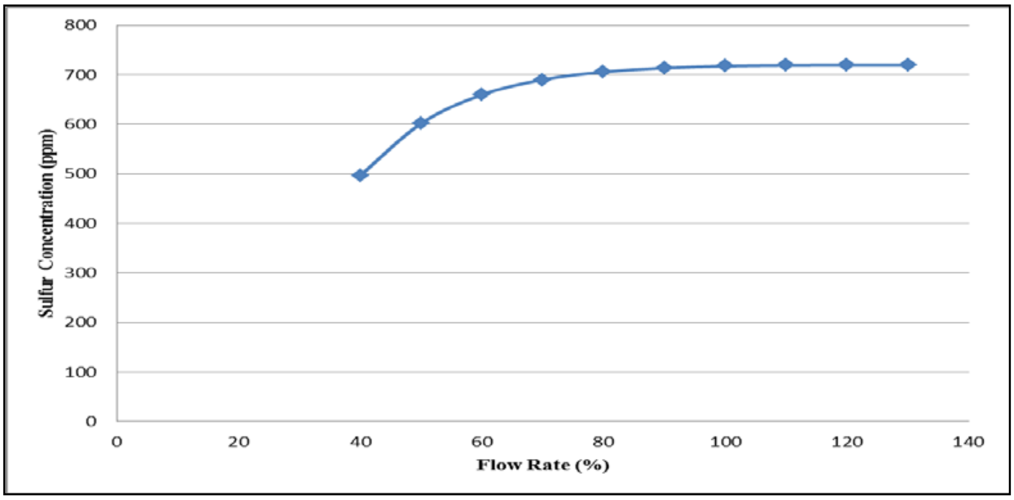


Figure 35: Effect of Dominant of Sulfur Element against Feed Flow Rate in Back-up Unit

From Figure 34 and Figure 35, it can be seen that sulfur concentration is increasing as feed flow rate is increasing. This is because there are more feed come into the process which they will a lot of product as well as the components that contain sulfur element. Hence, to decrease the sulfur concentration in final product, the feed flow rate should be low. The lowest of sulfur concentration is 1494.14 ppm at 50% of feed flow rate and the highest of sulfur concentration is 2502.97 ppm at 110% of feed flow rate. For dominant of component which contains the highest sulfur concentration is 1Pentanthiol and the sulfur concentration is increasing as feed flow rate is increasing.

4.4.4 Effect of Feed Temperature

For this operating parameter, the author has maintained some others parameter which are heat duty of the heat exchanger, feed pressure, feed flow and also steam flow rate. The author has only change the feed temperature ranging from -100◦C till 40◦C. When changing parameter, we want to see the effect on RVP as well as sulfur concentration which both of them can affect the quality of final product. The result has shown in Figure 28.

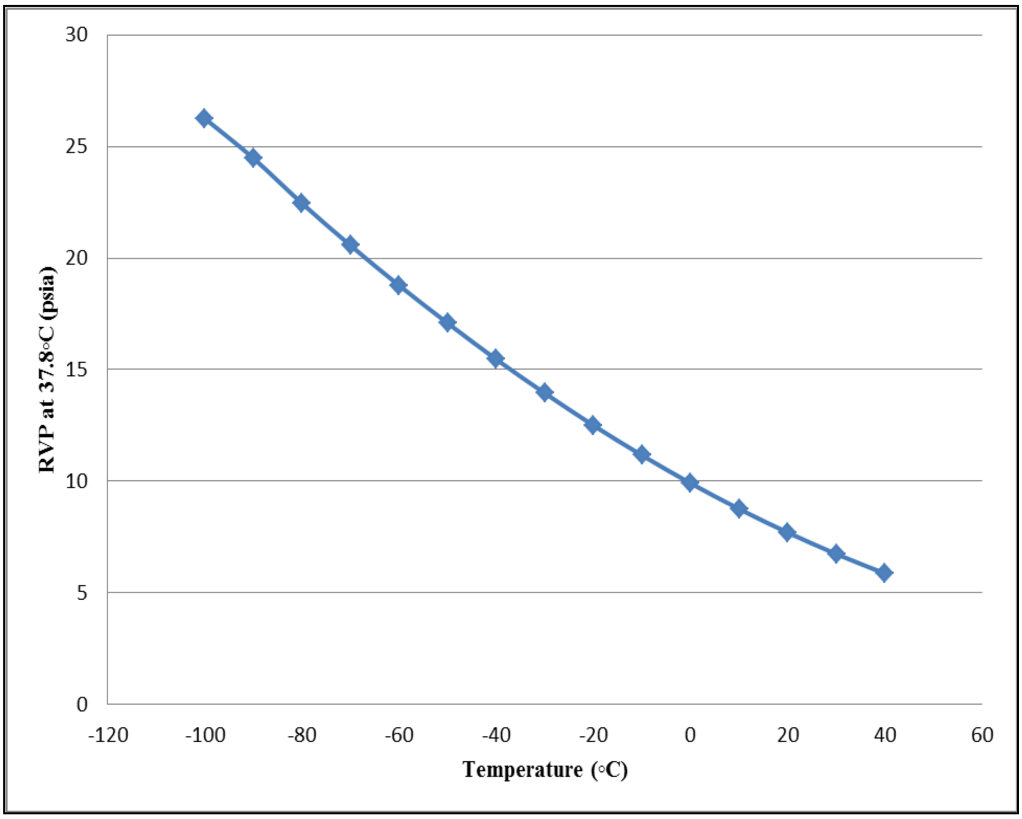


Figure 28: Variation of Feed Temperature in Back-Up Unit to Find Optimum Condition

Based on the Figure 36, it can be seen that RVP is decreasing as feed temperature is increasing. This result shows that we want to have lower RVP which we want to recover the product and can be sold at larger quantities. From the summer case which at 10 psai the range of feed temperature should be -10◦C till 20◦C and the original feed temperature is 17.7◦C which causes 7.932 psai. However, 30◦C and higher of feed temperature will cause temperature cross in the heat exchanger and the best condition for the process is 10◦C till 20◦C.

For sulfur content's results shown in Figure 29 and Figure 30.

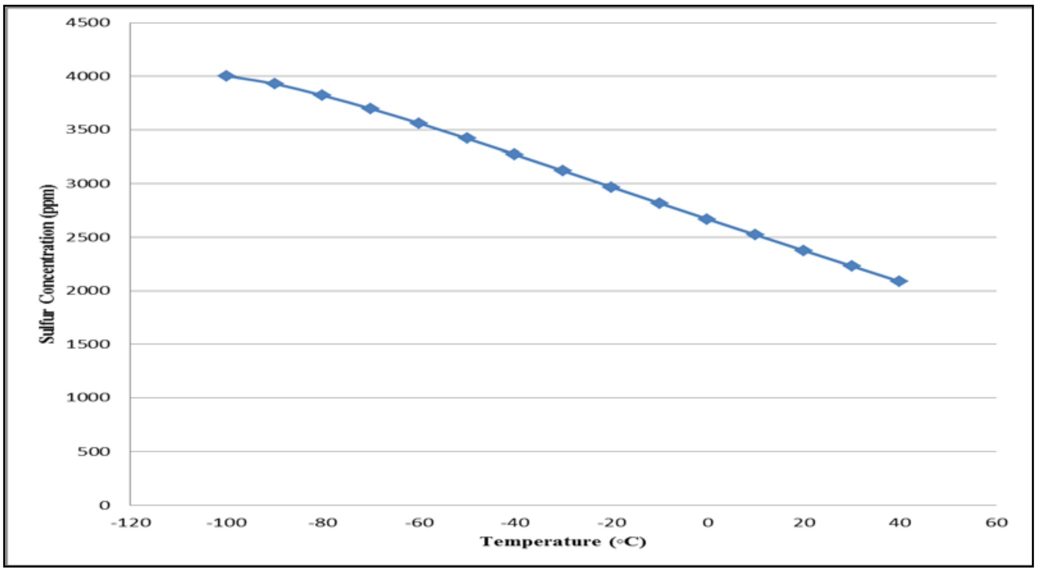


Figure 37: Variation of Feed Flow Rate against Sulfur Concentration in Back-up Unit

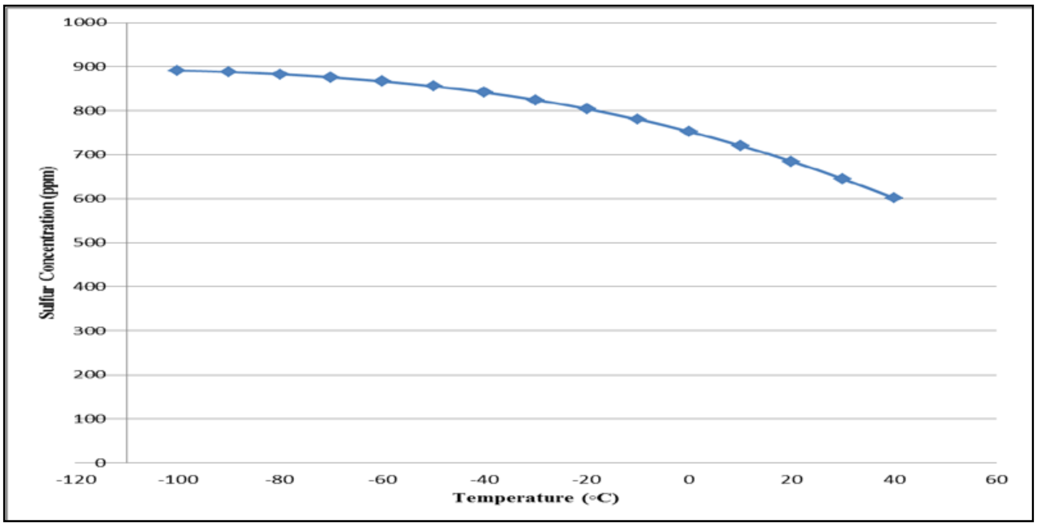


Figure 38: Effect of Dominant of Sulfur Element against Feed Temperature in Back-up Unit

From Figure 37 and Figure 38, it can be seen that sulfur concentration is decreasing as feed temperature is increasing. This is because the unwanted components including containing sulfur element have been removed at higher temperature. Therefore, the feed temperature should be higher as possible until it does not go against the temperature difference in the heat exchanger which is 20◦C. The lowest sulfur concentration is 2375.65 ppm and the highest of sulfur concentration is 4002.05 ppm. For dominant of component which contains the highest sulfur concentration is nPMercaptan (890.98 ppm at 110◦C).

4.4.5 Feed Pressure

For this operating parameter, the author has maintained some constant variables like heat duty for heat exchanger, feed flow rate, feed temperature and also steam flow rate. The author has only changed the feed pressure ranging from 1170 kPa to 1300 kPa as we want to see the effect on RVP and also sulfur concentration. The result has shown in Figure 31.

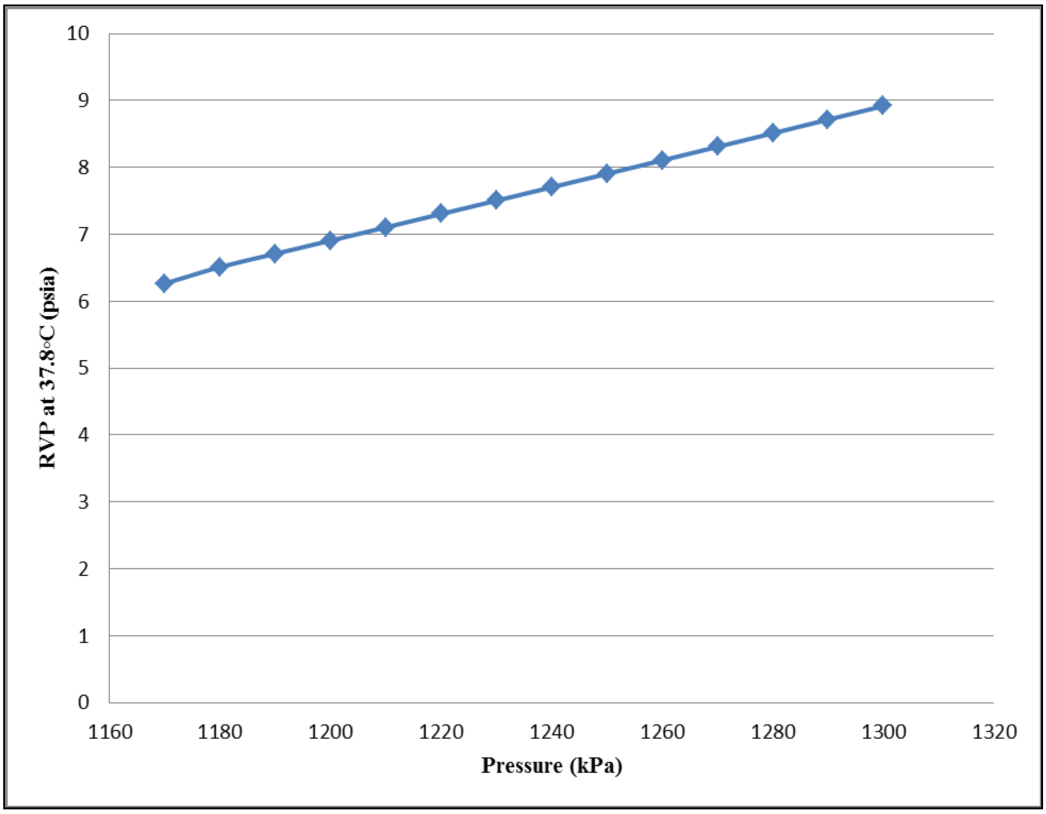


Figure 40: Variation of Feed Pressure in Back-Up Unit to Find Optimum Condition

Based on the Figure 40, it can be seen that RVP is increasing as feed pressure is increasing. This because higher pressure of the feed will cause the feed to become liquid phase as in the 3-phase separator's pressure should be as low as possible to flash off the acidic gases. The lowest of pressure is 1200 kPa as below from that, there will be temperature cross and it is the lowest pressure that can be used in the process. From the range 1200 kPa till 1300 kPa, they cause the RVP changes from 6.908 psai to 8.919 psia and the best condition is 1251 kPa which causes RVP 7.932 psai. This shows that feed pressure is one of the factor that will affect the process especially RVP.

For sulfur's content has shown in Figure 32 and Figure 33.

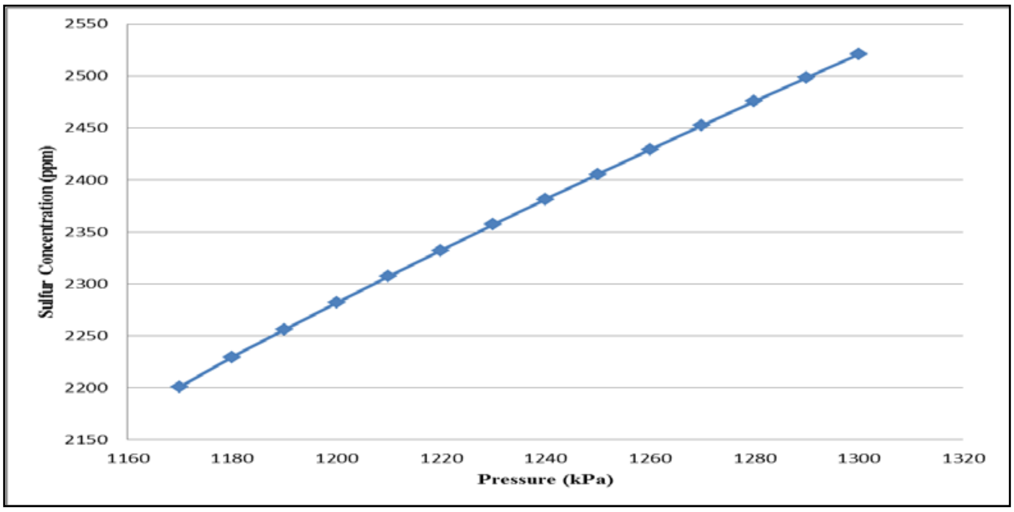


Figure 41: Variation of Feed Pressure against Sulfur Concentration in Back-up Unit

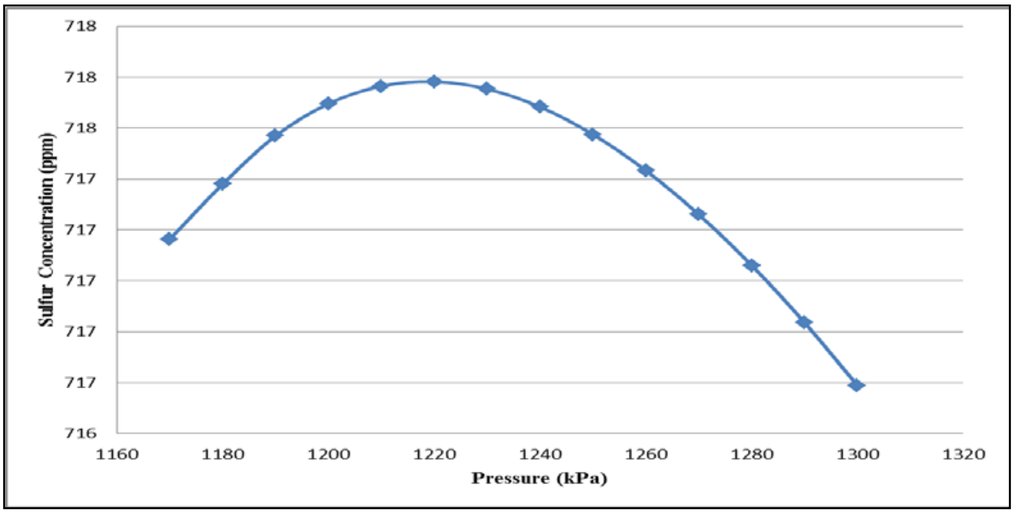


Figure 42: Effect of Dominant of Sulfur Element against Feed Pressure in Back-up Unit

From Figure 41 and Figure 42, it can be seen that the sulfur concentration is increasing as feed pressure is increasing. From this trend, higher temperature will cause the components that contain sulfur element will not be removed in the separator. The lowest sulfur concentration is 2281.76 ppm (1200 kPa) and feed pressure should be low as possible until it not go against the temperature cross in the heat exchanger. For dominant of component which contain sulfur element in this operating parameter is 1Pentanthiol and its concentration is increase as feed pressure is increasing but it decrease at 1220 kPa as 1Pentanthiol in highly water solubility and they can be flashed off in term of gas phase. It needs us to consider the factor of feed pressure.

CHAPTER 5

CONCLUSION AND RECOMMENDATIONS

5.1 Conclusion

This project is carried out based on two main objectives, which are simulating a back-up Condensate Stabilization Unit (CSU) that is able to bring down the Reid Vapor Pressure (RVP) of the Summer Rich Condensate of maximum 10 psia for summer season and 12 psia for winter season and finding the best operating parameters for each of the equipment in a back-up condensate stabilization unit.

For validation data of this project, the data have been compared with the actual plant in Iran and also Pro/II Software. From the comparison, the results show the composition from each of data is nearly and very feasible to build in Malaysia. Although there are some data is deviated from actual data plant a little bit, it does not concern with the simulation.

This research shows steam temperature, steam pressure, feed flow rate, feed temperature and feed pressure are important parameters to adjust the amount of RVP as well as sulfur concentration. It has been found that for steam temperature which the most optimum condition is 143◦C that gives RVP 7.932 psia which is below than 10 psia in summer season. Hence, it is very essential for these parameters to be monitored closely which unfavorable content should be in specified range and to ensure that they will not exceed the limit that affect the overall quality of final product.

Besides that, in the literature review, there are some studies about the sulfur content and salt which can affect the back-up CSU in term of equipment and also final product. From the studies, it shows that they give slight effect which the salt is affecting the column reboilers which there are no any column in the back-up CSU and sulfur needs to be treated for more in order to produce the quality product.

5.2 Recommendation

The projects objectives were successfully achieved and continuation on the project lays the possibility of extending the project actual potential. There are some recommendations of this research that can be used in order to build in the future. The recommendations are as below:

a. Another parameter that can be studies on the effect of RVP and also sulfur concentration is steam flow rate.

b. Comparing the data with the feed from Malaysian market or reservoir so that the back-up CSU can be built in Malaysia.

c. 3-phase parameters like temperature and pressure can be also studied on the effect of RVP and sulfur content.

d. Costing of economic in terms utilities and also equipment should be also considered in order to maximize the cost.

e. Additional of equipment like 3-phase separator can also be investigated as there more separators, the higher of flashing off the unwanted component.

f. Using another simulator like iCON and other will give different of final composition of product which to validate the previous parameters.

REFERENCES

[1] Mokhatab, S., Poe, W.A. and Speight, J.G. 2006, *Handbook of Natural Gas Transmission and Processing*, UK, Gulf Professional Publishing, Page 281.

[2] Discovery Channel. 20 June 2012 <http://curiosity.discovery.com/question/what-is-reid-vapor-pressure>.

[3] A. Khanmohammadi and Kh. Mohammadbeigy, 2007, “Studying of the Effectiveness Parameters for Condensate Stabilization”, Iran; Research Institute of Petroleum Industry.

[4] Arash Esmaeili, 2010, “Optimization of effective parameters on Reid vapor pressure (RVP) in an Iranian Condensate Stabilization plant and a Back-up unit”, Taipei; The 13th Asia Pacific Confederation of Chemical Engineering Congress.

[5] Benoy, J. and Kale, R.N. 2010, “Condensate Stabilization” in *Offshore World,* India; Chemtech Foundation.

[6] James H. Gary and Glenn E. Handwerk (2001). *Petroleum Refining: Technology and Economics* (4th ed.). CRC Press. ISBN 0-8247-0482-7.

[7] J. Javier Vazquez-Esparragoza, Gustavo Arturo Iglesias-Silva, Michael W. Hlavinka,Jerry A. Bullin, “How To Estimate Reid Vapor Pressure (RVP) of Blends”, USA, Bryan Research & Engineering, Inc.

[8] Knovel. 2 September 2012 < http://www.google.com.my/url?sa=t&rct=j&q=&esrc=s&source=web&cd=1&cad=rja&ved=0CDEQFjAA&url=http%3A%2F%2Fkolmetz.com%2Fpdf%2FEDG%2FENGINEERING_DESIGN_GUIDELINE__separator%2520vessel_REV01.pdf&ei=H4CgUOXsCI3QrQffxYG4Bg&usg=AFQjCNHVW1hzNrU6q86eeAgGU5vQ4T7KYQ>.

[9] Bubbico. Reborto, 2 October 2012 <http://www.scribd.com/doc/50371317 /comparison-horizontal-vs-vertical-separator>

[10] Trifilieff, O., Wines, Thomas H. and Daire, F. 2011, *Improve Hydrocarbon Condensate Dehydration Performance Diagnostics and Solutions*, UK, Pall Corporation.

[11] Cameron Inc., 2010, *NATCO Electro-Dynamic Desalter*, USA, Cameron Inc.

[12] Bojes, J., Lerbscher, J., and Wamburi Wellingtion. 2010, *Elemental Sulfur in 3-Phase Sour Gas Systems – Is Condensate Really Your Ally?.* USA, Baker Hughes Inc.

[13] Oild Industry Jargon De-coded, 1 November 2012 <www.naturalhub.com/slweb/defin_oil_and_Natural_Gas.html>

[14] De Jong, Jan-Peter., Dowling N., Sargent M., 2007, *Effect Of Mercaptans And Other Organic Sulfur Species On Hightemperature Corrosion In Crude And Condensate Distillation Units*, USA, NACE International.

[15] Elenbaas D., Mohamed D., Khan M. Hidayati D., Huang F., and Lee R., 1996, *The Environmental Impact of Mercaptans in Underground Storage*, USA, Society of Petroleum Engineers, Inc.

[16] Bader A., Charles E. Baukal, and Bussman W, 2011, *Selecting the Proper Flare Systems*, USA, American Institute of Chemical Engineers.

[17] Environmental Quality (Clean Air) Regulation 1978, 6 November 2012 <http://www.google.com.my/url?sa=t&rct=j&q=&esrc=s&source=web&cd=1&cad=rja&ved=0CCEQFjAA&url=http%3A%2F%2Fcp.doe.gov.my%2Fcpvc%2Fwp-content%2Fuploads%2F2011%2F04%2FRegulations%2FEnvironmental_Quality(Clean_Air).pdf&ei=GNyXUOioM8XtrQeHiYHgAQ&usg=AFQjCNHK8N9exE_Xjz-VmAnviHJmjju5zA&sig2=ylSrrhKUO2UUB6VPkFhgGA>

[18] Fatin Nadiah, R. Technology Department, PETRONAS Penapisan Melaka, Melaka. Personal Interview. Nov. 7. 2012.

APPENDICES

Appendix A: Feed Composition

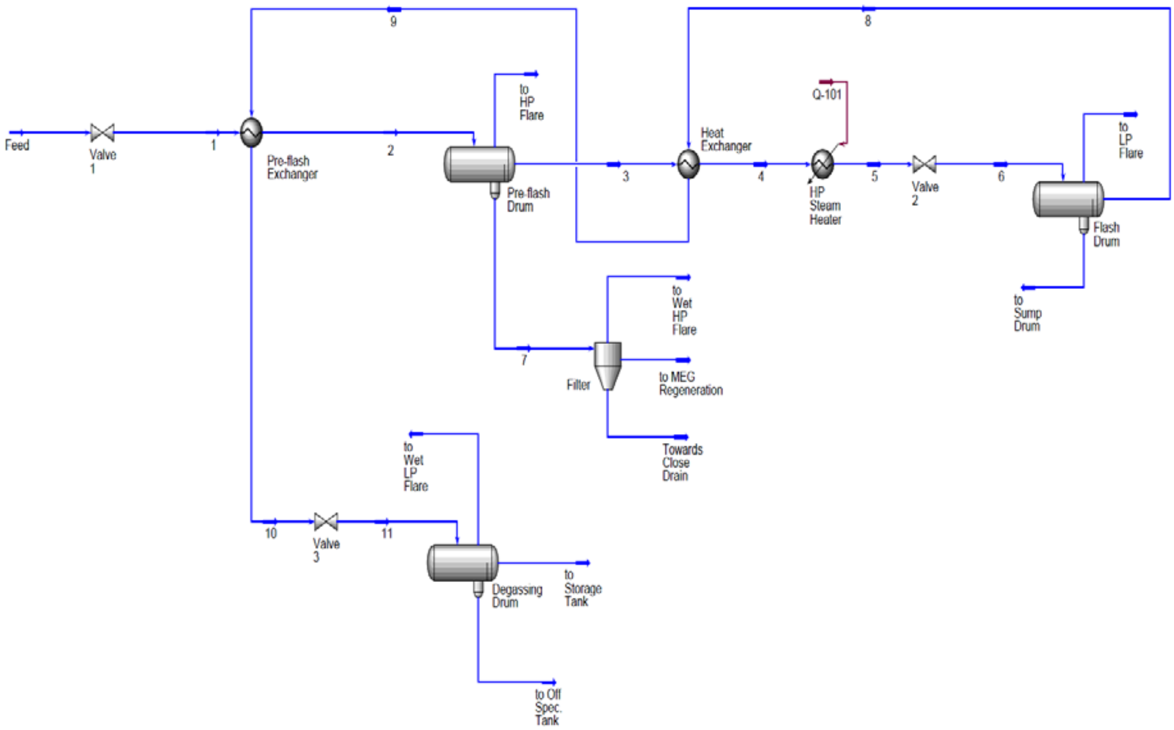
Component	Mole Fraction
Methane	0.218041
Ethane	0.054396
Propane	0.051802
i-Butane	0.018891
n-Butane	0.038908
i-Pentane	0.022982
n-Pentane	0.025847
Mcyclopentane	0.003284
Benzene	0.002242
n-Hexane	0.037976
Cyclohexane	0.004601
Mcyclohexane	0.012375
Toluene	0.003805
n-Heptane	0.046731
n-Octane	0.054126
p-Xylene	0.020163
n-Nonane	0.046275
Cumene	0.005448
n-Decane	0.037223
C11+	0.087779
Nitrogen	0.002623
Carbon Dioxide	0.012015
Hydrogen Sulphide	0.010165
Water	0.129249
M-Mercaptan	0.000130
E-Mercaptan	0.001688
COS	0.000007
nPMercaptan	0.001478
nBMercaptan	0.000505
1Pentanthiol	0.001092
MEG	0.048154

Appendix B: Clean Air Regulation 1978 Standard

Substance Emitted	Sources of Emission	Standards
(a) Acid Gases	Manufacture of Sulphuric Acid	1. Equivalent of : Standard A:7.5 Standard B: 6.0 Standard C: 3.5 gramme of sulphur trioxide/Nm^3 of effluent gas 2. Effluent gas free form persistent mist
(b) Sulphuric Acid mist or sulphur trioxide or both	Any source other than combustion process and plant for manufacture of sulphuric acid as in (a) above	1. Equivalent of : Standard A:0.3 Standard B: 0.25 Standard C: 0.2 gramme of sulphur trioxide/Nm^3 of effluent gas 2. Effluent gas free form persistent mist
(c) Chlorine	Any Source	Standard A:0.3 Standard B: 0.25 Standard C: 0.2 gramme of hydrogen chloride/Nm^3
(d) Hydrogen Chloride	Any source	Standard A:0.6 Standard B: 0.5 Standard C: 0.2 gramme of hydrogen chloride/Nm^3
(e) Fluorine, hydro-fluoric acid or inorganic fluorine compound	Manufacture of aluminum from alumina	Equivalent of : Standard C: 0.02 gramme of , hydrofluoric acid /Nm^3 of effluent gas
(f) Fluorine, hydro-fluoric acid or inorganic fluorine compound	Any source other than aluminum for manufacture of alumina as in (e) above	Equivalent of : Standard A:0.15 Standard B: 0.125 Standard C: 0.100 gramme of hydrofluoric acid /Nm^3 of effluent gas
(g) Hydrogen Sulphide	Any source	Standard A:6.25 Standard B: 5.00 Standard C: 5.00 parts per million volume for volume

(h) Oxide of nitrogen	Manufacture of nitric acid	Equivalent of : Standard A:4.60 Standard B: 4.60 Standard C: 1.7 and effluent gas substantially colorless gramme of sulphur trioxide/ Nm^3
(i) Oxide of nitrogen	Any source other than combustion process and plant for manufacture of nitric acid as in (h) above	Equivalent of : Standard A:3.0 Standard B: 2.5 Standard C: 02.0 gramme of sulphur trioxide /Nm^3

Appendix C: Process Flow Diagram in HYSYS Simulation



Appendix D: Results of HYSYS Simulation

Workbook: Case (Main)

Material Streams

Fluid Pkg: All

Name		Feed	1	2	4	7
Vapour Fraction		0.2568	0.2628	0.2934	0.0318	0.0000
Temperature	(C)	17.70 *	17.27	39.00 *	80.00 *	39.00
Pressure	(kPa)	1251 *	1151 *	1151 *	1151 *	1151
Molar Flow	(kgmole/h)	4645 *	4645	4645	2464	817.9
Mass Flow	(kg/h)	3.030e+005	3.030e+005	3.030e+005	2.456e+005	2.463e+004
Liquid Volume Flow	(m3/h)	462.8	462.8	462.8	354.4	23.29
Heat Flow	(kJ/h)	-9.178e+008	-9.178e+008	-9.012e+008	-4.842e+008	-2.696e+008

Name		3	to HP Flare	to Sump Drum	to LP Flare	Towards Close Drain
Vapour Fraction		0.0000	1.0000	0.0000	1.0000	0.0000
Temperature	(C)	39.00	39.00	128.4	128.4	39.00
Pressure	(kPa)	1151	1151	401.3	401.3	1151
Molar Flow	(kgmole/h)	2464	1363	0.0000	808.9	0.0000
Mass Flow	(kg/h)	2.456e+005	3.278e+004	0.0000	5.049e+004	0.0000
Liquid Volume Flow	(m3/h)	354.4	85.09	0.0000	84.06	0.0000
Heat Flow	(kJ/h)	-5.077e+008	-1.239e+008	0.0000	-9.553e+007	0.0000

Name		to Wet HP Flare	to MEG Regeneration	5	6	8
Vapour Fraction		0.0000	0.0000	0.1436	0.3282	0.0000
Temperature	(C)	39.00	39.00	143.0 *	128.4	128.4
Pressure	(kPa)	1151	1151	1151 *	401.3 *	401.3
Molar Flow	(kgmole/h)	0.0000	817.9	2464	2464	1655
Mass Flow	(kg/h)	0.0000	2.463e+004	2.456e+005	2.456e+005	1.951e+005
Liquid Volume Flow	(m3/h)	0.0000	23.29	354.4	354.4	270.3
Heat Flow	(kJ/h)	0.0000	-2.696e+008	-4.413e+008	-4.413e+008	-3.458e+008

Name		9	10	11	to Storage Tank	to Wet LP Flare
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000
Temperature	(C)	79.10	40.93	40.99	40.99	40.99
Pressure	(kPa)	331.3 *	261.3 *	151.3 *	151.3	151.3
Molar Flow	(kgmole/h)	1655	1655	1655	1655	0.0000
Mass Flow	(kg/h)	1.951e+005	1.951e+005	1.951e+005	1.951e+005	0.0000
Liquid Volume Flow	(m3/h)	270.3	270.3	270.3	270.3	0.0000
Heat Flow	(kJ/h)	-3.693e+008	-3.859e+008	-3.859e+008	-3.859e+008	0.0000

Name		to Off Spec. Tank				
Vapour Fraction		0.0000				
Temperature	(C)	40.99				
Pressure	(kPa)	151.3				
Molar Flow	(kgmole/h)	0.0000				
Mass Flow	(kg/h)	0.0000				
Liquid Volume Flow	(m3/h)	0.0000				
Heat Flow	(kJ/h)	0.0000				

Workbook: Case (Main) (continued)

Compositions

Fluid Pkg: All

Name	Feed	1	2	4	7
Comp Mole Frac (Methane)	0.2180 *	0.2180	0.2180	0.0363	0.0000
Comp Mole Frac (Ethane)	0.0544 *	0.0544	0.0544	0.0323	0.0000
Comp Mole Frac (Propane)	0.0518 *	0.0518	0.0518	0.0581	0.0000
Comp Mole Frac (i-Butane)	0.0189 *	0.0189	0.0189	0.0277	0.0000
Comp Mole Frac (n-Butane)	0.0389 *	0.0389	0.0389	0.0606	0.0000
Comp Mole Frac (n-Pentane)	0.0258 *	0.0258	0.0258	0.0455	0.0000
Comp Mole Frac (i-Pentane)	0.0230 *	0.0230	0.0230	0.0397	0.0000
Comp Mole Frac (Mcyclopentan)	0.0033 *	0.0033	0.0033	0.0061	0.0000
Comp Mole Frac (Benzene)	0.0022 *	0.0022	0.0022	0.0041	0.0000
Comp Mole Frac (n-Hexane)	0.0380 *	0.0380	0.0380	0.0699	0.0000
Comp Mole Frac (Mcyclohexane)	0.0124 *	0.0124	0.0124	0.0231	0.0000
Comp Mole Frac (Cyclohexane)	0.0046 *	0.0046	0.0046	0.0085	0.0000
Comp Mole Frac (Toluene)	0.0038 *	0.0038	0.0038	0.0071	0.0000
Comp Mole Frac (n-Heptane)	0.0467 *	0.0467	0.0467	0.0873	0.0000
Comp Mole Frac (n-Octane)	0.0541 *	0.0541	0.0541	0.1017	0.0000
Comp Mole Frac (p-Xylene)	0.0202 *	0.0202	0.0202	0.0379	0.0000
Comp Mole Frac (n-Nonane)	0.0463 *	0.0463	0.0463	0.0871	0.0000
Comp Mole Frac (Cumene)	0.0054 *	0.0054	0.0054	0.0103	0.0000
Comp Mole Frac (n-Decane)	0.0372 *	0.0372	0.0372	0.0701	0.0000
Comp Mole Frac (n-C11)	0.0878 *	0.0878	0.0878	0.1654	0.0000
Comp Mole Frac (Nitrogen)	0.0026 *	0.0026	0.0026	0.0002	0.0000
Comp Mole Frac (CO2)	0.0120 *	0.0120	0.0120	0.0043	0.0007
Comp Mole Frac (H2S)	0.0102 *	0.0102	0.0102	0.0072	0.0017
Comp Mole Frac (H2O)	0.1292 *	0.1292	0.1292	0.0007	0.7240
Comp Mole Frac (M-Mercaptan)	0.0001 *	0.0001	0.0001	0.0002	0.0000
Comp Mole Frac (E-Mercaptan)	0.0017 *	0.0017	0.0017	0.0028	0.0001
Comp Mole Frac (COS)	0.0000 *	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (nPMercaptan)	0.0015 *	0.0015	0.0015	0.0027	0.0000
Comp Mole Frac (nBMercaptan)	0.0005 *	0.0005	0.0005	0.0009	0.0000
Comp Mole Frac (1Pentanthiol)	0.0011 *	0.0011	0.0011	0.0021	0.0000
Comp Mole Frac (EGlycol)	0.0482 *	0.0482	0.0482	0.0000	0.2734

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	3	to HP Flare	to Sump Drum	to LP Flare	Towards Close Drain
Comp Mole Frac (Methane)	0.0363	0.6776	0.0016	0.1072	0.0000
Comp Mole Frac (Ethane)	0.0323	0.1270	0.0038	0.0906	0.0000
Comp Mole Frac (Propane)	0.0581	0.0715	0.0137	0.1491	0.0000
Comp Mole Frac (i-Butane)	0.0277	0.0144	0.0104	0.0630	0.0000
Comp Mole Frac (n-Butane)	0.0606	0.0231	0.0262	0.1309	0.0000
Comp Mole Frac (n-Pentane)	0.0455	0.0057	0.0312	0.0749	0.0000
Comp Mole Frac (i-Pentane)	0.0397	0.0065	0.0251	0.0696	0.0000
Comp Mole Frac (Mcyclopentan)	0.0061	0.0002	0.0058	0.0065	0.0000
Comp Mole Frac (Benzene)	0.0041	0.0002	0.0040	0.0044	0.0000
Comp Mole Frac (n-Hexane)	0.0699	0.0030	0.0661	0.0778	0.0000
Comp Mole Frac (Mcyclohexane)	0.0231	0.0004	0.0269	0.0154	0.0000
Comp Mole Frac (Cyclohexane)	0.0085	0.0003	0.0087	0.0082	0.0000
Comp Mole Frac (Toluene)	0.0071	0.0001	0.0084	0.0045	0.0000
Comp Mole Frac (n-Heptane)	0.0873	0.0013	0.1007	0.0599	0.0000
Comp Mole Frac (n-Octane)	0.1017	0.0006	0.1318	0.0402	0.0000
Comp Mole Frac (p-Xylene)	0.0379	0.0002	0.0504	0.0125	0.0000
Comp Mole Frac (n-Nonane)	0.0871	0.0002	0.1204	0.0191	0.0000
Comp Mole Frac (Cumene)	0.0103	0.0000	0.0141	0.0023	0.0000
Comp Mole Frac (n-Decane)	0.0701	0.0001	0.1003	0.0084	0.0000
Comp Mole Frac (n-C11)	0.1654	0.0000	0.2411	0.0106	0.0000
Comp Mole Frac (Nitrogen)	0.0002	0.0087	0.0000	0.0005	0.0000
Comp Mole Frac (CO2)	0.0043	0.0328	0.0003	0.0124	0.0007
Comp Mole Frac (H2S)	0.0072	0.0207	0.0010	0.0197	0.0017
Comp Mole Frac (H2O)	0.0007	0.0047	0.0001	0.0020	0.7240
Comp Mole Frac (M-Mercaptan)	0.0002	0.0001	0.0001	0.0004	0.0000
Comp Mole Frac (E-Mercaptan)	0.0028	0.0007	0.0015	0.0054	0.0001
Comp Mole Frac (COS)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (nPMercaptan)	0.0027	0.0001	0.0025	0.0031	0.0000
Comp Mole Frac (nBMercaptan)	0.0009	0.0000	0.0011	0.0007	0.0000
Comp Mole Frac (1Pentanthiol)	0.0021	0.0000	0.0026	0.0009	0.0000
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.2734

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	to Wet HP Flare	to MEG Regeneration	5	6	8
Comp Mole Frac (Methane)	0.0000	0.0000	0.0363	0.0363	0.0016
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0323	0.0323	0.0038
Comp Mole Frac (Propane)	0.0000	0.0000	0.0581	0.0581	0.0137
Comp Mole Frac (i-Butane)	0.0000	0.0000	0.0277	0.0277	0.0104
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0606	0.0606	0.0262
Comp Mole Frac (n-Pentane)	0.0000	0.0000	0.0455	0.0455	0.0312
Comp Mole Frac (i-Pentane)	0.0000	0.0000	0.0397	0.0397	0.0251
Comp Mole Frac (Mcyclopentan)	0.0000	0.0000	0.0061	0.0061	0.0058
Comp Mole Frac (Benzene)	0.0000	0.0000	0.0041	0.0041	0.0040
Comp Mole Frac (n-Hexane)	0.0000	0.0000	0.0699	0.0699	0.0661
Comp Mole Frac (Mcyclohexane)	0.0000	0.0000	0.0231	0.0231	0.0269
Comp Mole Frac (Cyclohexane)	0.0000	0.0000	0.0085	0.0085	0.0087
Comp Mole Frac (Toluene)	0.0000	0.0000	0.0071	0.0071	0.0084
Comp Mole Frac (n-Heptane)	0.0000	0.0000	0.0873	0.0873	0.1007
Comp Mole Frac (n-Octane)	0.0000	0.0000	0.1017	0.1017	0.1318
Comp Mole Frac (p-Xylene)	0.0000	0.0000	0.0379	0.0379	0.0504
Comp Mole Frac (n-Nonane)	0.0000	0.0000	0.0871	0.0871	0.1204
Comp Mole Frac (Cumene)	0.0000	0.0000	0.0103	0.0103	0.0141
Comp Mole Frac (n-Decane)	0.0000	0.0000	0.0701	0.0701	0.1003
Comp Mole Frac (n-C11)	0.0000	0.0000	0.1654	0.1654	0.2411
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0002	0.0002	0.0000
Comp Mole Frac (CO2)	0.0007	0.0007	0.0043	0.0043	0.0003
Comp Mole Frac (H2S)	0.0017	0.0017	0.0072	0.0072	0.0010
Comp Mole Frac (H2O)	0.7240	0.7240	0.0007	0.0007	0.0001
Comp Mole Frac (M-Mercaptan)	0.0000	0.0000	0.0002	0.0002	0.0001
Comp Mole Frac (E-Mercaptan)	0.0001	0.0001	0.0028	0.0028	0.0015
Comp Mole Frac (COS)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (nPMercaptan)	0.0000	0.0000	0.0027	0.0027	0.0025
Comp Mole Frac (nBMercaptan)	0.0000	0.0000	0.0009	0.0009	0.0011
Comp Mole Frac (1Pentanthiol)	0.0000	0.0000	0.0021	0.0021	0.0026
Comp Mole Frac (EGlycol)	0.2734	0.2734	0.0000	0.0000	0.0000

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	9	10	11	to Storage Tank	to Wet LP Flare
Comp Mole Frac (Methane)	0.0016	0.0016	0.0016	0.0016	0.3297
Comp Mole Frac (Ethane)	0.0038	0.0038	0.0038	0.0038	0.1557
Comp Mole Frac (Propane)	0.0137	0.0137	0.0137	0.0137	0.1671
Comp Mole Frac (i-Butane)	0.0104	0.0104	0.0104	0.0104	0.0518
Comp Mole Frac (n-Butane)	0.0262	0.0262	0.0262	0.0262	0.0952
Comp Mole Frac (n-Pentane)	0.0312	0.0312	0.0312	0.0312	0.0360
Comp Mole Frac (i-Pentane)	0.0251	0.0251	0.0251	0.0251	0.0378
Comp Mole Frac (Mcyclopentan)	0.0058	0.0058	0.0058	0.0058	0.0021
Comp Mole Frac (Benzene)	0.0040	0.0040	0.0040	0.0040	0.0015
Comp Mole Frac (n-Hexane)	0.0661	0.0661	0.0661	0.0661	0.0251
Comp Mole Frac (Mcyclohexane)	0.0269	0.0269	0.0269	0.0269	0.0038
Comp Mole Frac (Cyclohexane)	0.0087	0.0087	0.0087	0.0087	0.0026
Comp Mole Frac (Toluene)	0.0084	0.0084	0.0084	0.0084	0.0010
Comp Mole Frac (n-Heptane)	0.1007	0.1007	0.1007	0.1007	0.0131
Comp Mole Frac (n-Octane)	0.1318	0.1318	0.1318	0.1318	0.0059
Comp Mole Frac (p-Xylene)	0.0504	0.0504	0.0504	0.0504	0.0017
Comp Mole Frac (n-Nonane)	0.1204	0.1204	0.1204	0.1204	0.0019
Comp Mole Frac (Cumene)	0.0141	0.0141	0.0141	0.0141	0.0003
Comp Mole Frac (n-Decane)	0.1003	0.1003	0.1003	0.1003	0.0006
Comp Mole Frac (n-C11)	0.2411	0.2411	0.2411	0.2411	0.0005
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0021
Comp Mole Frac (CO2)	0.0003	0.0003	0.0003	0.0003	0.0250
Comp Mole Frac (H2S)	0.0010	0.0010	0.0010	0.0010	0.0303
Comp Mole Frac (H2O)	0.0001	0.0001	0.0001	0.0001	0.0042
Comp Mole Frac (M-Mercaptan)	0.0001	0.0001	0.0001	0.0001	0.0004
Comp Mole Frac (E-Mercaptan)	0.0015	0.0015	0.0015	0.0015	0.0034
Comp Mole Frac (COS)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (nPMercaptan)	0.0025	0.0025	0.0025	0.0025	0.0010
Comp Mole Frac (nBMercaptan)	0.0011	0.0011	0.0011	0.0011	0.0002
Comp Mole Frac (1Pentanthiol)	0.0026	0.0026	0.0026	0.0026	0.0001
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	to Off Spec. Tank				
Comp Mole Frac (Methane)	0.0000				
Comp Mole Frac (Ethane)	0.0000				
Comp Mole Frac (Propane)	0.0000				
Comp Mole Frac (i-Butane)	0.0000				
Comp Mole Frac (n-Butane)	0.0000				
Comp Mole Frac (n-Pentane)	0.0000				
Comp Mole Frac (i-Pentane)	0.0000				
Comp Mole Frac (Mcyclopentan)	0.0000				
Comp Mole Frac (Benzene)	0.0000				
Comp Mole Frac (n-Hexane)	0.0000				
Comp Mole Frac (Mcyclohexane)	0.0000				
Comp Mole Frac (Cyclohexane)	0.0000				
Comp Mole Frac (Toluene)	0.0000				
Comp Mole Frac (n-Heptane)	0.0000				
Comp Mole Frac (n-Octane)	0.0000				
Comp Mole Frac (p-Xylene)	0.0000				
Comp Mole Frac (n-Nonane)	0.0000				
Comp Mole Frac (Cumene)	0.0000				
Comp Mole Frac (n-Decane)	0.0000				
Comp Mole Frac (n-C11)	0.0000				
Comp Mole Frac (Nitrogen)	0.0000				
Comp Mole Frac (CO2)	0.0008				
Comp Mole Frac (H2S)	0.0040				
Comp Mole Frac (H2O)	0.5085				
Comp Mole Frac (M-Mercaptan)	0.0002				
Comp Mole Frac (E-Mercaptan)	0.0017				
Comp Mole Frac (COS)	0.0000				
Comp Mole Frac (nPMercaptan)	0.0006				
Comp Mole Frac (nBMercaptan)	0.0001				
Comp Mole Frac (1Pentanthiol)	0.0000				
Comp Mole Frac (EGlycol)	0.4840				

Energy Streams

Fluid Pkg: All

Name		Q-101				
Heat Flow	(kJ/h)	4.293e+007				

Unit Ops

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
Valve 1	Valve	Feed	1	No	500.0 *
Valve 2	Valve	5	6	No	500.0 *
Valve 3	Valve	10	11	No	500.0 *
HP Steam Heater	Heater	4 Q-101	5	No	500.0 *
Pre-flash Drum	3 Phase Separator	2	3 to HP Flare 7	No	500.0 *
Flash Drum	3 Phase Separator	6	8 to LP Flare to Sump Drum	No	500.0 *
Degassing Drum	3 Phase Separator	11	to Storage Tank to Wet LP Flare	No	500.0 *

Workbook: Case (Main) (continued)

Unit Ops (continued)

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
Degassing Drum	3 Phase Separator		to Off Spec. Tank	No	500.0 *
Filter	Simple Solid Separator	7	Towards Close Drain	No	500.0 *
			to Wet HP Flare		
			to MEG Regeneration		
Pre-flash Exchanger	Heat Exchanger	1	2	No	500.0 *
		9	10		
Heat Exchanger	Heat Exchanger	3	4	No	500.0 *
		8	9		

Process Simulation of a Back-up Condensate Stabilization Unit

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Abstract- **Condensate stabilization is a process of increasing the amount of heavy components in the condensate. Thus, in this study, it aims to to investigate Reid Vapor Pressures (RVP) values in a back-up Condensate Stabilization Unit (CSU) with a given feed of condensate and obtaining the best actual operating parameter for each of equipment. On the basis specified target for stabilized in this unit, two properties of product should stabilize before storing in storage tanks and export which for RVP of maximum 10 psia for summer season and 12 psia for winter season. Based on study, it shows that back-up CSU uses only flash vaporization which only heating and flashing processes, In back-up CSU, salt and sulfur content are not affect the process as there are no any distillation in column and it operate only for shut-down plant as well as not a continuous process. Results show that CSU's RVP and sulfur content is 7.932 psia and 2408.52 ppm which is the optimum condition for the process.**

Keywords: RVP, sulfur content, salt, CSU, flash vaporization

I. INTRODUCTION

Nowadays, the stable and sweet product of condensate and also the gasoline produced by modern plant processes must meet established pipeline and marketing standards. So, stabilization of condensate refers to the stripping of the light ends content (C1 and C2) from the raw liquids and the removal of all acidic constituents to produce a suitable product for the market.

The stabilization operations are simple and the principles are similar to the LPG fractionation systems. In general, condensate stabilization accomplishes several goals, the foremost of which are:

a) To increase the recovery of methane-ethane and LPG products.
b) To lower the vapor pressure of the condensate, which making it more suitable for blending and reducing the evaporation losses while the product is in storage or during shipment.
c) To sweeten the raw liquids entering the plant by removing the acid gases like hydrogen sulphide and carbon dioxide contents, in order to meet the required specifications.
d) To maintain the purity and molecular weight of the lean absorption oil, free of certain components like pentanes and heavier hydrocarbons.

II. LITERATURE REVIEW

A. Background of CSU

Hydrocarbon condensate recovered from the natural gas may be not transferred for further processing but they will be stabilized first in order to bending with crude oil stream and then sold as crude oil. For the case of raw condensate, there are no any specific requirements for the product other than the process specification. So, the process of increasing the amount of intermediates (C3 to C5) and heavy (C6+) components in the condensate is called "condensate stabilization" **[1]**. Hence, the hydrocarbon condensate stabilization is required to minimize the hydrocarbon losses from the storage tank **[2]**. This process is needed to be done because a vapour phase will not produce upon flashing to the atmospheric storage tank in order to reduce the vapor pressure of the condensate liquid. Besides that, the purpose of this process is to separate light hydrocarbon gases such as methane and ethane from heavier hydrocarbon components such as propane and others. Heavier components can be used for oil refinery cracking processes which allow the production of light production such as liquefied petroleum gas (LPG) and gasoline **[3]**. Nevertheless, stabilized liquid has vapor pressure specifications as; the product will be transferred into pipelines which have limitation of pressure **[1]**.

In order to measure the vapor pressure of the condensate is by measuring the Reid Vapor Pressure (RVP). It is a way to measure how quickly fuels evaporate; it's often used in determining gasoline and other petroleum product blends **[4]**. It means that higher RVP of a fuel, the more it quickly evaporates indicating the loss of the product. RVP represents the fuel's evaporation at 100 degrees Fahrenheit (37.8 degrees Celsius), and is measured in pounds per square inch, or PSIs **[2]**. Hence, the property that RVP measures often is referred to as the gasoline's volatility. RVP can be estimated without performing the actual test by using algorithm **[5]**.

B. Natural Gas Processing

Figure 1 shows the flow of condensate to be stabilized before transferring to the storage tank which is starting from the natural gas well.

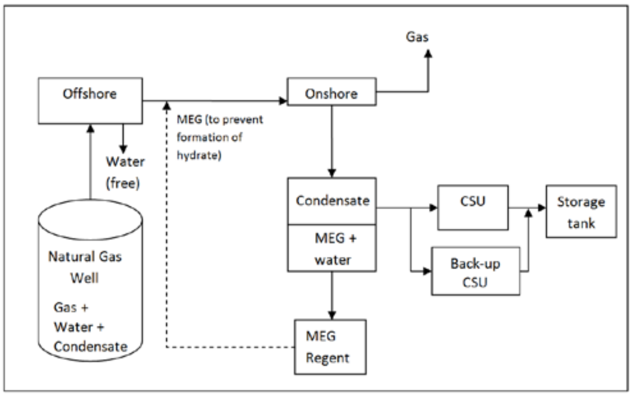

Figure 1: Flow Diagram of Condensate Stabilization [12]

Firstly, at natural gas well, a mixture of natural gas which consists of gas, water and condensate will be extracted before it is being transferred to the offshore plant (oil rig). Then, some water will be removed out from the mixture and transported to the onshore plant. The transportation of the treated gas will be done through a pipeline about 120km from offshore plant to onshore plant. As the result, the gas mixture will dehydrate and form a blockage which the flow of gas will not go smoothly. Hence, monoethylene glycol (MEG) is channeled to the pipeline in order to prevent the formation of gas hydration.

Once gas mixture reaching in onshore plant, it will be separated into two stream; gas stream and liquid stream. The gas stream will be transferred to gas plant and the liquid stream that consists of condensate, MEG and water is further separated which form a condensate stream and mixture of MEG and water stream. The mixture MEG and water will be treated in MEG regeneration unit which MEG will be recycled to the pipeline. Then condensate stream will send to the condensate stabilization unit (CSU) with a back-up unit to run the plant during failure. After treated in CSU, the condensate will be stored in the storage tanks.

C. Condensate Stabilization

1. Flash Vaporization

Stabilization by flash vaporization is a simple operation employing only two or three flash tanks **[1]**. This process is similar to stage separation utilizing the equilibrium principles between vapor and condensate phases. Equilibrium vaporization occurs when the vapor and condensate phases are in equilibrium at the temperature and pressure of separation **[1]**.

Figure 2 shows the typical flash vaporization process for the condensate stabilization. Based on the Figure 1, the main feed which is condensate coming from the inlet separator is passing through a heat exchanger entering the high-pressure flash tank where the pressure is maintained at 600 psia. A pressure drop which costly 300 psia helps the flashing of large amounts of lighter ends which they will be discharge to sour vapor stream after recompression. The discharged ones can be sent to the further units or recycled into the reservoir. After that, the bottom liquid from the high-pressure tank will enter the middle pressure flash tank where the additional methane and ethane will be released. Then, the bottom the product will enter again to the low-pressure tank and they will enter the condensate stripper for the purification before sending to the storage tank.

To ensure efficient separation, condensate is degassed in the stripper vessel at the lowest possible pressure prior to storage **[1]**. This reduces excess flashing of condensate in the storage tank and reduces the inert gas blanket pressure required in it. Multistage flashing is based on the principle of progressively lowing the pressure of condensate during each stage **[5]**. This will enhance for the flashing of lighter components from the condensate.

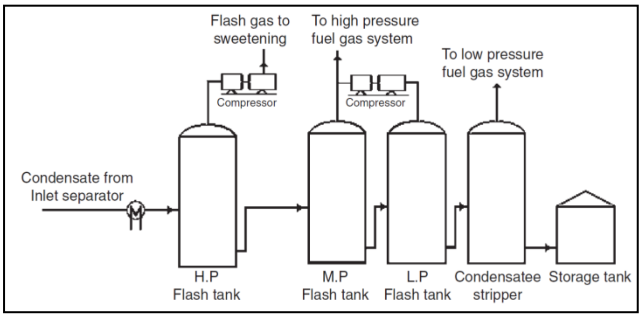

Figure 2: Flash Vaporization Method **[1]**

For the back-up unit, it is found that it just use only simple heating and cooling process as we want to reduce the cost as well as the it is not in continuous process. Hence, back-up unit prefers to use flash vaporization method to run its operation. This method just uses only some pressure to stabilize the condensate before sending to the storage tank.

D. Impact of Salt on Back-up CSU

Apart from crude oil in the mixture of condensate, there are also presence of salty, acidic water and solid particulate which cause various problems in the stabilization plant. Separation of water phase from the condensate can be problematic as many fields or plants from various regions experience it. Although the condensate viscosity is very low and the difference of density with water is high, other impurities tent to create stable condensate/water emulsion that are difficult to separate efficiently **[6]**.

There are many consequences on the impact of impurities in the condensate stabilization plant. Many plants in worldwide have reported that several following consequences may arise due to water carry over that contains dissolved salt like:

a) Plant upsets and stability of the plant is reduced.
b) Quality issues of the final products for example gasoline and LPG.
c) Excessive corrosion and deposits inside the stabilizer and re-boiler.
d) Power consumption is increased due to the ingression of excessive levels of water and loss of heat transfer caused by the contaminants.

e) There will be frequent shutdown of the stabilization train for the cleaning purposes, causing a drop in production and hence loss of revenue if the flow rate cannot be compensated by the other stabilization trains.

f) The corrosion products will be created in the export condensate storage tank and in the export pipeline also referred to as ‘black powder’.

In order to separate impurities from the condensate, we need to have desalter/dehydrators in the plant. Mostly, desalter is electrostatic precipitators and utilizes new technologies which are three grid-grid electrode system and horizontal emulsion distribution for better separation performance **[7]**. This equipment should be installed in the stabilization plant. However, in back-up condensate stabilization unit, desalter is not included as this unit brings to the operation only after the main CSU shutdown. This is save a lot of money as well as it can get more profit from selling the product. Besides that, we can see the effect of the impurities on the stabilizer which the distillation column and the impurities affect the reboiler performance. As the result, we cannot get desired product and the desalter should be installed in the main condensate stabilization unit.

E. Impact of Sulfur Concentration on Final Product

Elemental sulfur is a powerful oxidant. It means that the strong oxidizing property In the oil and gas industry, sulfur is recognized as aggressive corrosion accelerators, particularly for pitting and other forms of localized corrosion **[8]**. Normally, sulfur is formed in sour oil and gas systems from some of the following mechanisms; differential solubility of sulfur in high pressure sous gas, destabilization of hydrogen polysulfide presents in sour gas an others. If there is more than 2.5% sulfur present in crude, they are called sour crude **[9]**.

According to McConomy curve, measure corrosion rates of carbon, low allow, and stainless steels are significantly high where significant concentrations of Mercaptans which containing sulfur element are present in crude oils and hydrocarbon condensate **[10]**. It suggested that sulfur element that containing hydrocarbon condensate cause higher corrosion rate than sulfur species in general. Besides that, there more species of Mercaptans in the condensate, the higher corrosion rate will occur. Thus, we need to concern about presence of Mercaptans in the final product in the back-up condensate stabilization unit.

In addition to that, Mercaptans will also give bad smell on the condensate. This will affect the quality of condensate before selling to the customer. Nevertheless, Mercaptans are added to odorless natural gas for safety reason which in normal operations, gas companies add it to deliver to the city gas stations and commercial usage **[11]**. This is because Mercaptans will prevent the potential underground water contamination which natural gas will be not in good condition.

III. METHODOLOGY

A.Feed of Process

In analyzing back-up CSU system performance, plant simulation is modeled first by using HYSYS (ver. 2006). It is essential to have a model that reliable in representing CSU system as some of the data is unavailable from the plant and only available from the estimation from HYSYS model. To achieve this objective, the plant simulation is using the actual operating value, gained from data available in real plant South Pars gas field (Assaluyeh, Iran).

Based on Figure 3, the envelope curve show that the feed consist of 0.57 liquid phase, 0.26 vapor phase and 0.18 aqueous phase.

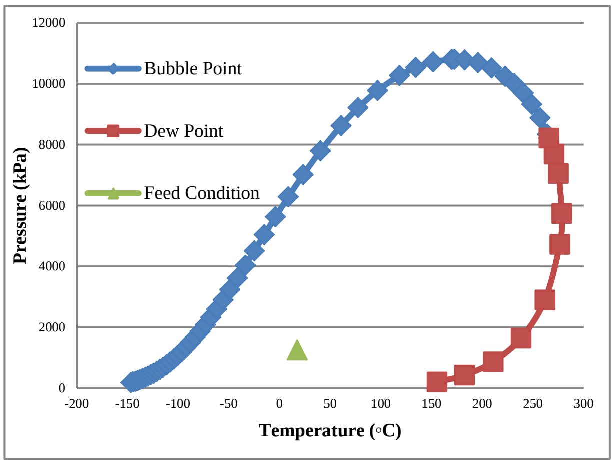

Figure 3: Envelope Curve of the Feed

Table 1: Feed stream conditions and composition

Properties	
Total	
Normal Flow, kmol/hr	4645
Normal Flow, Kg/hr	325604
Pressure, barg	11.5
Temperature, °C	17.7
Molecular Weight	70.1
Heat Flow, kW	4009
Molecular Weight	70.1
Vapor	
Molar Flow, MMSCFD	24
Normal Flow, kg/h	25957
Density, kg/cu m @ P, T	11.7
Liquid	
Standard Liq Vol Flow, SBPD	61349
Normal Flow, kg/h	299647
Actual cu m/h @ P , T	389
S. G. Liquid @ P, T	0.770
Composition	
Components	Mole Fraction
Methane	0.218041
Ethane	0.054396
Propane	0.051802
i-Butane	0.018891
n-Butane	0.038908
i-Pentane	0.022982
n-Pentane	0.025847
Mcyclopentane	0.003284
Benzene	0.002242
n-Hexane	0.037976
Cyclohexane	0.004601
Mcyclohexane	0.012375
Toluene	0.003805
n-Heptane	0.046731

n-Octane	0.054126
p-Xylene	0.020163
n-Nonane	0.046275
Cumene	0.005448
n-Decane	0.037223
C11+	0.087779
Nitrogen	0.002623
Carbon Dioxide	0.012015
Hydrogen Sulphide	0.010165
Water	0.129249
M-Mercaptan	0.000130
E-Mercaptan	0.001688
COS	0.000007
nPMercaptan	0.001478
nBMercaptan	0.000505
1Pentanthiol	0.001092
MEG	0.048154

B. Plant Simulation

The model is constructed based on reference CSU plant operation. Based on Figure 4, the purpose of this process is separation of aqueous phase and gaseous hydrocarbon from the condensate and then to stabilize it for the export by adjusting Reid Vapor Pressure (RVP) which indicating the volatility of the condensate. This is because the quality of the product depends on composition and also RVP before selling to the customers.

Firstly, main feed from the onshore plant is entered to pre-flash drum to remove light hydrocarbons, most value of acid gases and lighter paraffin's will be excited in this step. Next, condensate temperature is increased in two sequential heat exchanger and High Pressure (HP) heater up to 80°C and 143°C respectively. Lastly, this fluid with crossing from of two first shell tube exchanger and degassing in the last flash drum is stored in storage tanks.

The off-gas for example light hydrocarbon like methane, ethane and propane, sulfur components like hydrogen sulfide, and others will be burnt in the appropriate flare system. For aqueous phase like MEG and others are sent to further processing in the suitable units for instance MEG regeneration unit. Besides that, components that have sulfur element like Mercaptans and also water will be sent to off specification tank and then will be transferred to the waste treatment.

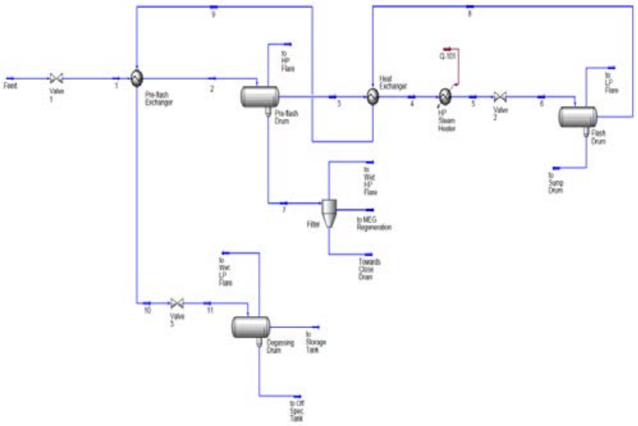

Figure 4: Process Flow Diagram in HYSYS Simulation

C. Adjusting Process Parameter

There are a few of process parameters that will change the final product specification including changing steam temperature, steam pressure, feed flow rate, feed temperature and feed pressure. These parameters are simulated in one dimensional condition where others parameter is kept constant at a time. The product specification as they will be analyzed by Reid Vapor Pressure (RVP), and sulfur content

Table 1: Condition of parameters for the study of effects of changing parameters

Parameter	**Study of the effect of:**				
	Steam Temperature	Steam Pressure	Feed Flow Rate	Feed Temperature	Feed Pressure
Steam Temperature	No	Constant	Constant	Constant	Constant
Steam Pressure	Constant	No	Constant	Constant	Constant
Feed Flow Rate	Constant	Constant	No	Constant	Constant
Feed Temperature	Constant	Constant	Constant	No	Constant
Feed Pressure	Constant	Constant	Constant	Constant	No

IV. RESULTS AND DISCUSSION

A. Comparison of Actual Plant Data, Pro II Software, HYSYS of Condensate Composition of Final Product at Normal Condition

For validation of data of final product, the obtained data have been compared with actual plant data, Pro/II software version 7.1 and HYSYS Software (ver. 2006).

Based on the Figure 5, it can be seen that the simulation of the process is exactly follow the trend as the plant data about 5% differences. Hence, this HYSYS data will be validated to the real plant. Besides that, the Pro/II Software looks also follows the trend of data with the real plant. Overall; data of final product should be valid for simulation software in order to validate the result for this process.

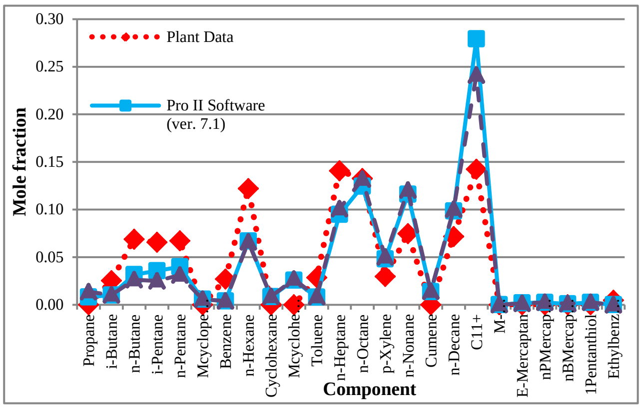

Figure 2: *Comparison of Actual Plant Data, Pro II Software, HYSYS of Condensate Composition of Final Product at Normal Condition*

Figure 6 shows the comparison of plant data and HYSYS data of final product of this process. It can be seen that the data is nearly the same of the comparison between plant data and Pro/II Software data. Light hydrocarbon components in HYSYS shows that their mole fraction is lower than the plant data which indicates the unwanted hydrocarbon is already flashed before sending to the storage tank. This will increase the quality of the product.

Furthermore, heavy hydrocarbon in the final product of HYSYS data shows that it is the nearly the same with the plant data. Although the plant data is slightly higher, we can consider that the quality of the product is the same as the plant data because their differences are not affecting the overall data.

Besides, sulfur element which contains in M-mercaptan, n-Pmercaptans also same with the plant which are very in small quantities. This shows that our final product should be safe to send to the customer and also the profit should be increased.

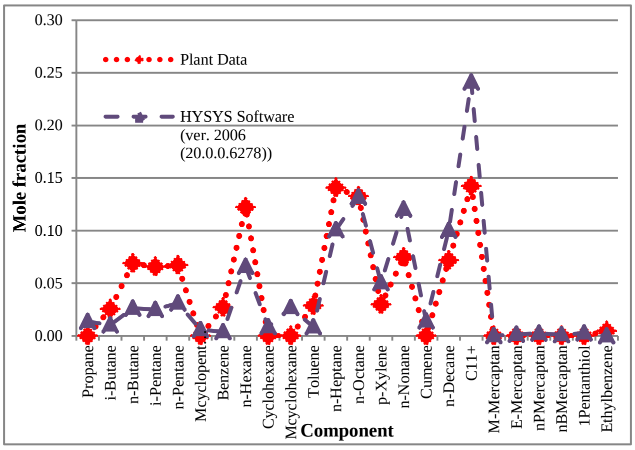

Figure 3: Comparison of Actual Plant Data and HYSYS Data

C. Effect of Changing Process Parameter

I. Effect of Steam Temperature

Figure 7 shows that the higher temperature gives lower RVP value. This means that higher temperature will remove more acid gases and light hydrocarbon which RVP changing between 8.385 psia and 6.336 psia. From this range, the best temperature for this process is to avoid more loss of Propane and Butane as well as stripping corrosive and sour components to promote value of the product is 143◦C which causes RVP of 7.932 psia.

For sulfur content, it shows that the concentration of sulfur decrease as temperature of steam increase. This is because the components which contain sulfur elemenet will be removed rapidly as at higher temperature and it will flash the acidic component. The highest sulfur concentration is 2500 ppm which is very high at low temperature and should be removed in this stage.

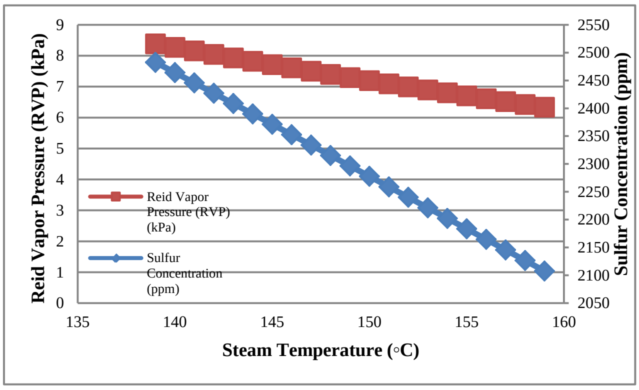

Figure 7: Effect of Steam Temperature on the RVP and Sulfur Concentration of the product

II. Effect of Steam Pressure

Figure 8 shows that RVP is decreasing as steam pressure is increasing. The lowest pressure is 10 kPa and the highest pressure is 65 kPa as lower pressure and higher pressure in this range will give temperature cross in the heat exchanger which is not valid for this process. From this range of the steam pressure, it will cause the RVP changes from 7.942 to 7.921 psia. The optimum condition is 35 kPa to remove the unwanted hydrocarbon and also stripping sour component which causes RVP of 7.932 psia. It means that higher steam pressure will increase the steam heat duty. As the result of higher steam heat duty, there are more flashing of acidic gases.

For sulfur content, it shows that the sulfur concentration is decreasing as steam pressure is increasing. From this trend, it can be seen that the high temperature will remove the components which contain sulfur element faster in the in separator. The highest sulfur concentration is 2410 ppm and needed to reduce as low as possible by increasing the steam pressure.

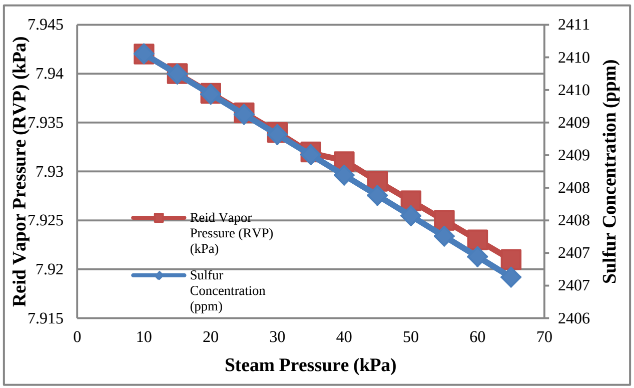

Figure 8: Effect of Steam Pressure on the RVP and Sulfur Concentration of the product

III. Effect of Feed FlowRate

Figure 9 shows that RVP is increasing as feed flow rate is increasing**.** This is because the highest feed which needs to be separated in the separators, the higher heat is required for a while the heat has been kept constant. As the result, RVP will increase insufficient heat to maintain the operation of the separator. From this trend, at 1848 kmole/hr which is 40% from the original one, the plant will turn down as there will be a temperature cross in the heat exchanger. Furthermore, at 5574 kmole/hr (120%), the fee will be overflowed because temperature cross also occurred in the heat exchanger. Therefore, the optimum condition for feed flow rate is ranging from 50% to 110%.

For sulfur content, it can be seen that sulfur concentration is increasing as feed flow rate is increasing. This is because there are more feed into the process, result in more sulfur feed to the plant which no adjustment of heat is being carried out; here more sulfur in the product. Hence, to decrease the sulfur concentration in final product, the feed flow rate should be low or heat supply should be adjusted. The lowest of sulfur concentration is 1494 ppm at 50% of feed flow rate and the highest of sulfur concentration is 2502 ppm at 110% of feed flow rate.

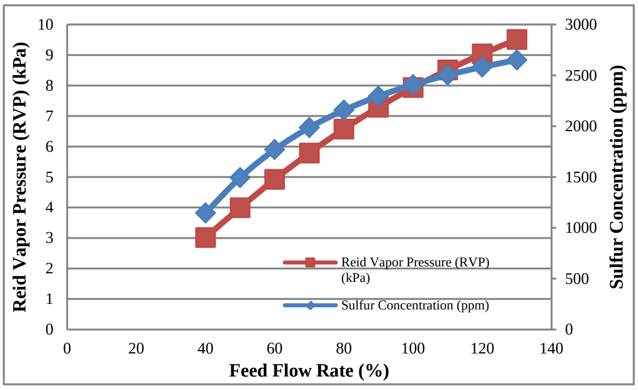

Figure 9: Effect of Feed Flow Rate on the RVP and Sulfur Concentration of the product

IV. Effect of Feed Temperature

Figure 10 shows that RVP is decreasing as feed temperature is increasing. This result shows that we want to have lower RVP which we want to recover the product and can be sold at larger quantities. From the summer case which at 10 psia the range of feed temperature should be -10◦C till 20◦C and the original feed temperature is 17.7◦C which causes 7.932 psia. However, 30◦C or higher of feed temperature will cause temperature cross in the heat exchanger and the best condition for the process is 10-20◦C.

For sulfur content, it can be seen that sulfur concentration is decreasing as feed temperature is increasing. This is because the unwanted components including containing sulfur element have been removed at higher temperature. Therefore, the feed temperature should be as high as possible until it does not go against the temperature difference in the heat exchanger which is 20◦C. The lowest sulfur concentration is 2375 ppm and the highest of sulfur concentration is 4002 ppm.

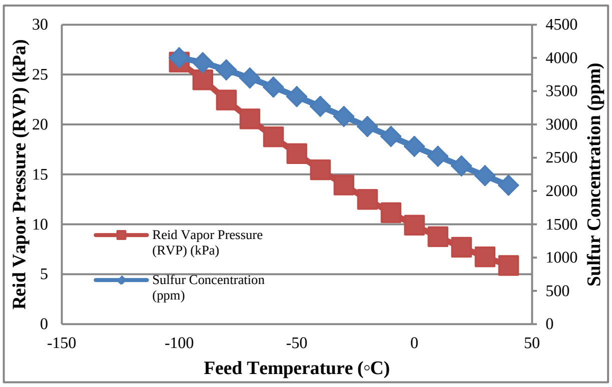

Figure 10: Effect of Feed Temperature on the RVP and Sulfur Concentration of the product

V. Effect of Feed Pressure

Figure 11 shows that RVP is increasing as feed pressure is increasing. This because higher pressure feed will cause the feed to become in liquid phase which in the 3-phase separator should be the pressure as low as possible to flash off the acidic gases. The lowest pressure is 1200 kPa as below from that, there will be temperature cross and it is the lowest pressure that can be used in the process in range of 1200-1300 kPa, is causing the RVP changes from 6.9 psia to 8.9 psia and the best optimum condition is 1251 kPa which causes RVP 7.932 psia. This shows that feed pressure is one of the factor that will affect the process especially RVP.

For sulfur content, it can be seen that the sulfur concentration is increasing as feed pressure is increasing. From this trend, higher temperature will cause the components that contain sulfur element will not be removed in the separator. The lowest sulfur concentration is 2281 ppm at 1200 kPa and feed pressure should be low as possible until it does not beyond the temperature cross in the heat exchanger.

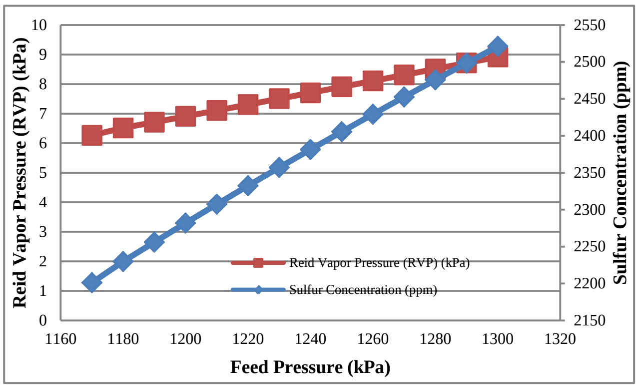

Figure 11: Effect of Feed Pressure on the RVP and Sulfur Concentration of the product

V. CONCLUSION

This project is carried out based on two main objectives, which are simulating a back-up Condensate Stabilization Unit (CSU) that is able to bring down the Reid Vapor Pressure (RVP) of the Summer Rich Condensate of maximum 10 psia for summer season and 12 psia for winter season and finding the best operating parameters for each of the equipment in a back-up condensate stabilization unit.

For validation data of this project, the data have been compared with the actual plant in Iran and also Pro/II Software. From the comparison, the results show the composition from each of data is nearly and very feasible to build in Malaysia. Although there are some data is deviated from actual data plant a little bit, it does not concern with the simulation.

This research shows steam temperature, steam pressure, feed flow rate, feed temperature and feed pressure are important parameters to adjust the amount of RVP as well as sulfur concentration. It has been found that for steam HP heater temperature which the most optimum condition is 143◦C that gives RVP 7.932 psia which is below than 10 psia in summer season. Hence, it is very essential for these parameters to be monitored closely which unfavorable content should be in specified range and to ensure that they will not exceed the limit that affect the overall quality of final product.

Besides that, in the literature review, there are some studies about the sulfur content and salt which can affect the back-up CSU in term of equipment and also final product. From the studies, it shows that they give slight effect which the salt is affecting the column reboilers which there are no any column in the back-up CSU and sulfur needs to be treated for more in order to produce the quality product.

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REFERENCES

[1] Mokhatab, S., Poe, W.A. and Speight, J.G. 2006, *Handbook of Natural Gas Transmission and Processing*, UK, Gulf Professional Publishing, Page 281.

[2] Benoy, J. and Kale, R.N. 2010, “Condensate Stabilization” in *Offshore World*, India; Chemtech Foundation.

[3] James H. Gary and Glenn E. Handwerk (2001). *Petroleum Refining: Technology and Economics* (4th ed.). CRC Press. ISBN 0-8247-0482-7.

[4]Discovery Channel. 20 June 2012 <http://curiosity.discovery.com/question/what-is-reid-vapor-pressure>.

[5] J. Javier Vazquez-Esparragoza, Gustavo Arturo Iglesias-Silva, Michael W. Hlavinka,Jerry A. Bullin, “How To Estimate Reid Vapor Pressure (RVP) of Blends”, USA, Bryan Research & Engineering, Inc.

[6] Trifilieff, O., Wines, Thomas H. and Daire, F. 2011, *Improve Hydrocarbon Condensate Dehydration Performance Diagnostics and Solutions,* UK, Pall Corporation.

[7] Cameron Inc., 2010, *NATCO Electro-Dynamic Desalter,* USA, Cameron Inc.

[8] Bojes, J., Lerbscher, J., and Wamburi Wellingtion. 2010, *Elemental Sulfur in 3-Phase Sour Gas Systems – Is Condensate Really Your Ally?.* USA, Baker Hughes Inc.

[9] Oild Industry Jargon De-coded, 1 November 2012 <www.naturalhub.com/slweb/defin_oil_and_Natural_Gas.html>

[10] De Jong, Jan-Peter., Dowling N., Sargent M., 2007, *Effect Of Mercaptans And Other Organic Sulfur Species On Hightemperature Corrosion In Crude And Condensate Distillation Units,* USA, NACE International.

[11] Elenbaas D., Mohamed D., Khan M. Hidayati D., Huang F., and Lee R., 1996, *The Environmental Impact of Mercaptans in Underground Storage,* USA, Society of Petroleum Engineers, Inc.

[12] Dr. Nejat R., L. Chemical Engineering Department, University Technology PETRONAS, Perak. Personal Interview. Sept. 7. 2012.