Process Simulation of Condensate Stabilisation Unit

by

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Dissertation submitted in partial fulfilment of
the requirements for the
Bachelor of Engineering (Hons)
(Chemical Engineering)

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CERTIFICATION OF APPROVAL

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A project dissertation submitted to the
Chemical Engineering Programme
Universiti Teknologi PETRONAS
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(CHEMICAL ENGINEERING)

Approved by,

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(DR. NEJATOLLAH 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.

(LAILA SAKINAH BINTI JUSOH)

ABSTRACT

Natural gas condensate is a liquid phase that is present in natural gas wells. From these wells, condensates are sent to oil refineries for it to be processed into marketable petroleum products. However, before being transported to the refineries, the condensate need to be stabilized beforehand since it contains light components that could flash off in low pressure conditions. This paper aims to find the suitable conditions to stabilise a feed of “Summer Rich” condensate to maximum Reid Vapour Pressures (RVPs) of 10 psia and 12 psia for summer and winter conditions respectively. A simulation of the process has been conducted by using Aspen HYSYS (ver. 2006) software. It was found that at a column pressure of 8.5 barg and reboiler temperature of 180°C, the condensate is successfully stabilised to an RVP of 8.778 psia. The effects of four parameters, i.e. feed flow rate, feed temperature, feed pressure and reboiler temperature towards the product RVP and sulphur content are also studied. The increase in feed flow rate causes an increase in the product RVP and sulphur content. On the other hand, increasing feed temperature and reboiler temperature causes the product RVP and sulphur content to decrease. The increase in feed pressure initially causes the product RVP to increase but after reaching a certain point, the RVP starts to decrease. Besides that, the increase in feed pressure does not give a major impact on the sulphur content.

ACKNOWLEDGEMENTS

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ABBREVIATIONS AND NOMENCLATURES

API	American Petroleum Institute
ASTM	American Standard Testing Method
BCSU	Back-up Condensate Stabilisation Unit
CSU	Condensate Stabilisation Unit
DOE	Design of Experiments
LLE	Liquid-liquid Equilibria
MEG	Monoethylene Glycol
PFD	Process Flow Diagram
ppm	Parts per Million
PTB	Pounds per Thousand Barrels
RVP	Reid Vapour Pressure
VLE	Vapour-liquid Equilibria

CHAPTER 1

1.0 INTRODUCTION

1.1 Project Background

Natural gas condensate (also called as condensate, gas condensate or natural gasoline) is a liquid hydrocarbon phase that is present as gaseous component in the raw natural gas produced from many natural gas fields. Based on the Schlumberger Oilfield Glossary, this mixture of hydrocarbon liquids has a low density and a high API gravity and will condense out of the raw gas if the temperature is reduced to below the hydrocarbon dew point temperature of the raw gas.

There are three types of well where natural gas could be found and they are crude oil well, dry gas well and condensate well. A crude oil well is a well that contains both crude oil and natural gas. Natural gas from this kind of well is called ‘associated gas’ and it can be present either separate from the crude oil or dissolved in the crude oil. On the other hand, a dry gas well produces only raw natural gas and does not contain any hydrocarbon liquid. Gas from this well is called ‘non-associated’ gas. Lastly, the condensate wells contains raw natural gas together with natural gas liquid and gas from this well is also called as ‘non-associated gas’ and sometimes referred to as ‘wet gas’.

Condensate from these well contains a large amount of light components that would flash off at low pressure and high temperature state. This condition is not ideal for condensate storage and transportation. Therefore, condensate stabilization needs to be done prior to its further processing.

1.2 Problem Statement

According to Schlumberger Oilfield Glossary, the hydrocarbon condensate is mainly composed of propane, butane, pentane and heavier hydrocarbons. However, there is also a percentage of lighter components present in the mixture. When brought to a condition with lower pressure, these lighter components will flash off. This condition is unsafe for storage and transportation and will also cause losses when the hydrocarbon evaporates into the atmosphere. Therefore, to avoid flashing in storage tanks and during transportation, the condensate needs to be stabilized beforehand.

For that, a Condensate Stabilization Unit needs to be developed. Stabilization is actually a process that lowers down the vapour pressure of the condensate. This ensures that at lower pressure, the condensate will not flash off. The required vapour pressure is usually denoted in terms of Reid Vapour Pressure (RVP) and varies according to usage and customer demands.

1.3 Objective and Scope of Study

The main aim of this project is to simulate a Condensate Stabilization Unit with a given composition of “Summer Rich” feed to obtain a stabilized condensate with Reid Vapour Pressures (RVPs) of 10 psia for summer and 12 psia for winter. In order to accomplish that main aim, the following objectives need to be achieved.

The objectives of this project are:

- To study the process of condensate stabilization, why it is required and what technologies are currently being used.
- To develop a model of Condensate Stabilization Unit (CSU) and simulate it using ASPEN HYSYS software.
- To study the effects of the changes in feed conditions and reboiler temperature towards the product properties.

CHAPTER 2

2.0 LITERATURE REVIEW

2.1 Natural-Gas Processing

The diagram below shows the simple block flow diagram of natural gas processing starting from the natural gas well and to the onshore processing plant.

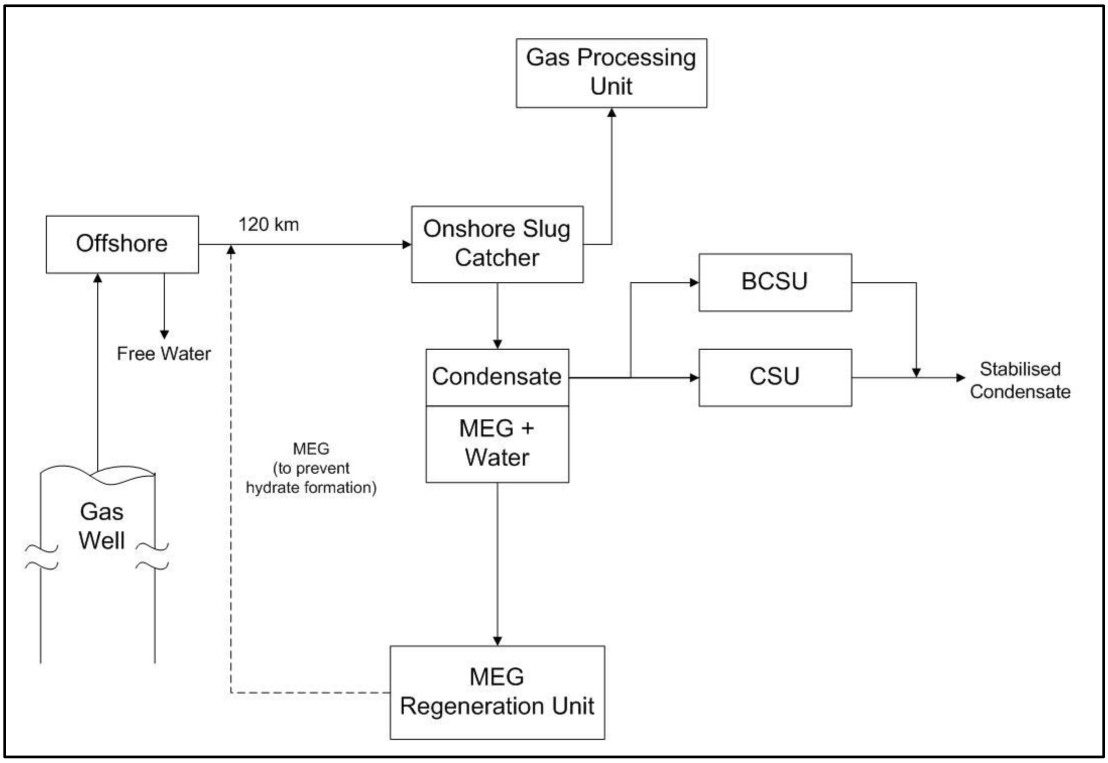


Figure 1: Flow Diagram of Natural-Gas Processing

Firstly, a mixture of gas, water and condensate will be extracted from the well and free water will be removed from that mixture in the offshore plant. The remaining mixture would then be transported through a pipeline to the onshore plant to be further processed. Without any injection of desiccant, the gas mixture would form hydrates in the pipeline and that would in turn form a blockage that could restrict the flow of gas. Therefore, monoethylene glycol (MEG) is injected in the pipeline in order to inhibit the formation of hydrates.

Upon reaching onshore, the mixture would first be separated into two streams; a gas stream and a liquid stream. The gas stream is sent to the gas plants to be further processed. On the other hand, the liquid stream which comprises of condensate, MEG and water is further separated to a stream of condensate and a stream of MEG and water.

The mixture of MEG and water is treated in the MEG regeneration unit where the MEG would be regenerated and then reused in the pipeline whereas the condensate would be sent to the condensate stabilization unit (CSU). This is where the stabilization process takes place.

2.2 Condensate Stabilization

Mokhatab et al. (2006) stated that there are two main methods for the stabilization of condensate. They are flash vaporization and fractionation.

2.2.1 Flash Vaporization

The method of flash vaporization utilizes the equilibrium principles between the vapour and liquid phases. The vapour phase of the condensate is flashed off by gradually lowering the pressure of the condensate during each stage (Benoy and Kale, 2010). The liquid mixture is partially vaporized (Geankoplis, 2003) and then equilibrium between the vapour and liquid would be reached when the two phases are in equilibrium at the temperature and pressure of separation.

Figure 2 and Figure 3 show the process flow of condensate stabilization through Ash Vaporization (Mokhatab et al., 2006) and Two Stage Flashing (Benoy and Kale, 2010). Both of these methods fall under the Flash Vaporization technique. As can be seen in the figures, the process of flash vaporization would usually comprise of two or three flash tanks. The number of tanks depends on how many stages of flashing are required to achieve the desired Reid Vapour Pressure (RVP).

Generally, in a flash vaporization process, the condensate would enter the first separator through the inlet separator (the inlet separator removes water from the condensate). This first separator is the one with the highest pressure. Here, a large amount of the lighter components are flashed off and released through the top of the separator. The bottom liquid from this separator is then sent to the next separator where the operating pressure is lower than the previous one. This condition enables more light components to flash off from the condensate. This is repeated until the last separator. For the process in Figure 2, the condensate is degassed in a stripper vessel prior to storage in order to increase the efficiency of separation. Meanwhile, the process in Figure 3 sends the stabilized condensate to a heat exchanger to recover the heat while heating up the feed. It is then cooled before being stored.

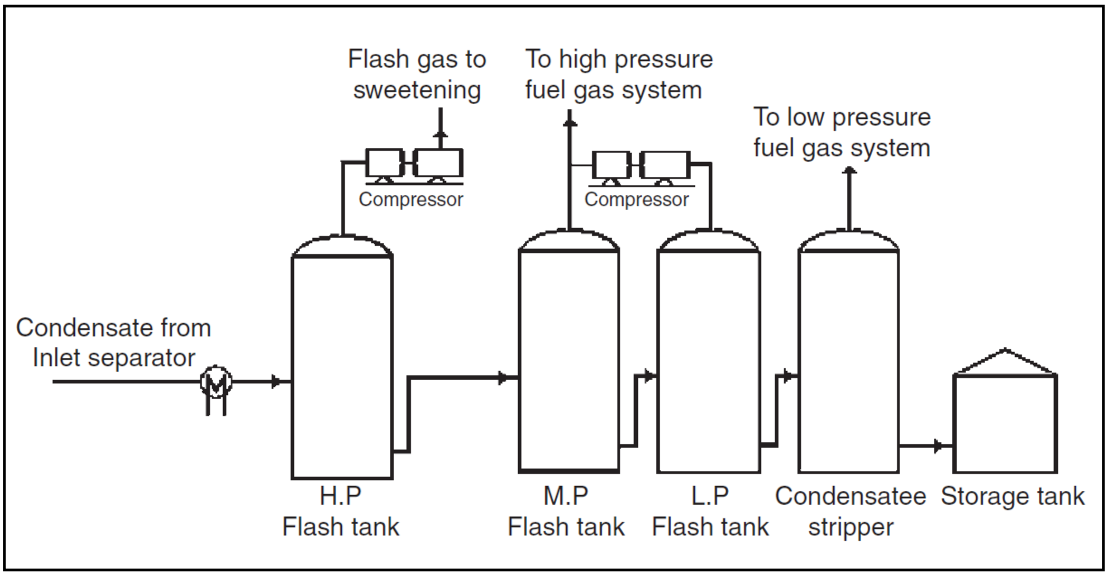


Figure 2: Schematic of Condensate Stabilization through Ash Vaporization Process (Mokhatab et al., 2006)

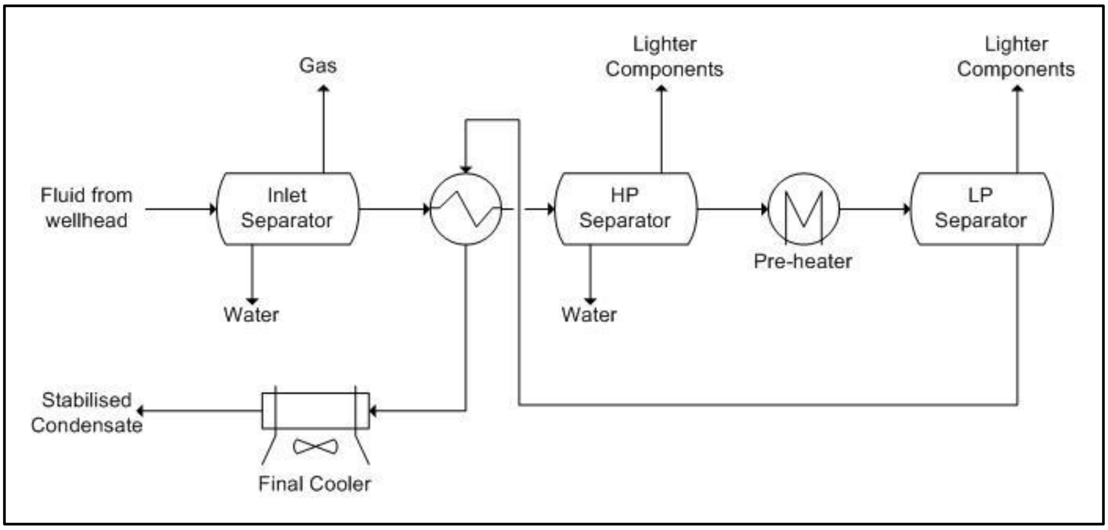


Figure 3: Schematic of Condensate Stabilization through Two-Stage Flashing (Benoy and Kale, 2010)

The method of stabilization through flash vaporization is actually an old technology and is not used in a modern gas plant. However, it can be used as a Back-up Condensate Stabilization Unit (BCSU) in the event of a shutdown of the main CSU. Figure 4 shows an example of a BCSU in Iran’s Phases 6, 7&8 Gas Refinery (Esmaeili, 2010).

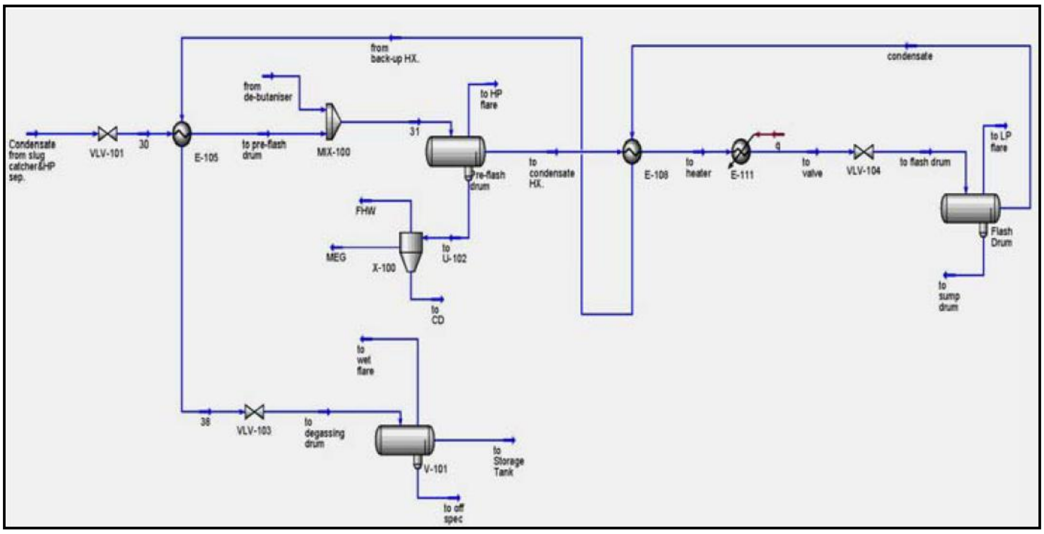
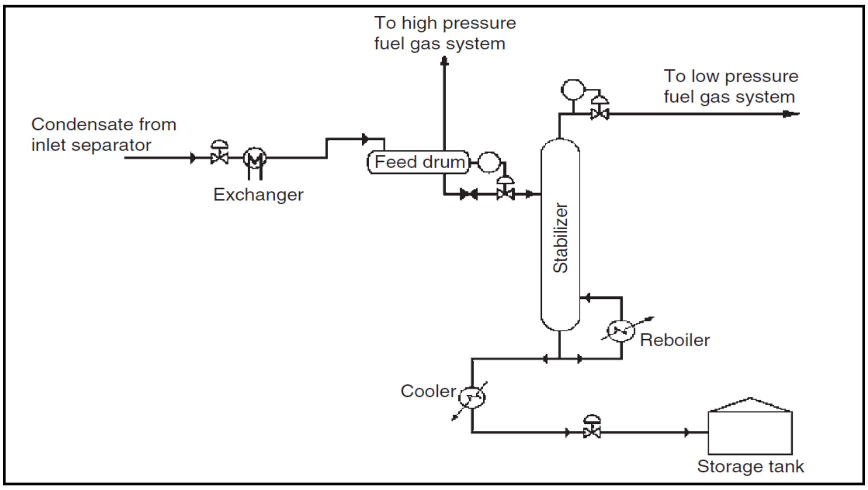


Figure 4: Process Flow Diagram of Back-up Condensate Stabilization Unit in Phases 6, 7 & 8 Gas Refinery (Esmaeili, 2010)

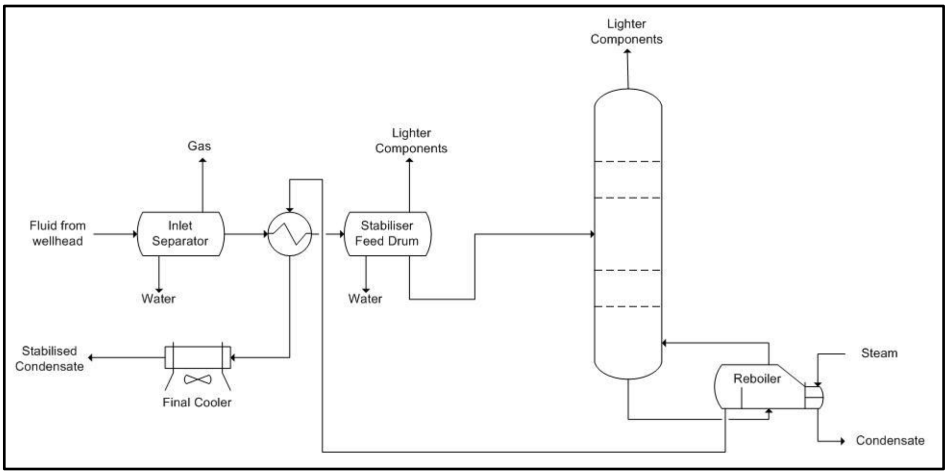
2.2.2 Stabilization by Fractionation

The second method of condensate stabilization is stabilization by fractionation. This method is very popular in the industry and is precise enough to produce liquids of suitable vapour pressure (Mokhatab et al., 2006). In this process, light fractions are removed from the condensate so the finished product would be composed of the heavy fractions which are mainly pentanes and other heavier hydrocarbons. Thus, the bottom product obtained is a vapour free liquid that can be safely stored at atmospheric pressure. This technique of stabilization is a more modern technique compared to the flash vaporization method and it is more economically attractive.

Figures 5(a) (Mokhatab et al., 2006) and 5(b) (Benoy and Kale, 2010) show two examples of process flow of condensate stabilization through fractionation. In these processes, the feed would first enter the feed drum through the inlet separator. The inlet separator here has the same function as in flash vaporization where it removes entrained water from the condensate. In the feed drum, any gaseous components would be separated from the feed and sent to the fuel gas system. The condensate then enters the stabilizer on or near the top tray. This column basically acts as a reboiled absorber where the light components are removed from the condensate (Mokhatab et al., 2006).



5(a): Schematic of a Condensate Stabilization System (Mokhatab et al., 2006)



5(b): Schematic of Stabilization by Non-Refluxed Stabilizer (Benoy and Kale, 2010)

Figure 5: Examples of Condensate Stabilization through Fractionation

For a better separation, a refluxed distillation tower could be used. The process flow of refluxed distillation stabilization can be seen in Figure 6 (Benoy and Kale, 2010). As can be seen in the figure, the early part of the process is similar to the process in stabilization through fractionation.

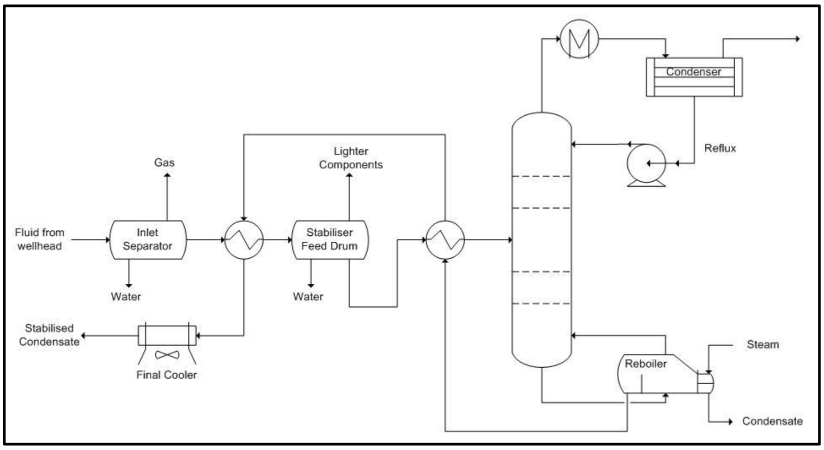


Figure 6: Schematic of Condensate Stabilization through Refluxed Distillation (Benoy and Kale, 2010)

The difference is in the location where the feed enters the column and also the existence of the reflux section in the refluxed distillation. Instead of being fed to the top part of the column, the feed in this process is fed at the tray where the feed temperature is the same as the tray temperature (Benoy and Kale, 2010). A refluxed stabilizer can recover more intermediate products from the stabilizer overhead vapour compared to non-refluxed stabilizer. However, the extent of liquid recovery varies from case to case basis (Benoy and Kale, 2010). Furthermore, a refluxed stabilization requires more capital cost as it requires more equipment.

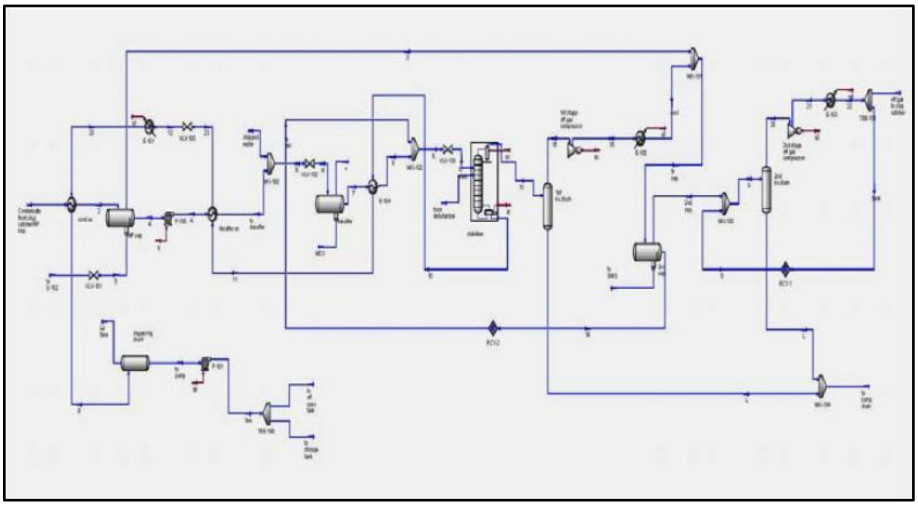


Figure 7: Process of Simulated Condensate Stabilization Plant in Phases 6, 7 & 8 Gas Refinery (Esmaeili, 2010)

Figure 7 shows the process of simulated condensate stabilisation plant in Phases 6, 7 & 8 in South pars gas field, Iran. This condensate stabilisation plant uses the fractionation method with reflux. In his paper, Esmaeili (2010) found that the most suitable operating conditions for the stabiliser column is at a pressure of 8.6 barg and reboiler temperature of 170°C. At these conditions, the resulting product has an RVP that is neither too low so as to lose more lighter components nor is the water content too high (Esmaeili, 2010).

2.3 Design Considerations of Stabilization Column

According to Mokhatab et al. (2006), in most cases, condensate stabilization columns operate as a non-refluxed tower. This type of column is much simpler in its operation as it requires no external cooling sources which makes in applicable in remote locations. However, it is less efficient in terms of separation. On the other hand, a column with reflux can recover more intermediate components from the gas but it requires more capital cost as it consists of more equipments.

Mokhatab et al. (2006) stated that the stabilization tower pressure in the column depends on the amount of liquid to be stabilized and whether it’s sweet or sour. If the feed is sweet, the pressure is to be set as high as possible to minimize the overhead vapour recompression. This is because this vapour will be mixed with the separator vapour. This will also help to decrease the cost of reflux cooling. However, relative volatility decreases with pressure and driving H_2S overhead requires a relatively low pressure. Therefore, the pressure needs to be at optimum for best operations.

Figure 8 shows the maximum recommended feed temperature to cold-feed stabilizer. Campbell (1992) stated that this curve can be used as a convenient guide to estimating the operating range of a non-refluxed stabilizer. However, when the feed temperature-pressure relationship exceeds the one shown in Figure 8, the non-refluxed tower should not be used. Though there could be an exception where small quantities of liquid are being handled and first cost is very critical.

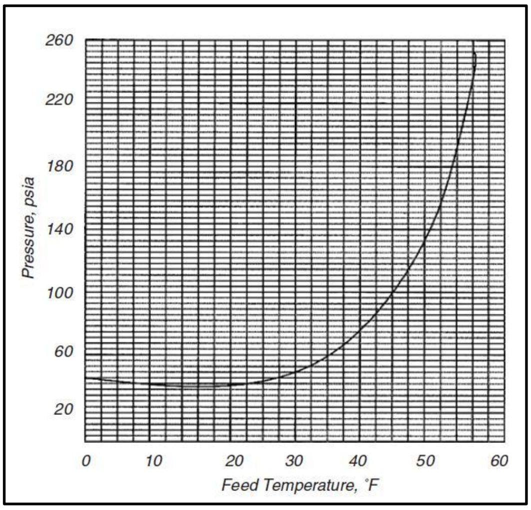


Figure 8: Maximum recommended feed temperature to cold-feed temperature (Campbell, 1992)

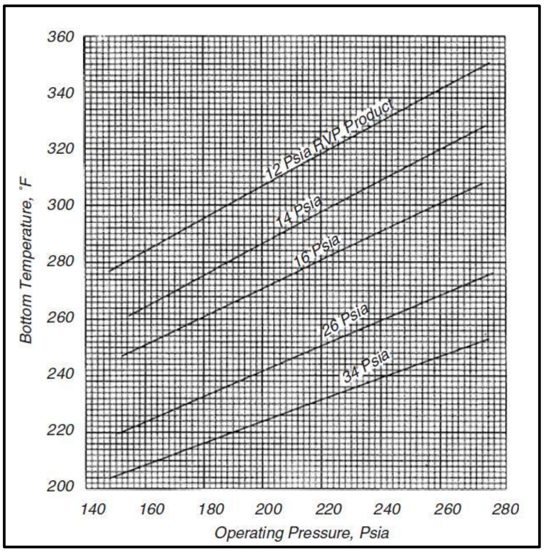


Figure 9: Estimation of proper bottom temperature of a non-refluxed stabilizer (Campbell, 1992)

Campbell (1992) also stated that Figure 9 can be used to estimate the bottom (reboiler) temperature for producing a specified Reid vapour pressure product. The figure shows the bottom reboiler temperature as a function of operating pressure and each line represents the RVP product desired.

Once the pressure and operating temperatures for the tower has been chosen, the split in the tower should be predicted. The most convenient method for this is by using the pseudo-equilibrium constant (K) values for each component between the top and the bottom of the tower. By using this concept and a simple flash calculation, the separation across the non-refluxed tower stabilizer can be estimated. From the flash calculation, the vapour will be the composition of the overhead product and the liquid will be the composition of the liquid product.

As stated earlier, the main purpose of condensate stabilization is to remove the lighter components and this is done by removing most of the C4 contained in the condensate. Therefore, for the split calculation, split for nC4 is assumed (Mokhatab et al., 2006). For this assumption, the mole fraction of the component in the liquid can be estimated by:

$$X_i = \frac{L_i}{\sum_{i=1}^{n} L_i}$$

Where,
X_i = mole fraction of component i in liquid
n = no. of components in bottom liquid
L_i = no. of moles of component i in liquid

The number of moles of component i in liquid (L_i) can be calculated by:

$$L_i = \frac{F_i(nC4\ split)}{RV_i}$$

Where,
nC4 split = assumed ratio of nC4 in bottom and nC4 feed
RV_i = relative volatility of component i

In this split method, the vapour pressure of the components is the main property being used. It is assumed that the mole fraction of each component multiplied by its vapour pressure represents the contribution of that component to the total mixture vapour pressure. The total mixture vapour pressure can be calculated by using the following equation:

$$P_v = \sum_{i=1}^{n} \left(P_{v_i} \times X_i\right)$$

Where,
P_v = vapour pressure of mixture
P_{vi} = vapour pressure of component i
X_i = mole fraction of component i

If the resulting vapour pressure of mixture is higher than the desired RVP, then a lower number should be chosen for the nC4 split. On the other hand, if the resulting vapour pressure of mixture is lower than the desired RVP, then a higher number should be chosen for the nC4 split.

2.4 Salt Content in Condensate

Besides ensuring that the product specification is within the accepted limit, it is also important to ensure that the equipments in the unit is maintained in a good condition so as to ensure that all operation can run smoothly. One of the causes that can lead to equipment failure or inefficiency is the presence of salt in the condensate.

Wauquier (2000) stated that in most cases, oil produced from wells would contain some degree of salt; the majority of it being sodium chloride. There are also variable amounts of alkaline earth salt associated with the oil. The structural position of the well and the physical characteristics of the reservoir rock will affect the salt content in the oil.

2.4.1 Types and Characteristics of Salts

The salts present in oil are mainly sodium chloride ($NaCl$), magnesium chloride ($MgCl_2$) and calcium chloride ($CaCl_2$). These salts could be in the solid phase (in form of crystal) or liquid phase (ionised in water in the oil). Theoretically, all the ionised salts can be removed by simple settling. However, this method would require a large storage capacity. On the other hand, the crystal salts can be removed by washing it with water where the crystals would be ionised and then hydrated. These hydrated salts are more advantageous because they have a high solubility in water. Therefore, it is shown that it is important to add water in the process of desalting (Wauquier, 2000).

Wauquier (2000) also mentioned that besides salt, there are also other contaminants present in oil. They are mineral sediments, rust and also iron sulphide and these contaminants can generate stable emulsions that are hard to break.

2.4.2 Implication of Presence of Salt in Processing Unit

According to Wauquier (2000), the presence of salt would cause drawbacks on the processing unit. Firstly, the salt content would cause fouling in preheating exchangers. Wauquier (2000) stated that once the concentration of salt exceeds 40 ppm, there would be a change in the fouling resistance of the exchangers and deposits would lay in the tube bundle more rapidly. This condition would increase the heat resistivity in the exchanger and thus increase utility requirement for the heating purposes. Secondly, Wauquier (2000) stated that "the alkaline earth chlorides are hydrolyzed with generation of hydrogen chloride, which causes corrosion in the overhead equipment of topping units."

The accepted limit of salt content in the overhead water is 10 ppm. If the concentration exceeds this value, severe corrosion can occur in the processing unit (Wauquier, 2000).

2.5 Removal of Salt

According to Kleintits et al. (2003), in the Northern German gas reservoirs, it has been observed that the precipitation of salt from the reservoir water is increasing as recovery progresses. As a result of this halite scale, the production rate of the well is significantly decreasing. This could lead to the complete blockage of the flow paths and would eventually cause the abandonment of wells. This phenomenon can also occur at process units where the precipitation of the salt would cause blockage in the piping and equipments. Also, as mentioned in the previous section, the presence of salt would also cause corrosion and affect the efficiency of the equipments and processing. Therefore, in order to avoid these implications, it is important to remove the salt from the oil before processing it.

Vafajoo et al. (2012) stated that there are currently several different methods available for the removal of salt. They are chemical demulsification, gravity or centrifugal settling, pH adjustment, filtration, heat treatment, membrane separation and electrostatic demulsification. However, some of these methods have their disadvantages. The centrifugation method has a high operating cost, the pH adjustment is not effective for breaking the water-oil emulsion, and heat treatment is expensive as it requires high fuel consumption (Vafajoo et al., 2012). Therefore, the method that is most widely used for desalting is the electrostatic demulsification.

2.5.1 Desalting Mechanism

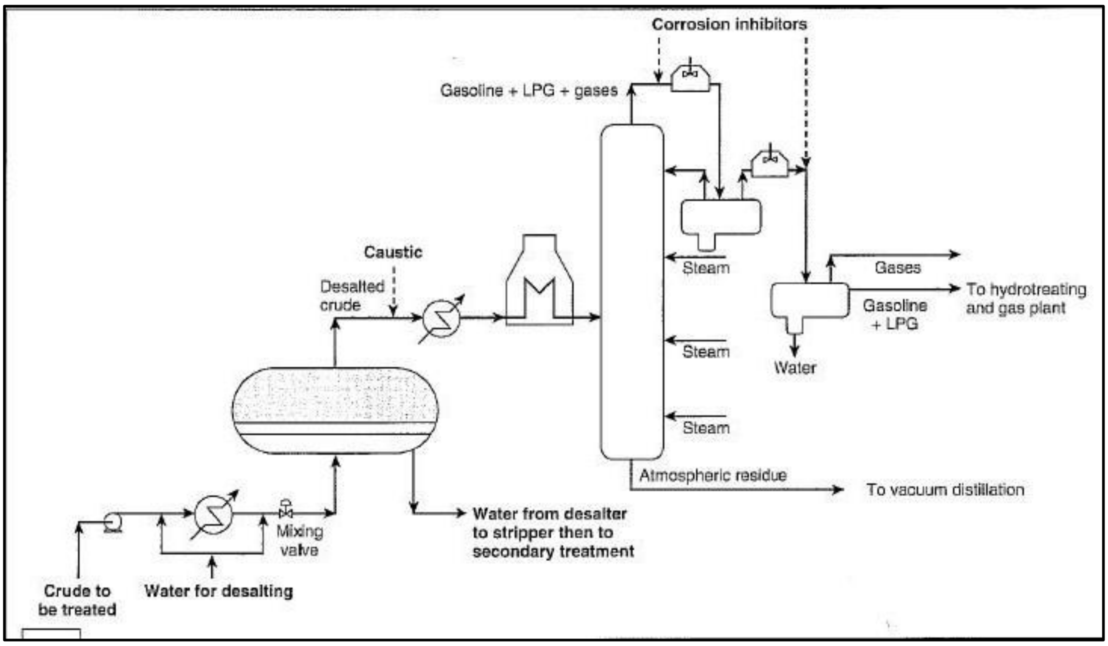


Figure 10: Process Flow of Crude Desalting (Wauquier, 2000)

The figure above shows the basic process flow of a crude desalting unit. In summary, in crude desalting, the crude is mixed with wash water to dissolve the salts and impurities, and then the wash water is separated from the crude by electrocoalescence. In essence, there are three consecutive steps in desalting (Waqauier, 2000) which are the diffusion of salts, coalescence of water droplets and settling.

2.5.1.1 Diffusion of Salts

The main purpose of this step is to dissolve the salt contained in the crude into the wash water. In order to ensure that all salt crystals are affected by this diffusion, the water-crude emulsion must be fine (Figure 11). For this purpose, the wash water is partly injected at the discharge of the water pump and partly at the mixing valve (Waqauier, 2000).

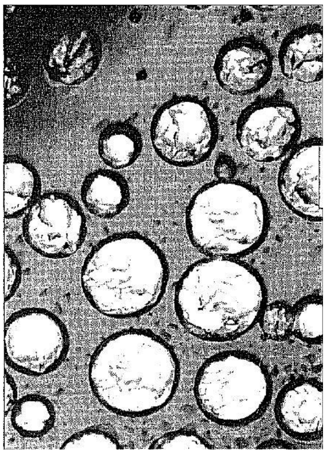


Figure 11: Water-crude Emulsion (Waqauier, 2000)

2.5.1.2 Coalescence

The water-crude emulsion is a thorough mixture of two non-miscible liquids. The crude is the continuous phase whereas the water is the dispersed phase. Other contaminants such as asphaltenes and finely dissolved solids act as agents that stabilise the emulsion. This is caused by their adsorption on the water oil interface and this forming a film (Waqauier, 2000).

In order to separate the water-oil emulsion, an electrostatic field is used to induce coalescence. In operating the desalter, attention must be given to the level of the water. If too much water reaches the grid, it will cause them to electrically overload and eventually lead to shutdown.

To assist in the separation of the oil and water, a chemical (emulsion breaking chemical/demulsifier) is injected

2.5.1.3 Settling

This step is the gravity settling of water in the vessel. The gravitational settling is governed by the Stoke’s law (Mahdi et al., 2008):

$$v = \frac{2\pi r^2 \Delta\rho g}{9\mu}$$

2.6 HYSYS Simulation Software

HYSYS is a process modelling tool that can be used for conceptual design, optimization and performance monitoring for oil & gas production, gas processing, petroleum refining, and air separation industries. In order to use this simulation software, there are several fundamental concepts that need to be understood.

2.6.1 Selection of Thermodynamics Package

In the design of any process, the thermodynamic data is often required, especially the phase equilibria. It is vital for the importance of thermodynamics to be appreciated since it is often that more than 40% of the cost in many processes is related to the separation units (Prausnitz et al., 1999). Table 1 below summarizes the type of data needed in the design of various separation processes.

Table 1: Phase equilibria data needed in the design of specific unit operations (Kontogeorgis and Folas, 2010)

Unit Operation	Phase Equilibria Type
Distillation *Azeotropic distillation* *Extractive distillation* *Evaporation, drying*	Vapour-liquid equilibria (VLE) VLE, liquid-liquid equilibria (LLE) LLE Gas-liquid equilibria
Absorption *Reboiled absorption* *Stripping*	VLE Gas-liquid equilibria Gas-liquid equilibria
Extraction *Supercritical fluid extraction*	LLE Gas-liquid and solid-gas equilibria
Adsorption	Vapour-solid equilibria Liquid-solid equilibria
Crystallization	Liquid-solid (vapour) equilibria
Leaching	Liquid-solid equilibria
Bioseparations *Extraction with aqueous two-phase systems* *Liquid-liquid extraction with reverse micelles*	LLE

In process simulators such as HYSYS a wide selection of thermodynamics is to be chosen from. For the selection of the most suitable thermodynamic package, the so-called ‘decision or selection trees’ is used (Kontogeorgis and Folas, 2010). Figure 12 below shows an example of the steps in the process of choosing the most suitable thermodynamic package where the type of compound involved is taken into account. In Figure 12, the thermodynamic model is determined by the characteristics of the compounds involved in the process. The characteristics that are taken into consideration in the selection are the polarity, real or pseudo-components, electrolyte and also the pressure.

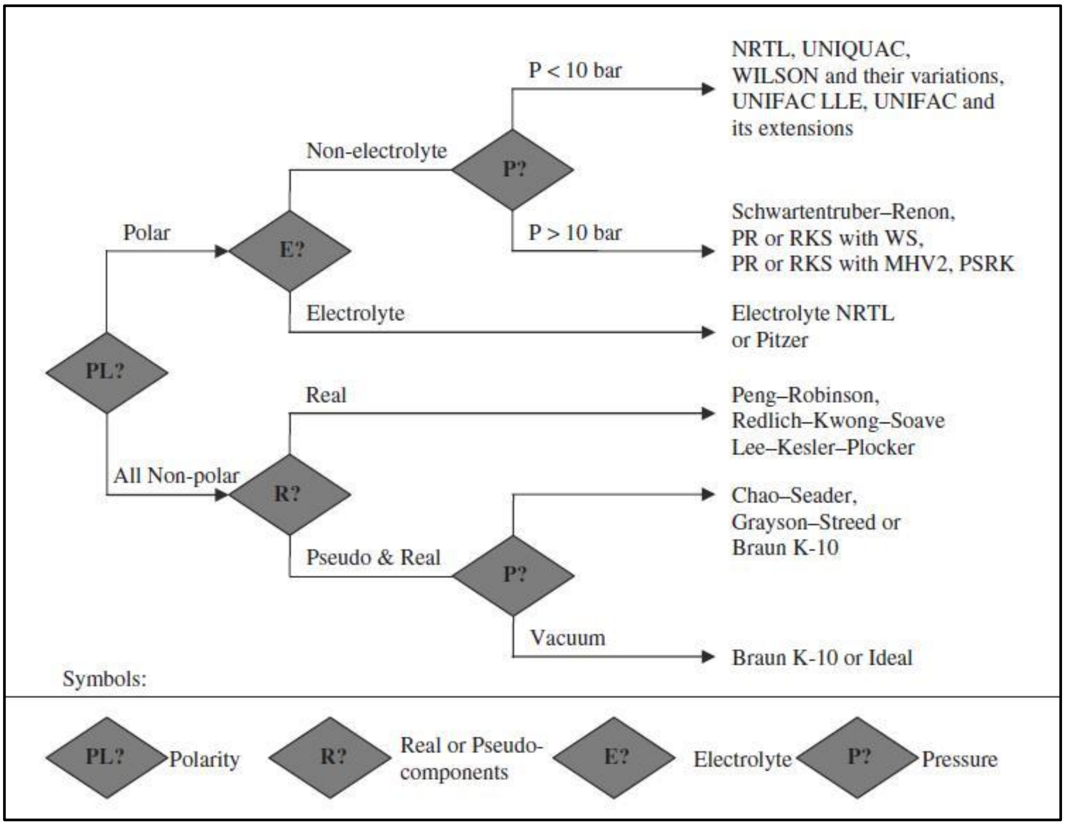


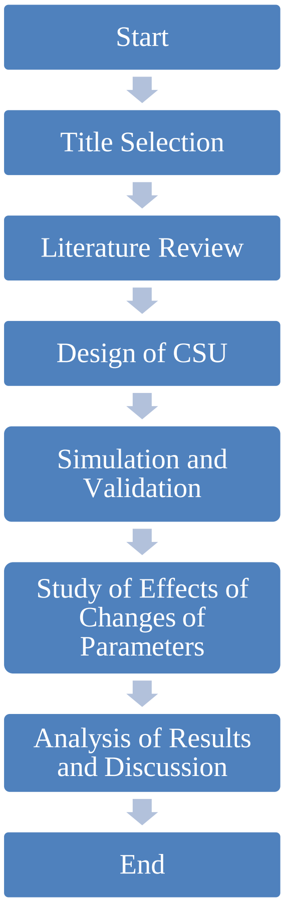
Figure 12: Example of selection tree for selecting the appropriate thermodynamic model (Kontogeorgis and Folas, 2010)

CHAPTER 3

3.0 METHODOLOGY

3.1 Research Methodology

The chart below shows the general flow of this project from the beginning till the end.



For this project, the methodology will be divided into two parts which are Project Research and Project Simulation.

3.1.1 Project Research

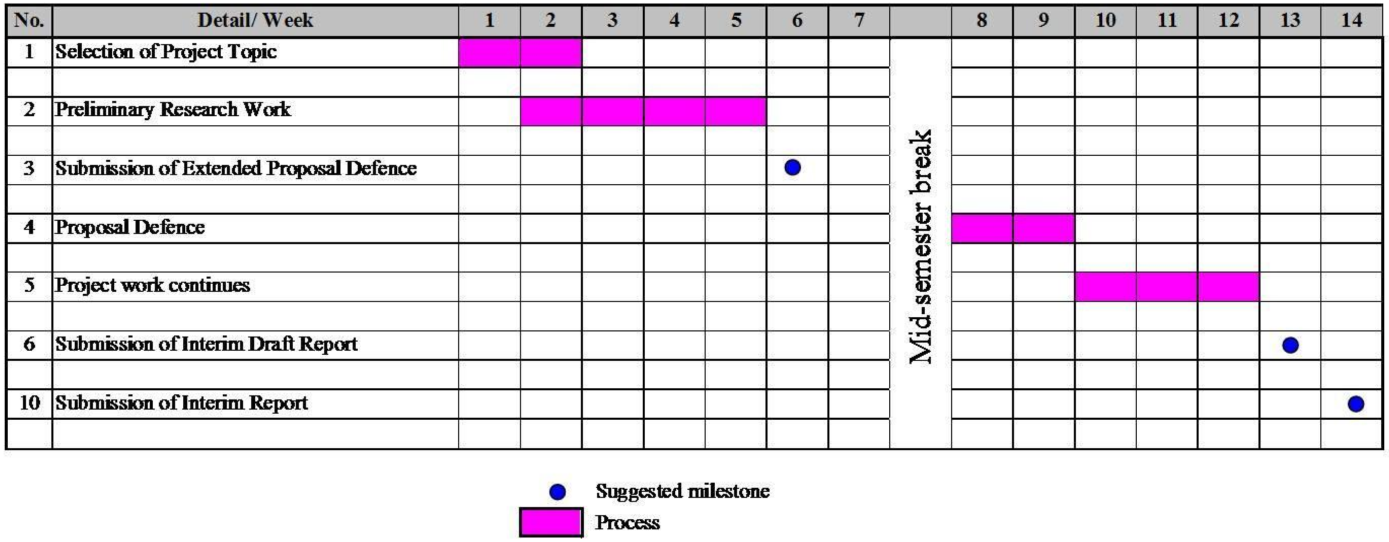
In this part of the project, research on the topic of condensate stabilization is conducted by reading books, journals and articles concerning the subject matter. Besides research on condensate stabilization itself, a brief background research is also done on the basics of natural gas processing. From this research, it can be known why the condensate stabilization is required and what are the current technologies being used.

3.1.2 Project Simulation

Once the first part of the project is completed and the basic process flow diagram as well as the estimates of parameters is obtained, the project simulation would be started by using the ASPEN HYSYS software. In this part, the process flow diagram is generated using the software and the parameters inside the process is adjusted little by little to obtain the desired result.

3.2 Key Milestone and Gantt Chart

The figure below shows the timeline of how the project has been conducted during the first semester of FYP.



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

Figure 13: FYP I Gantt Chart

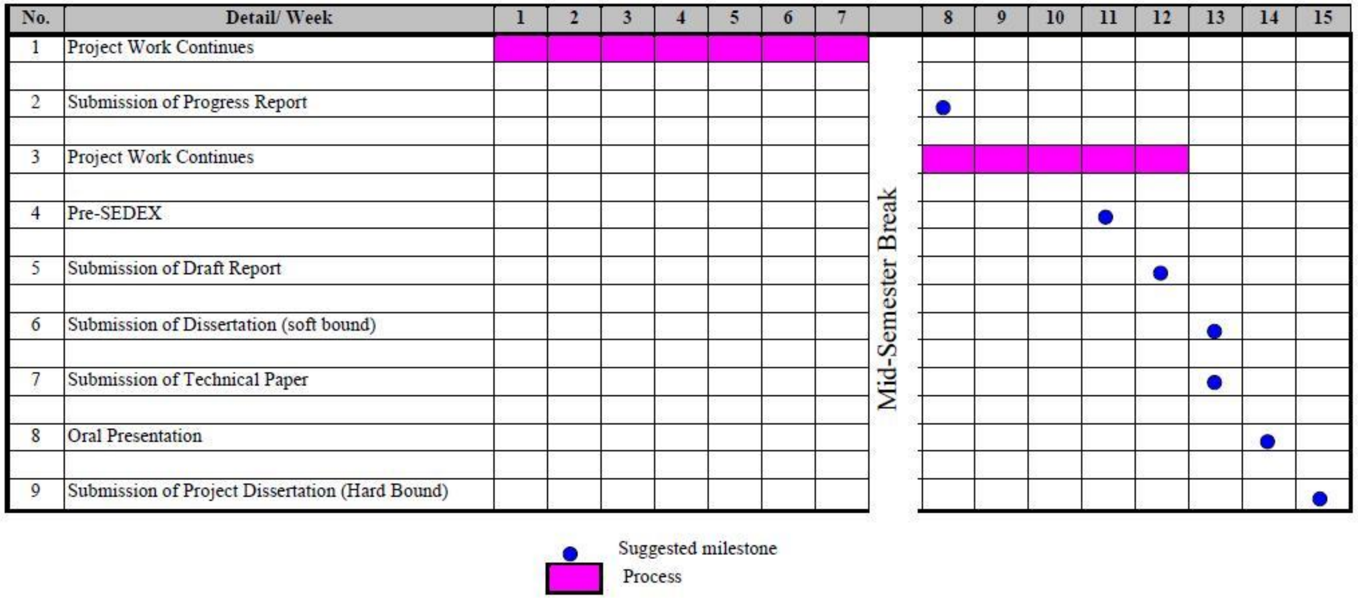
The tree main tasks to be completed for FYP I are:

a. Extended Proposal

b. Proposal Defence

c. Interim Report

On the other hand, 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

Figure 14: FYP II Gantt Chart

The main tasks for FYP II are:

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

3.3 Tools

For this project, the software being used to do the simulation is ASPEN HYSYS. With the feed properties, estimations of equipment parameters and the process flow diagram, this software can simulate the process and generate the composition of the products produced. Once a result is obtained, the product composition would be compared with the desired specifications. Should the specifications not meet, changes should be made in the process parameters until the desired product is obtained.

3.4 CSU Modelling

In the modelling of the condensate stabilisation unit, the main equipment that governs the process is the stabilisation column. Besides that, the feed stream is also one of the important objects that need to be defined in the simulator. This section briefly describes how these two objects are constructed using the HYSYS (ver. 2006) software. The following tables, Table 2, 3 and 4 summarises the data input for the said objects in order to run the simulation.

Table 2: Feed Stream Conditions

Feed Stream	
Normal Flow, kmol/h	4645
Pressure, barg	30.7
Temperature, °C	21.4

Table 3: Feed Stream Composition

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

Table 4: Stabiliser Column Operating Conditions

Distillation Column	
Reboiler Pressure, barg	8.5
Condenser Pressure, barg	8.0
Reboiler Temperature, °C	180

CHAPTER 4

4.0 RESULTS AND DISCUSSION

4.1 Process Description

Figure 15 shows the process flow diagram of the simulation of the main condensate stabilization unit using Aspen HYSYS (ver. 2006) software. As can be seen in the PFD, the main equipment governing the stabilization process is the stabilizer column. Before entering the column, the feed is first heated up by the product stream in heat exchanger E-100. Then, the feed is then routed to a desalter where the salt is removed by an electrostatic desalting process. However, the desalting unit is not shown in the simulation PFD since HYSYS cannot simulate the electrostatic desalting process. Nevertheless, this matter will be further discussed in section 4.5.

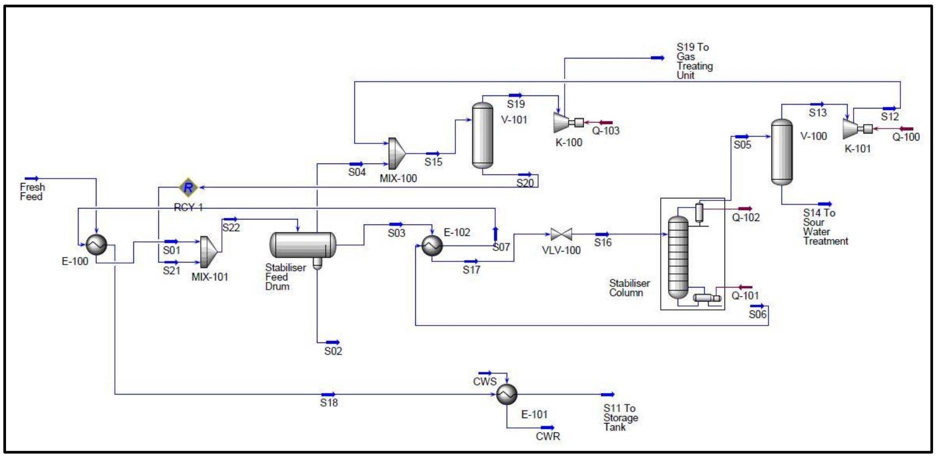


Figure 15: Process Flow Scheme of the Simulated Condensate Stabilization Unit

From the desalter, the brine water will be sent to the water treatment unit while the condensate will be sent to a 3-phase separator where gaseous and aqueous phases will be separated from the condensate. From the separator, the condensate is once again heated by the product stream and is then sent to the stabilizer column. Upon entering, the feed is routed through a valve to reduce the pressure of the feed. The column is operated at a pressure of 8.5 barg and reboiler temperature of 180°C. The bottom product of the column is the stabilized condensate which will then be cooled by the feed streams as well as cooling water. The final product would be stabilized condensate at 70°C with RVP of 8.778 psia. The complete simulation data can be found in Appendix I.

The top product of the column consists of the lighter components that have been removed from the condensate. This stream will be compressed and combined with the light gas stream from the 3-phase separator. The combined stream will be compressed again and is sent to the gas treating unit for further processing.

4.2 Process Feed

The feed used for the simulation in this project is the Summer Rich feed from an Iranian Reservoir South pars project. The composition and properties of the feed is as tabulated in Tables 5 and 6.

Table 5: Composition of Feed

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

Table 6: Properties of Feed

Properties	
Total	
Normal Flow, kmol/h	4645
Normal Flow, kg/h	325604
Heat Flow, kW	4009
Molecular Weight	70.1
Pressure, barg	30.7
Temperature, °C	21.4
Vapour	
Molar Flow, MMSCFD	16
Normal Flow, kg/h	15708
Density, kg/cu m @ P, T	28
Liquid	
Standard Liq Vol Flow, SBPD	65284
Normal Flow, kg/h	309896
Actual cu m/h @ P , T	411
S. G. Liquid @ P, T	0.753

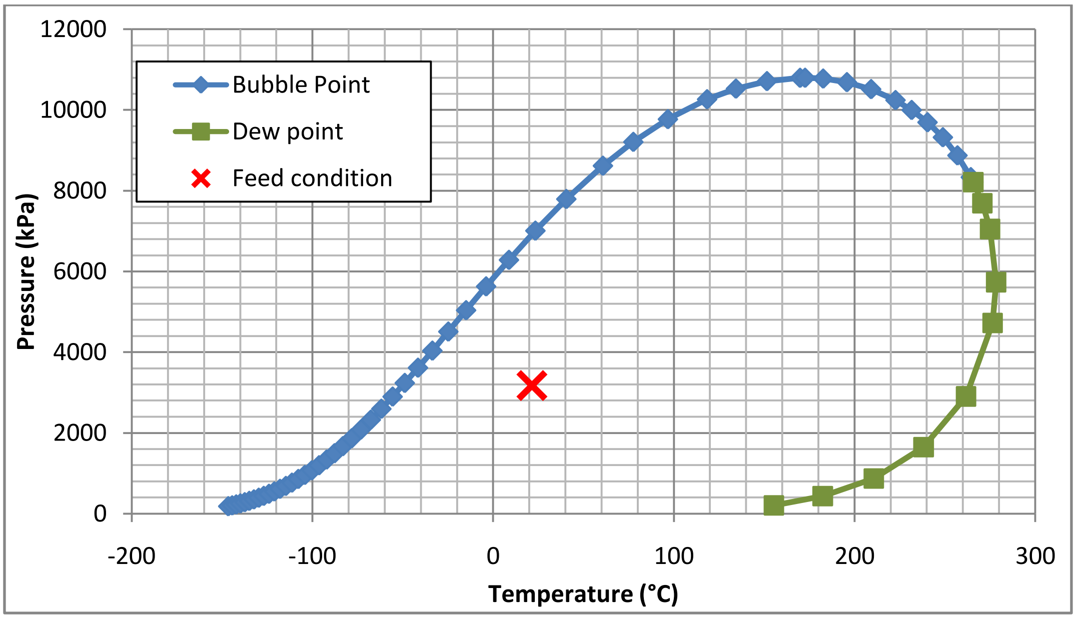


Figure 16: Phase Envelope Curve for Inlet Feed

The phase envelope diagram in Figure 16 shows the bubble points and dew points of the condensate at different pressures. According to the Schlumberger Oilfield Glossary, between the bubble point and dew point curves, the hydrocarbon is in two phases. The feed to the condensate stabilization unit is at 21.4°C and 3171kPa. As can be seen in the graph above, the red ‘X’ that represents the feed condition is located in the two-phase region. Besides that, from HYSYS simulation, it is found that the feed is 0.16 vapour, 0.66 liquid and 0.18 aqueous. This indicates that the feed is in both gaseous and liquid phases and also contains water in aqueous phase and thus, it can be processed in the CSU.

4.3 Simulation Validation

In order to ensure that the simulation done in this project is valid, the composition of the final product is compared to composition of final product in other simulation in previous study as well as in a real plant.

Figure 17 shows a graph of component mole fraction vs. component. The three different trends represents three different data which are the plant data from Iranian Reservoir South pars Project, data from Pro/II software (from previous study) and data from this project's simulation using HYSYS (ver. 2006) software. The complete data in table form can be referred to in Appendix II.

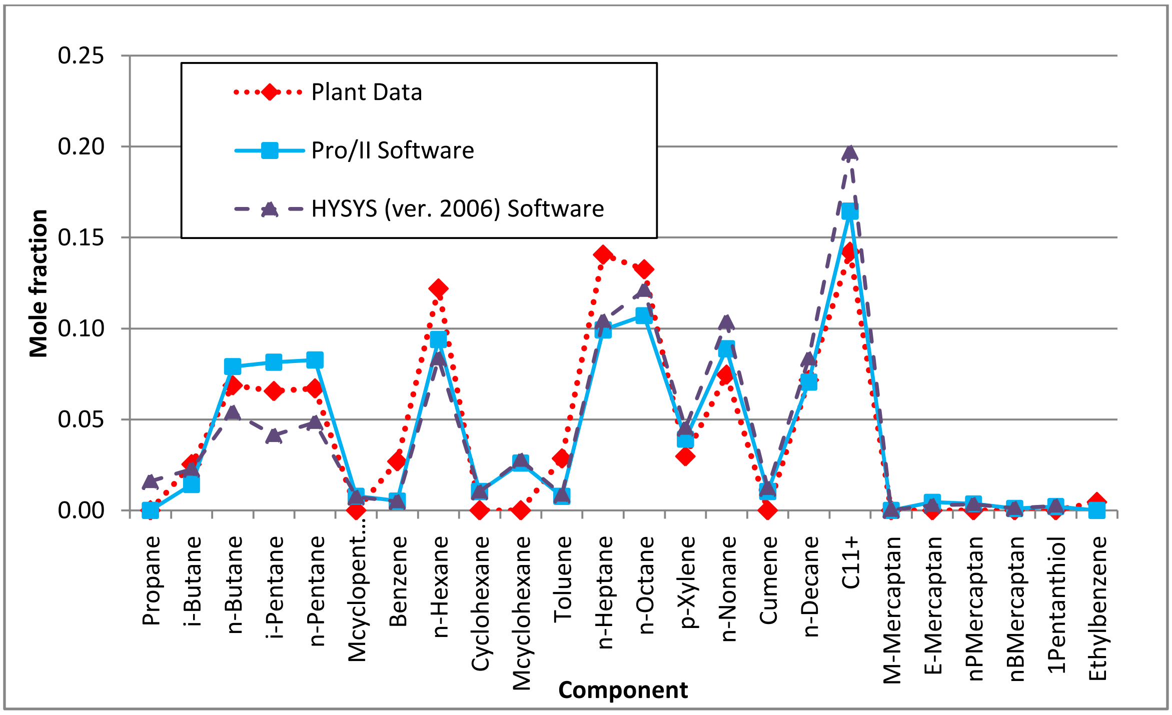


Figure 17: Comparison of Product Composition

In the first part of the graph from propane to mcyclopentane, it can be seen that the Pro/II software data gives the highest mole fractions followed by the plant data and then HYSYS data. Moving on to the heavier components, i.e. from benzene to n-

Octane, it is observed that the plant data gives off the highest mole fraction followed by HYSYS data and then Pro/II data. As for the six heaviest hydrocarbons from p-xylene to C11+, the HYSYS simulation results in the highest mole fraction followed by Pro/II simulation and plant data. Lastly, there are only trace amounts of mercaptans, 1pentanthiol and ethylbenzene in all three sets of data.

From the above analysis, it can be concluded that the Pro/II simulation condensate contains the most light components followed by the real plant data. Lastly, HYSYS simulation results in the condensate with the most heavy components. Therefore, it can be said that the Pro/II simulation results in the lightest condensate product, the HYSYS simulation results in the heaviest condensate product and the real plant results in a rather balanced condensate as compared to the other two.

However, in general, the trend of the mole fraction of the components is similar for all three sets of data. There are no major differences and thus, it is proven that the simulation done using the HYSYS software is valid and can be run in real life plants.

4.4 Effects of Different Operating Conditions

In real life plants, the process is not at a steady state since there are always fluctuations in the process parameters. This may be due to many reasons such as changing surrounding conditions, upset in other related process units upstream, and breakdown of related equipments. As a result of these parameter changes, the specifications of the product may also change. Therefore, it is important to know how much of these changes that the process can tolerate and at which point the parameter change will cause the product to become off specification. In order to obtain those data, a one-dimensional study is done on the simulated CSU by changing the feed flow rate, feed temperature, feed pressure and reboiler temperature, one at a time. The product specifications that are monitored in this study are the product RVP and the sulphur content.

In order to study the effects of the four different operating parameters, all other values except the parameter being studied need to be kept constant. Table 7 shows what parameter is kept constant for each study where C represents constant and V represents variable.

The findings from these four studies are discussed in the following section. The complete data obtained in the study can be referred to in Appendix III.

Table 7: Status of operating conditions for the study of effects of changing operating conditions

Parameter		Study of the effect of:			
		Feed Flow Rate	Feed Temperature	Feed Pressure	Reboiler Temperature
Feed Properties	Flow Rate	V	C	C	C
	Temperature	C	V	C	C
	Pressure	C	C	V	C
Heat Exchanger Duties		C	C	C	C
Reboiler Duty		C	C	C	V
Compressor Power		C	C	C	C
Column Pressure		C	C	C	C

4.4.1 Reid Vapour Pressure (RVP)

The main product specification that is considered for a CSU is the RVP of the condensate. Therefore, the RVP of the product is the most important specification that needs to be monitored closely during the operation of the CSU. The lower the RVP of the product, the higher it’s quality is. The standard method for measuring RVP is ASTM D323. As stated earlier, in this project, the effects of parameter changes on the RVP has been studied by changing four different parameters, i.e. feed flow rate, feed temperature, feed pressure and reboiler temperature.

4.4.1.1 Effect of Feed Flow Rate

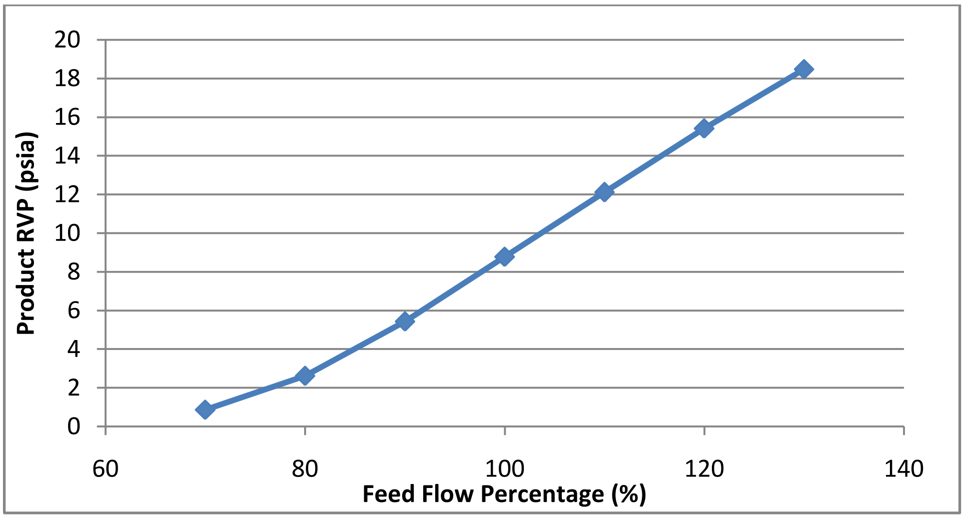


Figure 18: Effects of Feed Flow Rate towards Product RVP

The normal feed flow rate used for the base case study is 4645 kgmole/hr. The flow rate is then decreased to 70% and then increased to 140% in 10% intervals. Figure 18 shows how the change in feed flow rate affects the RVP of the condensate. From the graph, it can be seen that as the flow rate increases, the RVP also increases. This increase in RVP is because when the flow rate increases, more heat is required to flash off the light components in the condensate. Since the column reboiler duty is kept constant, there is insufficient heat to maintain a constant RVP. Therefore, the RVP would gradually increase with the increase of feed flow rate. For a maximum RVP of 10 psia and 12 psia, the maximum flow rate percentage that can be processed by the CSU is at 103% and 110% respectively.

4.4.1.2 Effect of Feed Temperature

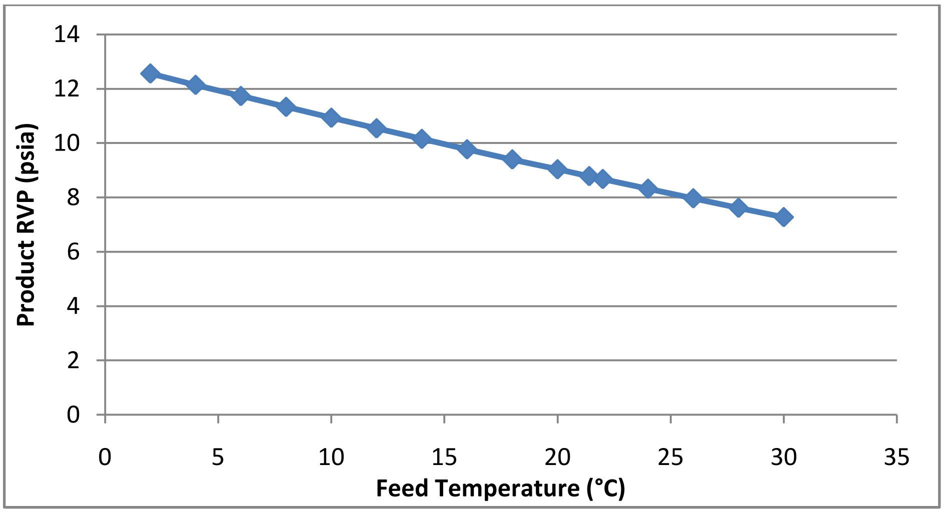


Figure 19: Effect of Feed Temperature Towards Product RVP

The condensate fed to the CSU is normally at 21.4°C. In order to study the effects of feed temperature towards product RVP, the temperature is decreased to 2°C and then increased to 30°C at 2°C intervals. As can be seen in Figure 19, as the temperature of the feed is increased, the product RVP gradually decreases. The increase in temperature would cause more portions of the light components to flash off from the condensate and thus reduce the RVP of the product. The minimum temperature that the CSU can tolerate in order to achieve the specified RVP for summer is 14°C and for winter is 4°C.

4.4.1.3 Effect of Feed Pressure

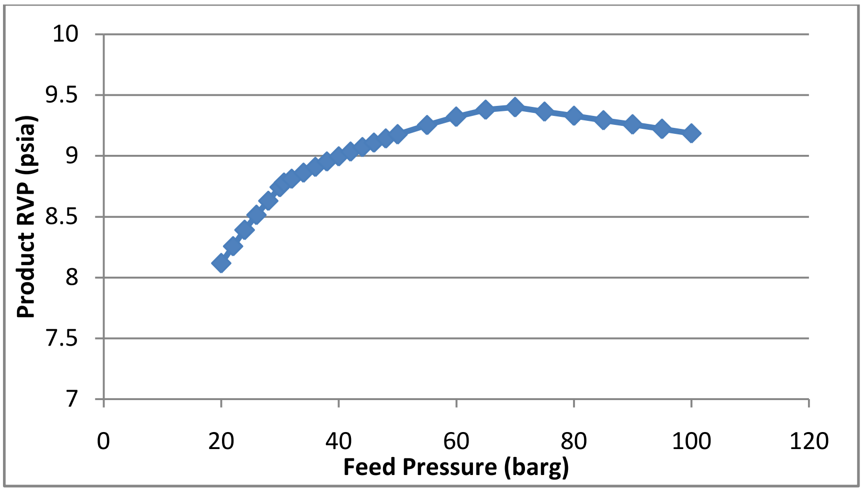


Figure 20: Effects of Feed Pressure Towards Product RVP

At normal conditions, the pressure of the feed to the CSU is at 30.7 barg. At first, to study the effects of the feed pressure towards the product RVP, the feed pressure is reduced to 20 barg and then increased to 50 barg at 2 barg intervals. However, after tabulating the data and plotting it in a graph, the trend was gradually increasing and still hasn't reached the 10 psia RVP limit for summer conditions. Therefore, in order to find the maximum pressure limit for 10 psia and 12 psia RVP specifications, the range of the feed pressure is increased from 50 barg to 100 barg in 5 barg intervals. As can be seen in Figure 20, even though the range is expanded to 100 barg, the RVP still doesn't rise to over 10 psia. On the contrary, once the feed pressure has reached 70 barg, the RVP climb stops and starts to decrease as the pressure further increases. This shows that even though the change in feed pressure will affect the product RVP, the RVP will not go off specification since the highest RVP reached is only at 9.4 psia.

4.4.1.4 Effect of Reboiler Temperature

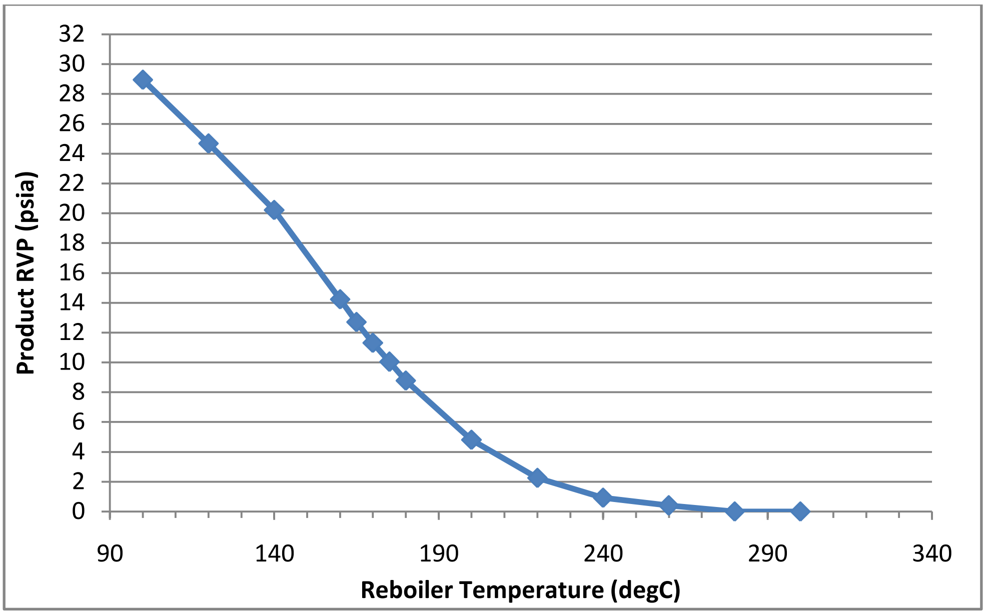


Figure 21: Effects of Reboiler Temperature Towards Product RVP

From the HYSYS simulation, it was found that 180°C is the optimum reboiler temperature required at normal conditions in order to achieve an RVP of lower than 10 psia. For the study of the effects of reboiler temperature towards the product RVP, the reboiler temperature is reduced to 100°C and then increased to 300°C at 20°C intervals. The graph in Figure 21 shows that as the column reboiler temperature is increased, the product RVP is decreased. This decrease in the RVP is because as the temperature increases, more light components will flash off from the condensate thus leaving less amount of volatile component in the product. This is similar to the effect of feed temperature increase. However, the trend for feed temperature is almost linear while the trend for reboiler temperature is polynomial. To ensure that the condensate product is within required specifications for summer and winter conditions, the reboiler temperature must not fall below 175°C and 167°C respectively.

4.4.2 Sulphur Content

The sulphur content in the product is measured in parts per million in weight (ppm wt.). The total mass of sulphur calculated by extracting the molar flow rate of each component that contains sulphur, multiplying it by the coefficient of sulphur in each component and then multiplying with the atomic mass of sulphur. The total mass is then divided by the total mass of the condensate and multiplied by 10^6. The formula used is as follows:

$$Sulfur\ content\ (ppm\ wt.) = \frac{\sum_{i=0}^{n} X_i \cdot A_i \cdot MW}{m} \times 10^6$$

Where:

i = component containing sulphur

n = total number of component containing sulphur

X_i = molar flow rate of component *i* containing sulphur

A_i = coefficient of sulphur in component *i*

MW = atomic mass of sulphur (32.065 kg/kgmole)

m = total mass flow rate of condensate (product)

The molar flow of the dominant sulphur component is also observed to see the effects of the changes in the parameters.

The manipulation of the four parameters for the study of their effects to sulphur content is done in the same method as the study of their effects on product RVP. The range and interval size of the parameter manipulation is also the same.

4.4.2.1 Effect of Feed Flow Rate

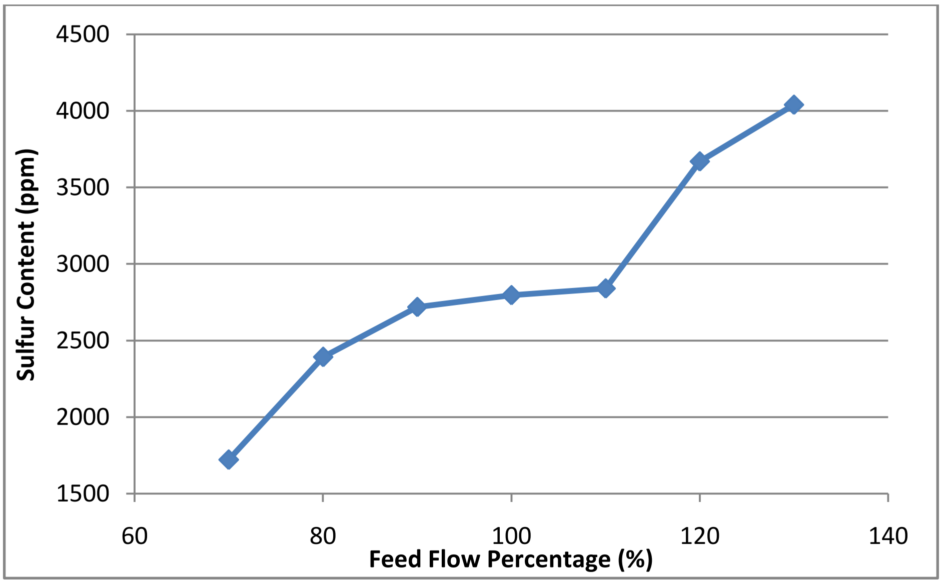


Figure 22: Effect of Feed Flow Rate towards Sulphur Content

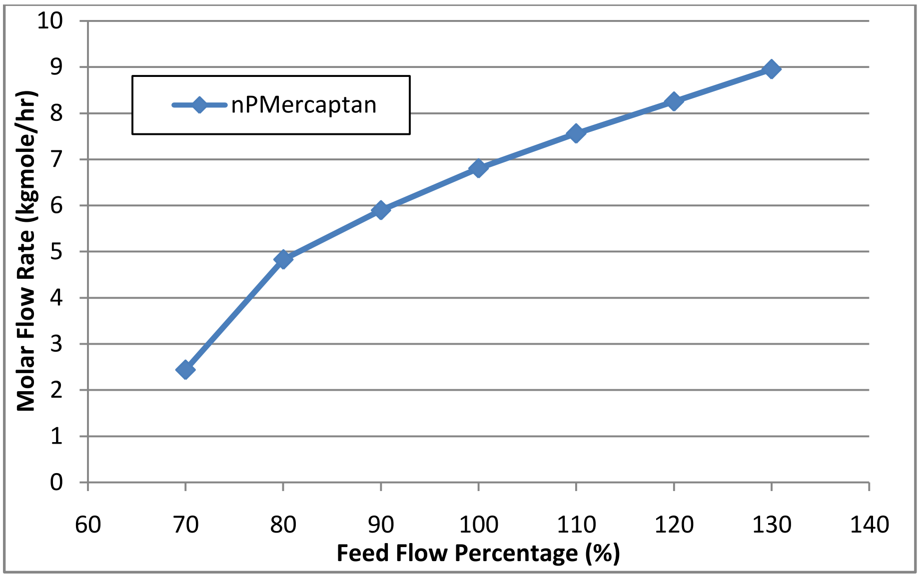


Figure 23: Effect of Feed Flow Rate towards nPMercaptan content

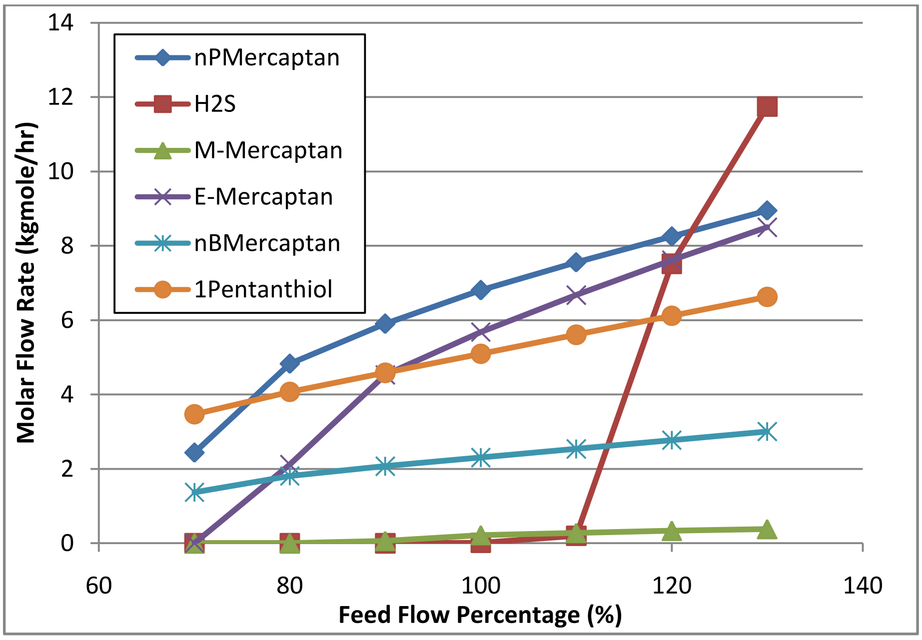


Figure 24: Effect of Feed Flow Rate towards Sulphur Component Content

As can be seen in Figure 22, as the feed flow rate is increased the sulphur content in the condensate also increases. This is because as the feed increases, then the total amount of sulphur is also increased. However, the heat exchanger duties as well as the column duty are kept constant. Therefore, the amount of sulphur that can be removed from the condensate is also constant despite the increase in feed flow rate, thus causing the total sulphur content to increase.

If the graph is closely observed, it can be seen that the sulphur content started to become constant at feed flow percentage of 90%-100%. After 100%, the sulphur content then started to increase rapidly again.

To investigate this occurrence, a graph of the molar fraction of the dominant sulphur component, nPMercaptan versus feed flow rate is plotted (Figure 23). However, from the said graph, it can be seen that the increment of the nPMercaptan flow rate is fairly linear to the increment of feed flow rate and that doesn’t indicate any kind of sudden increase like the one in Figure 22. To further investigate on this matter, a graph of flow rate of all sulphur components versus feed flow rate is constructed (Figure 24). As can be seen in the graph, all the sulphur component flow rates increase gradually as the feed flow rate is increased. All except H_2S that shows very little increase up until 110% flow rate after which the flow rate increases rapidly. Therefore, it can be concluded that this sudden increase in H_2S flow rate is the cause of the sudden increase in overall sulphur content in the condensate.

4.4.2.2 Effect of Feed Temperature

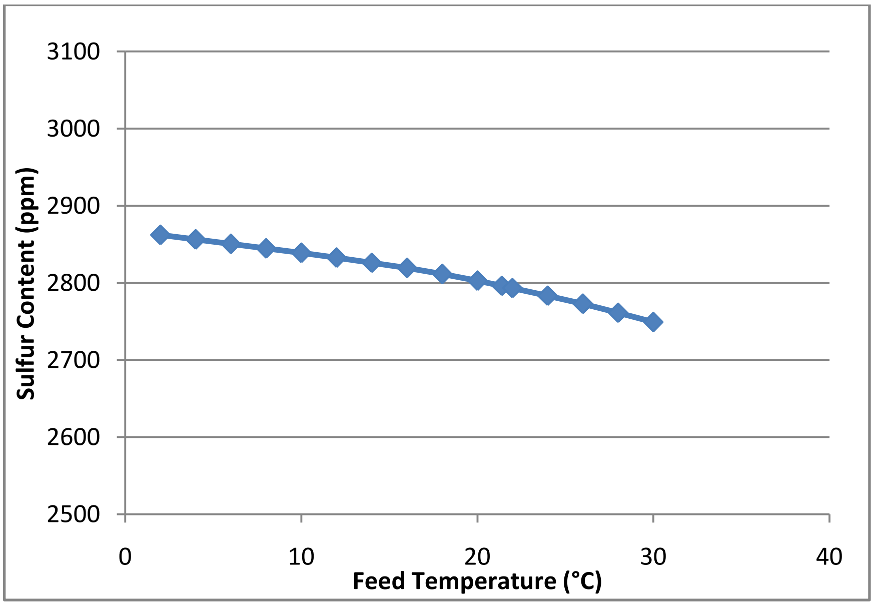


Figure 25: Effect of Feed Temperature towards Sulphur Content

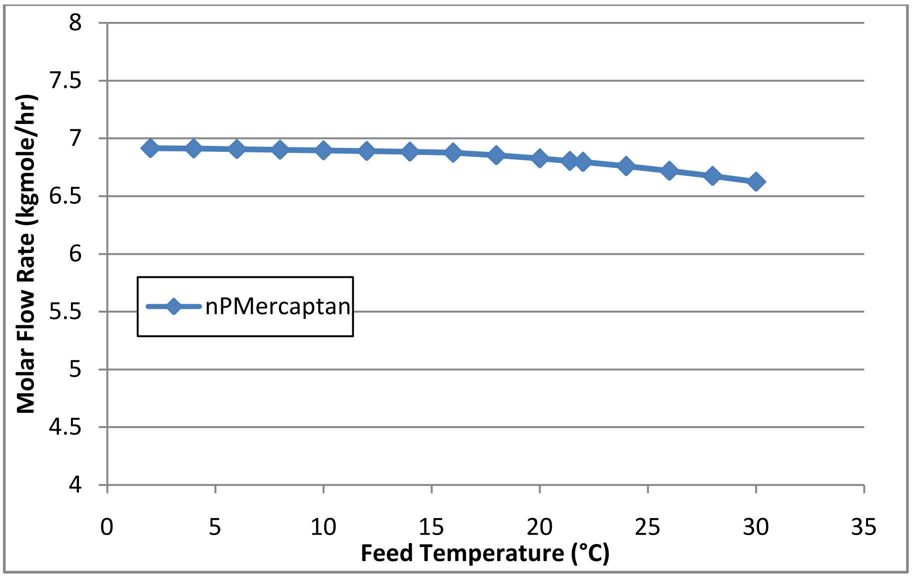


Figure 26: Effect of Feed Temperature towards nPMercaptan Content

When the feed temperature is increased, in can be seen in Figure 25 that the sulphur content will gradually decrease. This phenomenon is similar to the effects of the feed temperature towards product RVP. When the feed temperature is increased, the amount of sulphur that can be vaporised off from the condensate increases. Thus, it results in lower sulphur content in the stabilised condensate. Figure 26 shows that the increase in feed temperature also causes a gradual decrease in the molar flow rate of nPMercaptan in the condensate.

4.4.2.3 Effect of Feed Pressure

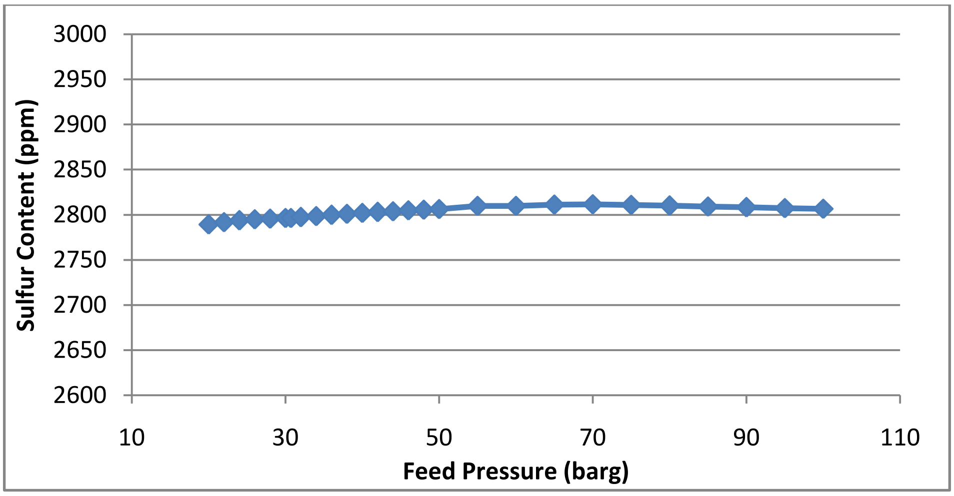


Figure 27: Effect of Feed Pressure towards Sulphur Content

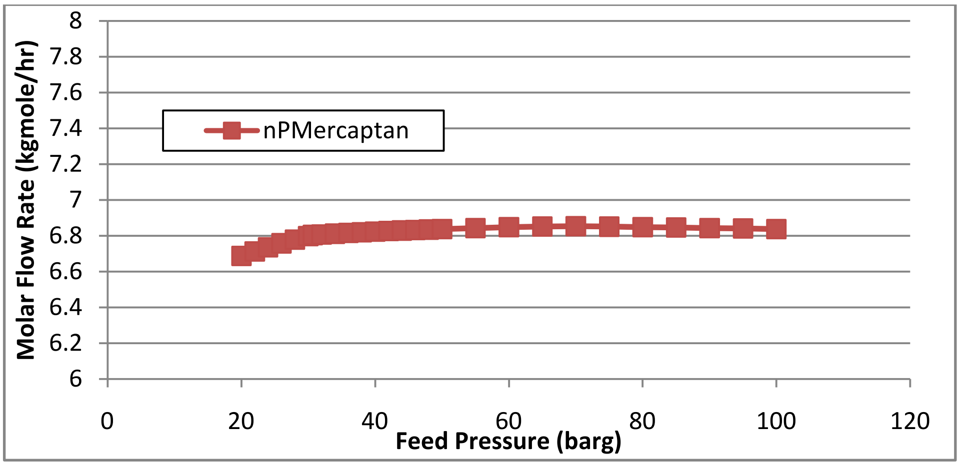


Figure 28: Effect of Feed Pressure towards nPMercaptan Content

The effect of feed pressure towards the sulphur content is rather different compared to feed flow rate and feed temperature. For feed flow rate and feed temperature, the sulphur content would either increase or decrease. However, for feed pressure, as the pressure is increased, the sulphur content is not affected as much. As can be seen in Figure 27, as the feed pressure is increased, the sulphur content only increases a little bit and then starts to maintain at a constant value. This shows that the feed pressure doesn’t give a big impact on the sulphur content of the condensate. Figure 28, shows the effect of feed pressure towards flow rate of nPMercaptan. The trend is similar to the previous graph.

4.4.2.4 Effect of Reboiler Temperature

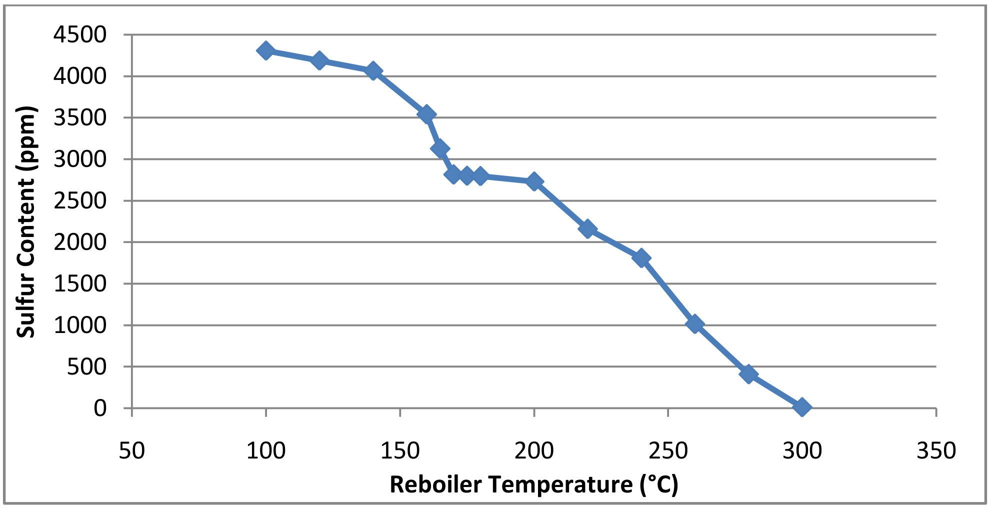


Figure 29: Effect of Reboiler Temperature towards Sulphur Content

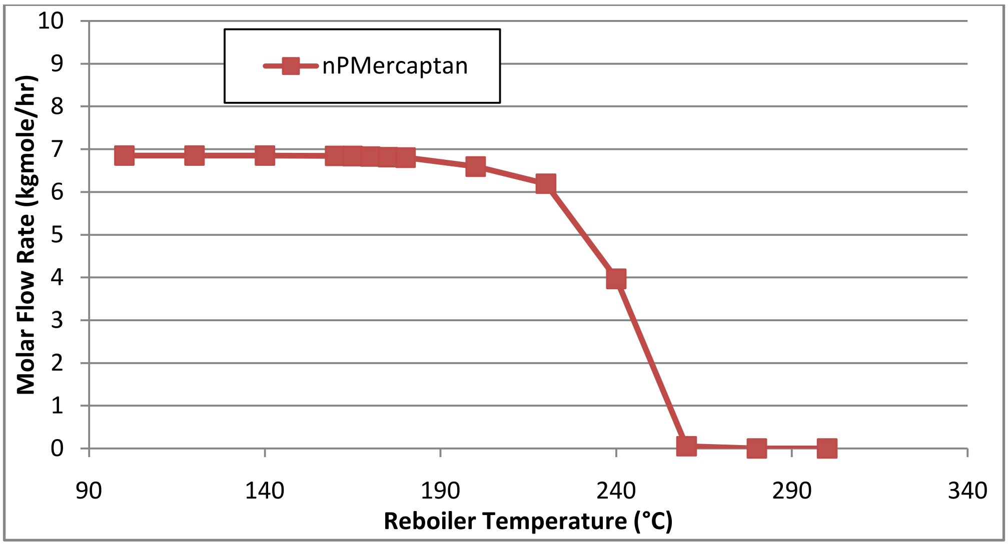


Figure 30: Effect of Reboiler Temperature towards nPMercaptan Content

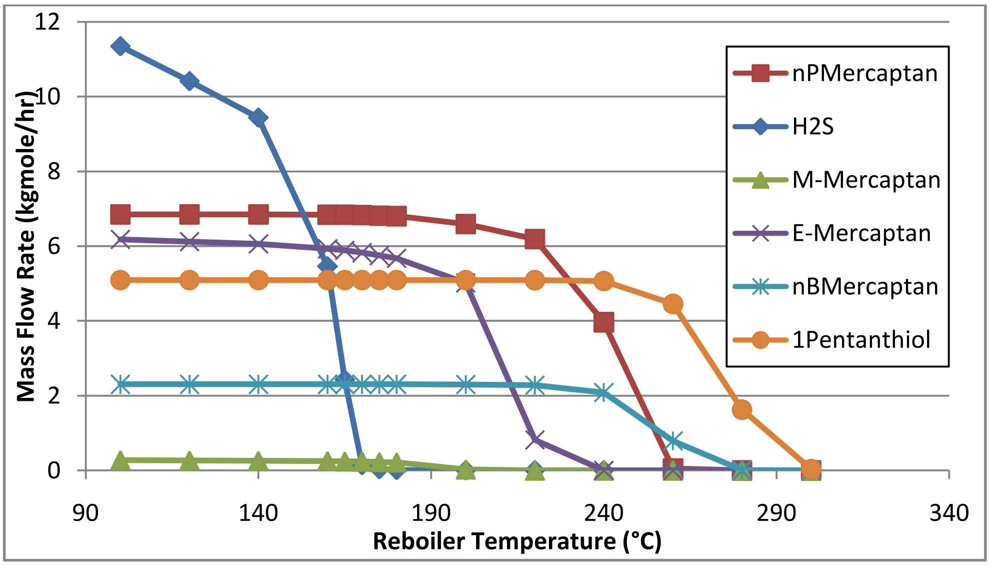


Figure 31: Effect of Reboiler Temperature towards Sulphur Component Content

Figure 29 shows how reboiler temperature affects the amount of sulphur in the condensate product. As can be observed in the graph, as the reboiler temperature is increased, the sulphur content decreases. This is because the high temperature in the column enables more sulphur components to be vaporised off from the condensate. The decrease of sulphur is almost constant up till 140°C after which the line became rather steep. From temperature 170°C to 200°C, the curve became quite horizontal where the sulphur content didn’t vary much. However, at temperature above 200°C, the sulphur content started decreasing again. A graph of flow rate of nPMercaptan versus reboiler temperature (Figure 30) is constructed to look further into the unusual trend in Figure 29. From Figure 30, it can be seen that the change in nPMercaptan flow with respect to reboiler temperature is rather normal.

To look further into this issue, the molar flow rate of all the sulphur components is plotted in a graph against the reboiler temperature (Figure 31). From Figure 31, it can be seen that the different components have a rapid decrease in flow rate at different temperatures. This is because of the different boiling points of the components that would cause each component to completely vaporise at different temperatures. Hence, the overall sulphur content would also be affected by these different trends of each sulphur component.

Table 8: Boiling Points of Each Sulphur Component

Component	Boiling Point (°C)
H_2S	-59.65
M-Mercaptan	5.946
E-Mercaptan	35.65
nPMercaptan	66.05
nBMercaptan	98.46
1Pentanthiol	126.6

Table 8 shows the boiling point of each of the components in ascending order. By comparing the order of these components in increasing boiling point and the order at which the components start decreasing rapidly in Figure 31, it can be seen that the order is the same. This further supports the statement that the different boiling points affect the rate of vaporisation of the sulphur components.

4.5 Salt Removal

Gary and Handwerk (2001) mentioned in their book that if the hydrocarbon contains salt more than 10 PTB (lb/1000bbl), it requires desalting to reduce fouling and corrosion. The salt content in the Summer Rich condensate is given to be 8 tons/day which is equivalent to 734.9 lb/hr. The total condensate flow rate is given as 325,604 kg/hr which is equivalent to 8,467.38 barrels. This results in a total salt content of 86.79 PTB in the condensate. Since it is more than 10 PTB, the condensate would be required to go through desalting process before it can be stabilised in the CSU.

As mentioned in section 2.5, the most commonly used desalting method is the electrostatic demulsification which utilises electrocoalescence to break the water-oil emulsion in the desalter vessel. Figure 32 shows the schematic of a crude oil desalting/dehydration plant. Mahdi et al. (2008) stated that the desalting process comprises of six main steps namely separation by gravity settling, chemical injection, heating, addition of less salty water (dilution), mixing and electrical coalescing.

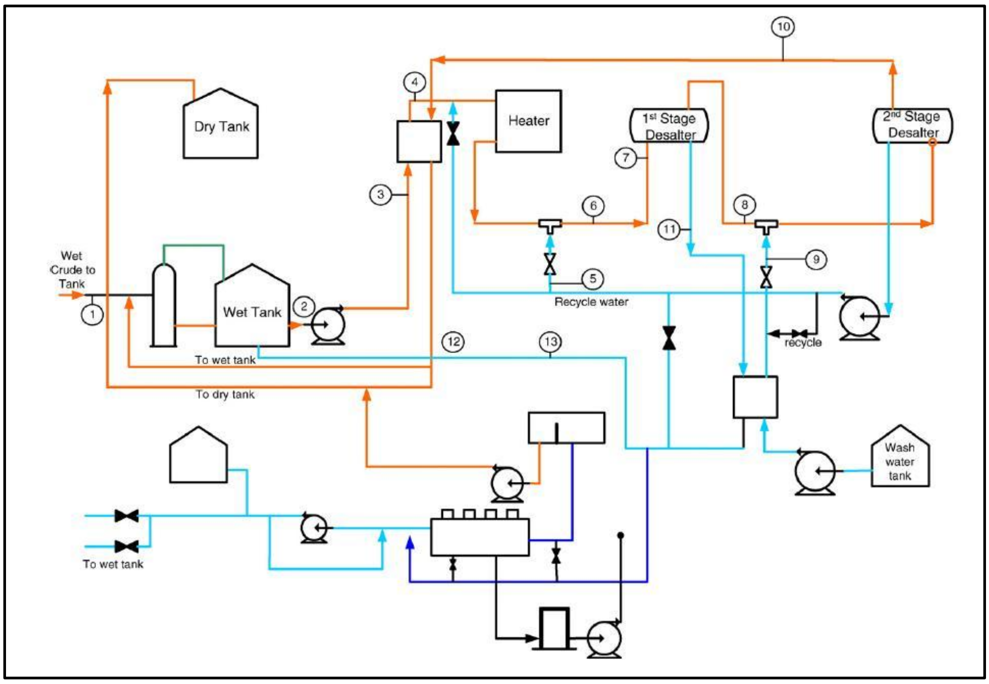


Figure 32: Schematic of Crude Oil Desalting/Dehydration Plant (Mahdi et al., 2008)

For a typical desalting/dehydration plant such as in Figure 32, the process starts at the wet tank where the crude is settled for a few hours (Mahdi et al., 2008). After settling, some of the water would have been separated from the crude and is flowed out of the tank through stream 13. On the other hand, the crude is pumped out of the tank after being injected with emulsifier. The pump then pumps the emulsion to the heat exchanger where it will be heated up by the desalted crude stream (stream 10). The emulsion coming out of the exchanger is then further heated up by the water bath heater. Then, the emulsion is mixed with recycle water from the second stage desalter. A mixing valve (globe valve) is used to agitate the recycled water and emulsion through induced shearing force. Stream 7 then leaves the mixing valve and enters the first stage desalter. In the desalter vessel, the emulsion is exposed to a high voltage electrostatic field that will cause coalescence of the dispersed water. The large water droplets will then settle at the bottom of the vessel and is removed through stream 11. The desalted crude is sent to a second stage desalter where the desalting process as in the first stage is repeated to remove any remaining salts and impurities. The crude is mixed with fresh water at the mixing valve and is then separated in the vessel. The brine water coming out of the vessel is recycled and used again in the first stage desalter. This recycling is for the sake of minimising fresh water consumption. On the other hand, the desalted crude is used to heat up the emulsion coming out of the wet tank and is then sent to the dry tank (Mahdi et al., 2008).

The process described in the previous paragraph is the typical two-stage desalting process used for desalting crude oil. The crude oil processed in the said desalter plant has an API gravity of 31.7°API. The Summer Rich condensate on the other hand, has an API gravity of 307.12°API which comes to show that the condensate is much lighter than the crude. Therefore, it should be much easier to remove salt from the condensate using the same process.

4.5.1 Optimum Parameters for Desalting

In their study, Vafajoo et al. (2012) have investigated the effects of temperature, injected chemicals and the pH of the crude oil associated water towards an electrostatic desalter on one of the oil platforms owned by Iranian Oil Offshore Company in the Persian Gulf. The crude oil used in the study has an API gravity of 19°API and salt content of 6161 PTB. From the study, it is found that an increase in the demulsifier concentration would require the temperature to be lowered. Besides that, a higher pH of associated water would increase the efficiency of the desalting process.

In another study, Mahdi et al. (2008) investigated the effects of the demulsifying agent concentration, temperature, wash water dilution ratio, settling time and mixing time towards the desalting process. The study used the Fractional Factorial Design (FFD) to find the factors that had a significant effect on the efficiency of the desalter. From the study, it was found that the most optimum parameters for desalting process are as follows:

Table 9: Optimum Parameters for Desalting Process (Mahdi et al., 2008)

Temperature (°C)	77
Settling time (min)	3
Mixing time (min)	9
Demulsifying agent concentration (ppm)	15
Wash water dilution ratio (%)	10

At the stated parameters, a removal of 93.28% of salt is achieved.

4.5.2 Desalting of Condensate in the CSU

As have been mentioned earlier, the given Summer Rich condensate has salt content of 86.79 PTB and an API gravity of 307.12°API. The condensate is much lighter that the crude used in the study by Mahdi et al. which will make it easier for salt removal. Therefore, it can be deduced that the condensate can be desalted at the similar conditions and result in similar salt removal if not better.

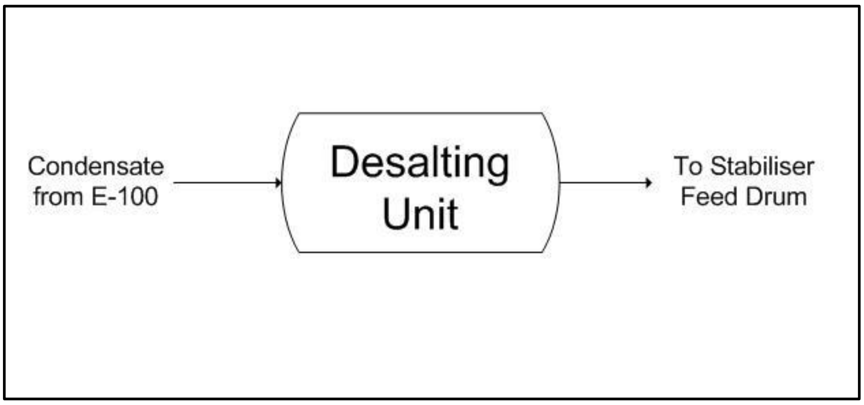


Figure 33: Desalting Unit in the CSU

Figure 33 above shows the desalting unit that is to be installed in the CSU to remove salt from the condensate. The desalting unit is to be comprised of a two stage desalting process similar to the one in Figure 32. According to Mahdi et al., at the parameters stated in Table 7, 93.28% of salt can be removed. Therefore, the condensate exiting the desalting unit should only contain 5.83 PTB of salt. This value is below 10 PTB thus is already satisfactory to minimise fouling and corrosion.

CHAPTER 5

5.0 CONCLUSION AND RECOMMENDATIONS

5.1 Conclusion

From this research, it has been found that at a stabiliser column pressure of 8.5 barg and reboiler temperature of 180°C, the condensate can be stabilised to an RVP of 8.778 psia. This would satisfy both the summer and winter condition limits of 10 psia and 12 psia. Based on the comparison of the HYSYS simulation product composition with the real plant data and simulation data from previous study, it can be concluded that the HYSYS simulation carried out is valid and can be used for further study.

From the study of the effects of changes in parameters towards the product properties, it has been found that the feed flow rate, feed temperature, feed pressure and reboiler temperature all affect the product RVP and sulphur content. The increase in feed flow rate causes an increase in RVP and sulphur content. The increase in feed temperature and reboiler temperature would both cause a decrease in RVP and sulphur content. On the other hand, when the feed pressure is increased, the RVP would increase at first and after 70 barg, it would start to decrease. Besides that, with increasing feed pressure, the sulphur content is not affected as much. At first it increases and then after a certain point, it starts to become constant.

Salt is one of the impurities that are usually present in gas wells and oil wells. A high concentration of salt can cause problems in downstream process as the presence of salt can cause corrosion and fouling in process equipments. Salt content cannot exceed 10 PTB in order to ensure corrosion and fouling can be minimised. The most efficient and most widely used method of desalting is the electrostatic desalting that uses electrostatic to cause coalescence in the water-oil emulsion.

5.2 Recommendations

There are a few other aspects of this research that can be approached in order to complete this project. The recommendations are as below:

a. Compare the simulation data with data from Malaysian market or Malaysian reservoir to validate whether this simulation is suitable in this country.
b. Another parameter that can be studied for its effects towards the product RVP and sulphur content is the column pressure.
c. Produce a simulation of the desalting process of the condensate to accurately predict the salt content at the outlet of the desalter.
d. Include an economics study that would take into account the cost of utilities and processing and find the most optimum conditions that would result in highest gross profit.
e. Conduct another simulation with a different simulator in order to compare its results with the current results and investigate what causes the differences (if any).
f. Conduct a design of experiments (DOE) study that could investigate the effects of more than one parameter at once towards the product properties so as to find the optimum parameters where the product of highest quality can be obtained.

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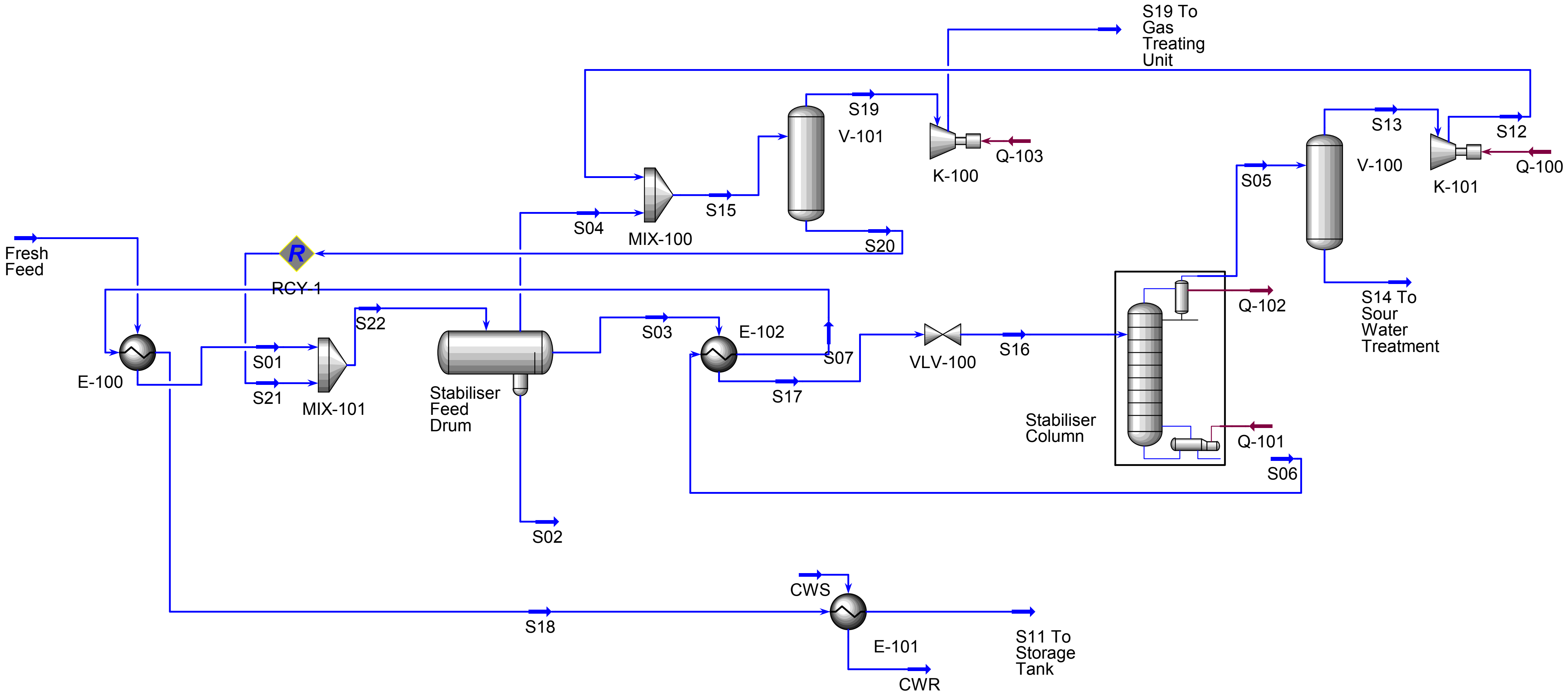
APPENDICES

APPENDIX I



Wed Oct 31 10:34:09 2012 Case: D:\ENGINEERING FUTURES~\~4th Year 1st Sem [Final Year Yawwww]\FINAL YEAR PROJECT\HYSYS\CSU_03 (DISTILLATION COLUMN)_REV12.HSC Flowsheet: Case (Main)

Workbook: Case (Main)

Material Streams

Fluid Pkg: All

Name		Fresh Feed	S01	S02	S03	S04
Vapour Fraction		0.1641	0.2114	0.0000	0.0000	1.0000
Temperature	(C)	21.40 *	55.00 *	55.00	55.00	55.00
Pressure	(kPa)	3171 *	3166	3166	3166	3166
Molar Flow	(kgmole/h)	4645 *	4645	819.7	2843	981.8
Mass Flow	(kg/h)	3.030e+005	3.030e+005	2.469e+004	2.564e+005	2.191e+004
Liquid Volume Flow	(m3/h)	462.7	462.7	23.36	380.2	59.20
Heat Flow	(kJ/h)	-9.186e+008	-8.933e+008	-2.686e+008	-5.380e+008	-8.670e+007

Name		S05	S06	S07	S11 To Storage Tank	S13
Vapour Fraction		0.9987	0.0000	0.0000	0.0000	1.0000
Temperature	(C)	26.63	180.0	146.1	70.00 *	26.63
Pressure	(kPa)	901.3	951.3	946.3	936.3	901.3
Molar Flow	(kgmole/h)	772.3	2071	2071	2071	771.3
Mass Flow	(kg/h)	2.576e+004	2.307e+005	2.307e+005	2.307e+005	2.574e+004
Liquid Volume Flow	(m3/h)	57.00	323.2	323.2	323.2	56.98
Heat Flow	(kJ/h)	-7.819e+007	-3.809e+008	-4.027e+008	-4.463e+008	-7.788e+007

Name		S14 To Sour Water Tr	S16	S17	S18	S12
Vapour Fraction		0.0000	0.0987	0.0381	0.0000	1.0000
Temperature	(C)	26.63	88.44	90.00 *	103.5	105.3
Pressure	(kPa)	901.3	2161	3161	941.3	3166 *
Molar Flow	(kgmole/h)	1.039	2843	2843	2071	771.3
Mass Flow	(kg/h)	21.68	2.564e+005	2.564e+005	2.307e+005	2.574e+004
Liquid Volume Flow	(m3/h)	2.131e-002	380.2	380.2	323.2	56.98
Heat Flow	(kJ/h)	-3.079e+005	-5.161e+008	-5.161e+008	-4.280e+008	-7.474e+007

Name		S15	S19 To Gas Treating	S19	S20	S22
Vapour Fraction		1.0000	1.0000	1.0000	0.0000	0.2114
Temperature	(C)	81.17	144.7	81.17	81.17	55.00
Pressure	(kPa)	3166	7101 *	3166	3166	3166
Molar Flow	(kgmole/h)	1753	1753	1753	0.0000	4645
Mass Flow	(kg/h)	4.765e+004	4.765e+004	4.765e+004	0.0000	3.030e+005
Liquid Volume Flow	(m3/h)	116.2	116.2	116.2	0.0000	462.7
Heat Flow	(kJ/h)	-1.614e+008	-1.562e+008	-1.614e+008	0.0000	-8.933e+008

Name		S21	CWS	CWR		
Vapour Fraction		0.3046	0.0000	0.0000		
Temperature	(C)	81.17 *	20.00 *	34.14		
Pressure	(kPa)	3166 *	101.3 *	96.32		
Molar Flow	(kgmole/h)	0.0000 *	1.665e+004	1.665e+004		
Mass Flow	(kg/h)	0.0000	3.000e+005 *	3.000e+005		
Liquid Volume Flow	(m3/h)	0.0000	300.6	300.6		
Heat Flow	(kJ/h)	0.0000	-4.773e+009	-4.755e+009		

* Specified by user.

Workbook: Case (Main) (continued)

Compositions

Fluid Pkg: All

Name	Fresh Feed	S01	S02	S03	S04
Comp Mole Frac (Methane)	0.2180 *	0.2180	0.0000	0.0995	0.7430
Comp Mole Frac (Ethane)	0.0544 *	0.0544	0.0000	0.0528	0.1044
Comp Mole Frac (Propane)	0.0518 *	0.0518	0.0000	0.0675	0.0497
Comp Mole Frac (i-Butane)	0.0189 *	0.0189	0.0000	0.0275	0.0098
Comp Mole Frac (n-Butane)	0.0389 *	0.0389	0.0000	0.0581	0.0158
Comp Mole Frac (i-Pentane)	0.0230 *	0.0230	0.0000	0.0359	0.0048
Comp Mole Frac (n-Pentane)	0.0258 *	0.0258	0.0000	0.0406	0.0043
Comp Mole Frac (n-Hexane)	0.0380 *	0.0380	0.0000	0.0612	0.0026
Comp Mole Frac (n-Heptane)	0.0467 *	0.0467	0.0000	0.0758	0.0014
Comp Mole Frac (n-Octane)	0.0541 *	0.0541	0.0000	0.0881	0.0007
Comp Mole Frac (n-Nonane)	0.0463 *	0.0463	0.0000	0.0755	0.0003
Comp Mole Frac (n-Decane)	0.0372 *	0.0372	0.0000	0.0607	0.0001
Comp Mole Frac (Mcyclopentan)	0.0033 *	0.0033	0.0000	0.0053	0.0002
Comp Mole Frac (Benzene)	0.0022 *	0.0022	0.0000	0.0035	0.0001
Comp Mole Frac (Toluene)	0.0038 *	0.0038	0.0000	0.0062	0.0001
Comp Mole Frac (Cyclohexane)	0.0046 *	0.0046	0.0000	0.0074	0.0002
Comp Mole Frac (p-Xylene)	0.0202 *	0.0202	0.0000	0.0329	0.0002
Comp Mole Frac (Cumene)	0.0054 *	0.0054	0.0000	0.0088	0.0000
Comp Mole Frac (Nitrogen)	0.0026 *	0.0026	0.0000	0.0006	0.0107
Comp Mole Frac (CO2)	0.0120 *	0.0120	0.0014	0.0087	0.0306
Comp Mole Frac (H2S)	0.0102 *	0.0102	0.0024	0.0106	0.0157
Comp Mole Frac (H2O)	0.1293 *	0.1293	0.7230	0.0013	0.0042
Comp Mole Frac (M-Mercaptan)	0.0001 *	0.0001	0.0000	0.0001	0.0001
Comp Mole Frac (E-Mercaptan)	0.0017 *	0.0017	0.0001	0.0026	0.0005
Comp Mole Frac (nPMercaptan)	0.0015 *	0.0015	0.0000	0.0024	0.0001
Comp Mole Frac (nBMercaptan)	0.0005 *	0.0005	0.0000	0.0008	0.0000
Comp Mole Frac (n-C11)	0.0878 *	0.0878	0.0000	0.1434	0.0001
Comp Mole Frac (COS)	0.0000 *	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (EGlycol)	0.0482 *	0.0482	0.2730	0.0000	0.0000
Comp Mole Frac (Mcyclohexane)	0.0124 *	0.0124	0.0000	0.0201	0.0004
Comp Mole Frac (1Pentanthiol)	0.0011 *	0.0011	0.0000	0.0018	0.0000

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	S05	S06	S07	S11 To Storage Tank	S13
Comp Mole Frac (Methane)	0.3665	0.0000	0.0000	0.0000	0.3670
Comp Mole Frac (Ethane)	0.1945	0.0000	0.0000	0.0000	0.1947
Comp Mole Frac (Propane)	0.2057	0.0159	0.0159	0.0159	0.2060
Comp Mole Frac (i-Butane)	0.0406	0.0226	0.0226	0.0226	0.0406
Comp Mole Frac (n-Butane)	0.0692	0.0540	0.0540	0.0540	0.0692
Comp Mole Frac (i-Pentane)	0.0217	0.0412	0.0412	0.0412	0.0218
Comp Mole Frac (n-Pentane)	0.0198	0.0484	0.0484	0.0484	0.0198
Comp Mole Frac (n-Hexane)	0.0016	0.0834	0.0834	0.0834	0.0016
Comp Mole Frac (n-Heptane)	0.0000	0.1041	0.1041	0.1041	0.0000
Comp Mole Frac (n-Octane)	0.0000	0.1210	0.1210	0.1210	0.0000
Comp Mole Frac (n-Nonane)	0.0000	0.1037	0.1037	0.1037	0.0000
Comp Mole Frac (n-Decane)	0.0000	0.0834	0.0834	0.0834	0.0000
Comp Mole Frac (Mcyclopentan)	0.0001	0.0073	0.0073	0.0073	0.0001
Comp Mole Frac (Benzene)	0.0001	0.0048	0.0048	0.0048	0.0001
Comp Mole Frac (Toluene)	0.0000	0.0085	0.0085	0.0085	0.0000
Comp Mole Frac (Cyclohexane)	0.0000	0.0102	0.0102	0.0102	0.0000
Comp Mole Frac (p-Xylene)	0.0000	0.0452	0.0452	0.0452	0.0000
Comp Mole Frac (Cumene)	0.0000	0.0121	0.0121	0.0121	0.0000
Comp Mole Frac (Nitrogen)	0.0021	0.0000	0.0000	0.0000	0.0021
Comp Mole Frac (CO2)	0.0319	0.0000	0.0000	0.0000	0.0319
Comp Mole Frac (H2S)	0.0389	0.0000	0.0000	0.0000	0.0389
Comp Mole Frac (H2O)	0.0050	0.0000	0.0000	0.0000	0.0037
Comp Mole Frac (M-Mercaptan)	0.0002	0.0001	0.0001	0.0001	0.0002
Comp Mole Frac (E-Mercaptan)	0.0022	0.0027	0.0027	0.0027	0.0022
Comp Mole Frac (nPMercaptan)	0.0001	0.0033	0.0033	0.0033	0.0001
Comp Mole Frac (nBMercaptan)	0.0000	0.0011	0.0011	0.0011	0.0000
Comp Mole Frac (n-C11)	0.0000	0.1969	0.1969	0.1969	0.0000
Comp Mole Frac (COS)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (EGlycol)	0.0001	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Mcyclohexane)	0.0000	0.0276	0.0276	0.0276	0.0000
Comp Mole Frac (1Pentanthiol)	0.0000	0.0025	0.0025	0.0025	0.0000

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	S14 To Sour Water Tr	S16	S17	S18	S12
Comp Mole Frac (Methane)	0.0000	0.0995	0.0995	0.0000	0.3670
Comp Mole Frac (Ethane)	0.0000	0.0528	0.0528	0.0000	0.1947
Comp Mole Frac (Propane)	0.0000	0.0675	0.0675	0.0159	0.2060
Comp Mole Frac (i-Butane)	0.0000	0.0275	0.0275	0.0226	0.0406
Comp Mole Frac (n-Butane)	0.0000	0.0581	0.0581	0.0540	0.0692
Comp Mole Frac (i-Pentane)	0.0000	0.0359	0.0359	0.0412	0.0218
Comp Mole Frac (n-Pentane)	0.0000	0.0406	0.0406	0.0484	0.0198
Comp Mole Frac (n-Hexane)	0.0000	0.0612	0.0612	0.0834	0.0016
Comp Mole Frac (n-Heptane)	0.0000	0.0758	0.0758	0.1041	0.0000
Comp Mole Frac (n-Octane)	0.0000	0.0881	0.0881	0.1210	0.0000
Comp Mole Frac (n-Nonane)	0.0000	0.0755	0.0755	0.1037	0.0000
Comp Mole Frac (n-Decane)	0.0000	0.0607	0.0607	0.0834	0.0000
Comp Mole Frac (Mcyclopentan)	0.0000	0.0053	0.0053	0.0073	0.0001
Comp Mole Frac (Benzene)	0.0000	0.0035	0.0035	0.0048	0.0001
Comp Mole Frac (Toluene)	0.0000	0.0062	0.0062	0.0085	0.0000
Comp Mole Frac (Cyclohexane)	0.0000	0.0074	0.0074	0.0102	0.0000
Comp Mole Frac (p-Xylene)	0.0000	0.0329	0.0329	0.0452	0.0000
Comp Mole Frac (Cumene)	0.0000	0.0088	0.0088	0.0121	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0006	0.0006	0.0000	0.0021
Comp Mole Frac (CO2)	0.0002	0.0087	0.0087	0.0000	0.0319
Comp Mole Frac (H2S)	0.0010	0.0106	0.0106	0.0000	0.0389
Comp Mole Frac (H2O)	0.9346	0.0013	0.0013	0.0000	0.0037
Comp Mole Frac (M-Mercaptan)	0.0000	0.0001	0.0001	0.0001	0.0002
Comp Mole Frac (E-Mercaptan)	0.0000	0.0026	0.0026	0.0027	0.0022
Comp Mole Frac (nPMercaptan)	0.0000	0.0024	0.0024	0.0033	0.0001
Comp Mole Frac (nBMercaptan)	0.0000	0.0008	0.0008	0.0011	0.0000
Comp Mole Frac (n-C11)	0.0000	0.1434	0.1434	0.1969	0.0000
Comp Mole Frac (COS)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (EGlycol)	0.0641	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Mcyclohexane)	0.0000	0.0201	0.0201	0.0276	0.0000
Comp Mole Frac (1Pentanthiol)	0.0000	0.0018	0.0018	0.0025	0.0000

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	S15	S19 To Gas Treating	S19	S20	S22
Comp Mole Frac (Methane)	0.5776	0.5776	0.5776	0.1581	0.2180
Comp Mole Frac (Ethane)	0.1441	0.1441	0.1441	0.1079	0.0544
Comp Mole Frac (Propane)	0.1185	0.1185	0.1185	0.1907	0.0518
Comp Mole Frac (i-Butane)	0.0234	0.0234	0.0234	0.0667	0.0189
Comp Mole Frac (n-Butane)	0.0393	0.0393	0.0393	0.1359	0.0389
Comp Mole Frac (i-Pentane)	0.0122	0.0122	0.0122	0.0753	0.0230
Comp Mole Frac (n-Pentane)	0.0112	0.0112	0.0112	0.0795	0.0258
Comp Mole Frac (n-Hexane)	0.0022	0.0022	0.0022	0.0313	0.0380
Comp Mole Frac (n-Heptane)	0.0008	0.0008	0.0008	0.0219	0.0467
Comp Mole Frac (n-Octane)	0.0004	0.0004	0.0004	0.0212	0.0541
Comp Mole Frac (n-Nonane)	0.0001	0.0001	0.0001	0.0152	0.0463
Comp Mole Frac (n-Decane)	0.0001	0.0001	0.0001	0.0104	0.0372
Comp Mole Frac (Mcyclopentan)	0.0002	0.0002	0.0002	0.0022	0.0033
Comp Mole Frac (Benzene)	0.0001	0.0001	0.0001	0.0014	0.0022
Comp Mole Frac (Toluene)	0.0001	0.0001	0.0001	0.0015	0.0038
Comp Mole Frac (Cyclohexane)	0.0001	0.0001	0.0001	0.0024	0.0046
Comp Mole Frac (p-Xylene)	0.0001	0.0001	0.0001	0.0063	0.0202
Comp Mole Frac (Cumene)	0.0000	0.0000	0.0000	0.0016	0.0054
Comp Mole Frac (Nitrogen)	0.0069	0.0069	0.0069	0.0010	0.0026
Comp Mole Frac (CO2)	0.0312	0.0312	0.0312	0.0134	0.0120
Comp Mole Frac (H2S)	0.0259	0.0259	0.0259	0.0210	0.0102
Comp Mole Frac (H2O)	0.0040	0.0040	0.0040	0.0019	0.1293
Comp Mole Frac (M-Mercaptan)	0.0001	0.0001	0.0001	0.0003	0.0001
Comp Mole Frac (E-Mercaptan)	0.0012	0.0012	0.0012	0.0051	0.0017
Comp Mole Frac (nPMercaptan)	0.0001	0.0001	0.0001	0.0011	0.0015
Comp Mole Frac (nBMercaptan)	0.0000	0.0000	0.0000	0.0002	0.0005
Comp Mole Frac (n-C11)	0.0001	0.0001	0.0001	0.0205	0.0878
Comp Mole Frac (COS)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0482
Comp Mole Frac (Mcyclohexane)	0.0002	0.0002	0.0002	0.0056	0.0124
Comp Mole Frac (1Pentanthiol)	0.0000	0.0000	0.0000	0.0004	0.0011

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	S21	CWS	CWR		
Comp Mole Frac (Methane)	0.1581 *	0.0000 *	0.0000		
Comp Mole Frac (Ethane)	0.1079 *	0.0000 *	0.0000		
Comp Mole Frac (Propane)	0.1907 *	0.0000 *	0.0000		
Comp Mole Frac (i-Butane)	0.0667 *	0.0000 *	0.0000		
Comp Mole Frac (n-Butane)	0.1359 *	0.0000 *	0.0000		
Comp Mole Frac (i-Pentane)	0.0753 *	0.0000 *	0.0000		
Comp Mole Frac (n-Pentane)	0.0795 *	0.0000 *	0.0000		
Comp Mole Frac (n-Hexane)	0.0313 *	0.0000 *	0.0000		
Comp Mole Frac (n-Heptane)	0.0219 *	0.0000 *	0.0000		
Comp Mole Frac (n-Octane)	0.0212 *	0.0000 *	0.0000		
Comp Mole Frac (n-Nonane)	0.0152 *	0.0000 *	0.0000		
Comp Mole Frac (n-Decane)	0.0104 *	0.0000 *	0.0000		
Comp Mole Frac (Mcyclopentan)	0.0022 *	0.0000 *	0.0000		
Comp Mole Frac (Benzene)	0.0014 *	0.0000 *	0.0000		
Comp Mole Frac (Toluene)	0.0015 *	0.0000 *	0.0000		
Comp Mole Frac (Cyclohexane)	0.0024 *	0.0000 *	0.0000		
Comp Mole Frac (p-Xylene)	0.0063 *	0.0000 *	0.0000		
Comp Mole Frac (Cumene)	0.0016 *	0.0000 *	0.0000		
Comp Mole Frac (Nitrogen)	0.0010 *	0.0000 *	0.0000		
Comp Mole Frac (CO2)	0.0134 *	0.0000 *	0.0000		
Comp Mole Frac (H2S)	0.0210 *	0.0000 *	0.0000		
Comp Mole Frac (H2O)	0.0019 *	1.0000 *	1.0000		
Comp Mole Frac (M-Mercaptan)	0.0003 *	0.0000 *	0.0000		
Comp Mole Frac (E-Mercaptan)	0.0051 *	0.0000 *	0.0000		
Comp Mole Frac (nPMercaptan)	0.0011 *	0.0000 *	0.0000		
Comp Mole Frac (nBMercaptan)	0.0002 *	0.0000 *	0.0000		
Comp Mole Frac (n-C11)	0.0205 *	0.0000 *	0.0000		
Comp Mole Frac (COS)	0.0000 *	0.0000 *	0.0000		
Comp Mole Frac (EGlycol)	0.0000 *	0.0000 *	0.0000		
Comp Mole Frac (Mcyclohexane)	0.0056 *	0.0000 *	0.0000		
Comp Mole Frac (1Pentanthiol)	0.0004 *	0.0000 *	0.0000		

Energy Streams

Fluid Pkg: All

Name	Q-101	Q-102	Q-100	Q-103	
Heat Flow (kJ/h)	6.142e+007	4.372e+006	3.141e+006	5.253e+006	

Unit Ops

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
Stabiliser Feed Drum	3 Phase Separator	S22	S03 S04 S02	No	500.0 *
Stabiliser Column	Distillation	S16 Q-101	S06 S05 Q-102	No	2500 *
E-100	Heat Exchanger	S07 Fresh Feed	S18 S01	No	500.0 *
E-102	Heat Exchanger	S06 S03	S07 S17	No	500.0 *
E-101	Heat Exchanger	S18 CWS	S11 To Storage Tank CWR	No	500.0 *
K-100	Compressor	S19 Q-103	S19 To Gas Treating Unit	No	500.0 *
K-101	Compressor	S13 Q-100	S12	No	500.0 *
V-100	Separator	S05	S14 To Sour Water Treatment S13	No	500.0 *
V-101	Separator	S15	S20	No	500.0 *

Workbook: Case (Main) (continued)

Unit Ops (continued)

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
V-101	Separator		S19	No	500.0 *
VLV-100	Valve	S17	S16	No	500.0 *
MIX-100	Mixer	S12	S15	No	500.0 *
		S04			
MIX-101	Mixer	S01	S22	No	500.0 *
		S21			
RCY-1	Recycle	S20	S21	No	3500 *

* Specified by user.

APPENDIX II

<u>Data for simulation validation</u>

Component	**Plant Data**		**Pro/II Software**	**HYSYS (ver. 2006) Software**
	mass %	mole fraction	mole fraction	mole fraction
Propane	0.00	0.0000	0.0000	0.0159
i-Butane	1.40	0.0253	0.0143	0.0226
n-Butane	3.80	0.0687	0.0791	0.0540
i-Pentane	4.50	0.0656	0.0814	0.0412
n-Pentane	4.60	0.0670	0.0827	0.0484
Mcyclopentane	0.00	0.0000	0.0078	0.0073
Benzene	2.00	0.0269	0.0052	0.0048
n-Hexane	10.00	0.1220	0.0940	0.0834
Cyclohexane	0.00	0.0000	0.0104	0.0102
Mcyclohexane	0.00	0.0000	0.0261	0.0276
Toluene	2.50	0.0285	0.0078	0.0085
n-Heptane	13.40	0.1406	0.0992	0.1041
n-Octane	14.40	0.1325	0.1071	0.1210
p-Xylene	3.00	0.0297	0.0392	0.0452
n-Nonane	9.10	0.0746	0.0888	0.1037
Cumene	0.00	0.0000	0.0105	0.0121
n-Decane	9.70	0.0717	0.0705	0.0834
C11+	21.14	0.1422	0.1645	0.1969
M-Mercaptan	0.00	0.0000	0.0000	0.0001
E-Mercaptan	0.00	0.0000	0.0045	0.0027
nPMercaptan	0.00	0.0000	0.0036	0.0033
nBMercaptan	0.00	0.0000	0.0011	0.0011
1Pentanthiol	0.00	0.0000	0.0022	0.0025
Ethylbenzene	0.46	0.0046	0.0000	0.0000
	100.0000	**1.0000**	**1.0000**	**1.0000**

APPENDIX III

Effect of Feed Flow Rate

Normal Flow = 4645 kgmole/hr

MW of S = 32.065 kg/kgmole

Feed Flow (kgmole/hr)	3251.5	3716	4180.5	4645	5109.5	5574	6038.5
Flow %	70.0	80.0	90.0	100.0	110.0	120.0	130.0
RVP (psia)	0.8683	2.621	5.436	8.779	12.11	15.41	18.48
Components							
Methane	3.73E-12	7.9E-11	9.63E-10	2.91E-08	2.6E-07	2.06E-05	0.000119
Ethane	9.08E-09	1.32E-06	5.08E-05	0.005243	0.085741	9.825171	30.09459
Propane	7.18E-06	0.004174	0.346877	32.92544	87.56977	120.264	142.9469
i-Butane	0.000389	0.564932	22.37664	46.84751	58.51905	68.85246	78.38199
n-Butane	0.003906	7.106194	77.6815	111.7624	135.2987	156.9184	177.0547
i-Pentane	0.313236	50.84607	71.00131	85.37574	99.08917	112.7502	125.5437
n-Pentane	0.947595	64.18872	84.51479	100.278	116.5463	133.292	147.9786
n-Hexane	67.78512	122.8129	149.9891	172.7023	191.58	209.231	226.8507
n-Heptane	131.6056	170.7307	193.793	215.566	237.2925	258.9987	280.6911
n-Octane	171.5955	200.2298	225.4466	250.6243	275.7801	300.9248	326.0621
n-Nonane	150.1526	171.7417	193.2846	214.8126	236.332	257.8469	279.3587
n-Decane	120.8136	138.1206	155.4146	172.7031	189.9883	207.2717	224.5539
Mcyclopentane	6.672159	10.80308	13.19229	15.05909	16.65205	18.18457	19.71472
Benzene	4.439484	7.222873	8.785299	10.03419	11.10643	12.12833	13.14853
Toluene	11.04579	13.98531	15.78736	17.55803	19.32553	21.09161	22.8567
Cyclohexane	10.88852	15.53974	18.82302	21.10892	23.26134	25.39767	27.53155
p-Xylene	65.15207	74.83662	84.24518	93.6424	103.0331	112.4207	121.806
Cumene	17.51084	20.02927	22.54196	25.0528	27.56258	30.07182	32.58069
Nitrogen	4.14E-17	2.5E-16	1.4E-15	1.85E-14	1.07E-13	4.55E-12	1.98E-11
CO2	1.17E-11	8.06E-10	2.07E-08	1.5E-06	2.1E-05	0.003307	0.024978
H2S	9.65E-09	1.9E-06	8.98E-05	0.011667	0.204799	7.517511	11.74459
H2O	2.01E-11	1.08E-09	2.13E-08	1.25E-06	1.52E-05	0.002069	0.012963
M-Mercaptan	8.03E-07	0.001016	0.067222	0.220907	0.282031	0.334386	0.382301
E-Mercaptan	0.002215	2.115335	4.529691	5.67819	6.674673	7.615042	8.497779
nPMercaptan	2.438149	4.82856	5.899473	6.804589	7.557377	8.254294	8.949857
nBMercaptan	1.372238	1.80769	2.072599	2.305933	2.538578	2.770972	3.003196
n-C11	285.3296	326.143	366.9429	407.7368	448.5272	489.3154	530.1023
COS	0	0	0	0	0	0	0
Eglycol	2.88E-09	1.04E-06	5.38E-05	0.005669	0.022612	0.031156	0.033179
Mcyclohexane	35.38544	45.36577	51.44959	57.22968	62.9971	68.75916	74.51756
1Pentanthiol	3.470563	4.070202	4.583059	5.095044	5.606542	6.1178	6.628896
Total (kgmole/hr)	7.283167	12.82281	17.15213	20.11633	22.864	32.61	39.20662
Total Mass of S (kg/hr)	233.5347	411.1633	549.9832	645.0301	733.1342	1045.64	1257.16
Total Product Mass Flow (kg/hr)	135781	171933.9	202350.3	230677.7	258203.4	285005.5	311223.1
Sulphur Concentration (ppm)	1719.937	2391.404	2717.976	2796.24	2839.367	3668.841	4039.418

Effect of Feed Temperature

Normal Feed Temperature =	21.4	degC
MW of S =	32.065	kg/kgmole

Feed Temperature (degC)	2	4	6	8	10	12	14
RVP	12.56	12.14	11.73	11.33	10.93	10.54	10.15
Components							
Methane	5.41E-08	4.94E-08	4.59E-08	4.30E-08	4.05E-08	3.84E-08	3.65E-08
Ethane	2.72E-02	2.23E-02	1.86E-02	1.56E-02	1.33E-02	1.13E-02	9.71E-03
Propane	81.94666	76.19079	70.65169	65.24972	60.0027	54.94576	50.0435
i-Butane	58.18197	57.04226	55.89027	54.71941	53.53232	52.33391	51.12405
n-Butane	133.3431	131.216	129.046	126.8223	124.5497	122.2365	119.8859
i-Pentane	96.57848	95.40835	94.22077	93.00105	91.74297	90.44476	89.11264
n-Pentane	114.5	113.2661	111.858	110.2843	108.5609	106.7072	104.7601
n-Hexane	175.2443	175.1401	175.0289	174.9103	174.7836	174.6471	174.4951
n-Heptane	216.3175	216.2617	216.2016	216.1369	216.0675	215.993	215.9132
n-Octane	251.0206	250.9925	250.9619	250.9287	250.8927	250.8538	250.8116
n-Nonane	214.9689	214.9583	214.9466	214.9338	214.9199	214.9047	214.888
n-Decane	172.7622	172.7584	172.7541	172.7494	172.7442	172.7385	172.7322
Mcyclopentane	15.22557	15.21713	15.20813	15.19854	15.18831	15.17734	15.16534
Benzene	10.1533	10.14787	10.14206	10.13587	10.12924	10.12209	10.11414
Toluene	17.60981	17.60598	17.60185	17.5974	17.59263	17.5875	17.582
Cyclohexane	21.24866	21.23883	21.22834	21.21715	21.20524	21.19254	21.17898
p-Xylene	93.75423	93.74638	93.73782	93.72852	93.71841	93.70744	93.69554
Cumene	25.07141	25.07013	25.06873	25.06721	25.06554	25.06372	25.06174
Nitrogen	1.67E-14	1.64E-14	1.65E-14	1.66E-14	1.69E-14	1.72E-14	1.76E-14
CO2	5.63E-06	4.77E-06	4.11E-06	3.58E-06	3.14E-06	2.77E-06	2.46E-06
H2S	6.67E-02	5.40E-02	4.46E-02	3.72E-02	3.13E-02	2.65E-02	2.25E-02
H2O	1.95E-06	1.82E-06	1.72E-06	1.64E-06	1.58E-06	1.53E-06	1.48E-06
M-Mercaptan	0.281666	0.275496	0.269296	0.263025	0.256698	0.250338	0.243936
E-Mercaptan	6.481529	6.404613	6.325281	6.243111	6.158165	6.070637	5.980686
nPMercaptan	6.915183	6.91087	6.906265	6.901346	6.89608	6.89039	6.883982
nBMercaptan	2.315047	2.314366	2.313635	2.312849	2.312006	2.311102	2.310135
n-C11	407.8007	407.7967	407.7923	407.7873	407.7818	407.7757	407.7689
COS	0	0	0	0	0	0	0
Eglycol	7.69E-03	7.62E-03	7.53E-03	7.41E-03	7.26E-03	7.09E-03	6.88E-03
Mcyclohexane	57.42893	57.41391	57.3978	57.38052	57.36203	57.34224	57.32109
1Pentanthiol	5.10355	5.102944	5.102286	5.101572	5.100799	5.099962	5.099058
Total (kgmole/hr)	21.16366	21.06225	20.96132	20.85909	20.75503	20.64891	20.54026
Total Mass of S (kg/hr)	678.6127	675.3611	672.1247	668.8467	665.51	662.1072	658.6234
Total Product Mass Flow (kg/hr)	237104.2	236455.8	235798	235125.6	234440.3	233745.5	233042.1
Sulphur Concentration (ppm)	2862.086	2856.183	2850.425	2844.636	2838.718	2832.599	2826.199

Effect of Feed Temperature (continued)

16	18	20	21.4	22	24	26	28	30
9.766	9.395	9.031	8.779	8.672	8.314	7.961	7.612	7.268
3.47E-08	3.27E-08	3.06E-08	2.91E-08	2.85E-08	2.62E-08	2.39E-08	2.13E-08	1.87E-08
8.31E-03	7.04E-03	5.93E-03	5.24E-03	4.96E-03	4.09E-03	3.33E-03	2.65E-03	2.06E-03
45.27807	40.60129	36.03775	32.92544	31.60915	27.25474	23.0643	18.98421	15.09109
49.92112	48.7778	47.64567	46.84751	46.50747	45.35979	44.18431	42.98545	41.73848
117.5518	115.3897	113.2638	111.7624	111.1235	108.9678	106.7564	104.516	102.2082
87.83463	86.87077	85.99516	85.37574	85.11088	84.21007	83.27105	82.30729	81.30196
102.9346	101.8134	100.9089	100.278	100.008	99.08894	98.12772	97.1351	96.09449
174.2754	173.811	173.1954	172.7023	172.4704	171.6236	170.6712	169.5828	168.3731
215.8277	215.7363	215.6385	215.566	215.5329	215.417	215.2954	215.1624	215.0224
250.7661	250.7169	250.6639	250.6243	250.6063	250.543	250.4758	250.4021	250.3237
214.8699	214.8502	214.8287	214.8126	214.8053	214.7795	214.7517	214.7212	214.6885
172.7252	172.7177	172.7094	172.7031	172.7002	172.6902	172.6792	172.6671	172.654
15.14976	15.12252	15.08749	15.05909	15.04551	14.99472	14.93525	14.86369	14.77979
10.10304	10.082	10.05536	10.03419	10.0242	9.987484	9.945603	9.896589	9.840641
17.5761	17.56979	17.56303	17.55803	17.55573	17.5477	17.53927	17.53005	17.52034
21.16401	21.14593	21.12514	21.10892	21.10121	21.07275	21.03986	20.9994	20.95012
93.68266	93.66872	93.65366	93.6424	93.63727	93.61918	93.59994	93.57881	93.55628
25.05959	25.05725	25.05471	25.0528	25.05193	25.04887	25.04559	25.04199	25.03812
1.80E-14	1.82E-14	1.84E-14	1.85E-14	1.85E-14	1.83E-14	1.81E-14	1.76E-14	1.68E-14
2.17E-06	1.90E-06	1.66E-06	1.50E-06	1.43E-06	1.22E-06	1.03E-06	8.49E-07	6.84E-07
1.90E-02	1.60E-02	1.33E-02	1.17E-02	1.10E-02	8.97E-03	7.21E-03	5.66E-03	4.32E-03
1.43E-06	1.37E-06	1.30E-06	1.25E-06	1.23E-06	1.15E-06	1.05E-06	9.47E-07	8.30E-07
0.237555	0.231384	0.225241	0.220907	0.219057	0.212795	0.206361	0.199724	0.192707
5.891197	5.811201	5.73337	5.67819	5.654668	5.575048	5.492802	5.409183	5.32275
6.8743	6.853326	6.826069	6.804589	6.794588	6.7585	6.718588	6.673744	6.624623
2.3091	2.307993	2.306809	2.305933	2.305531	2.304126	2.302651	2.301033	2.299325
407.7613	407.753	407.7438	407.7368	407.7336	407.7223	407.7099	407.6962	407.6812
0	0	0	0	0	0	0	0	0
6.63E-03	6.32E-03	5.95E-03	5.67E-03	5.54E-03	5.06E-03	4.53E-03	3.95E-03	3.33E-03
57.2985	57.27439	57.24868	57.22968	57.22097	57.19054	57.15874	57.12396	57.08746
5.098081	5.097027	5.095891	5.095044	5.094658	5.0933	5.091859	5.09028	5.088602
20.42926	20.3169	20.20069	20.11633	20.07951	19.95274	19.81947	19.67963	19.53233
655.0641	651.4614	647.7351	645.0301	643.8495	639.7846	635.5115	631.0272	626.3041
232350.5	231724.6	231111.8	230677.7	230490.3	229852.6	229198.7	228524.8	227830.5
2819.293	2811.361	2802.691	2796.24	2793.391	2783.455	2772.754	2761.307	2748.992

Effect of Feed Pressure

Normal Feed Pressure = 30.7 barg

MW of S = 32.065 kg/kgmole

Feed Pressure (barg)	20	22	24	26	28	30	30.7
RVP	8.116	8.257	8.389	8.513	8.63	8.742	8.779
Components							
Methane	1.54E-08	1.76E-08	2.00E-08	2.25E-08	2.52E-08	2.81E-08	2.91E-08
Ethane	2.73E-03	3.13E-03	3.57E-03	4.03E-03	4.52E-03	5.05E-03	5.24E-03
Propane	22.64632	24.68441	26.66754	28.60599	30.47243	32.31322	32.92544
i-Butane	46.34022	46.57377	46.72515	46.81452	46.85309	46.85565	46.84751
n-Butane	111.5642	111.8578	111.997	112.0216	111.9548	111.8235	111.7624
i-Pentane	85.50588	85.58	85.5941	85.56413	85.49997	85.41177	85.37574
n-Pentane	100.3071	100.3939	100.4243	100.4132	100.3702	100.3053	100.278
n-Hexane	169.8082	170.451	171.0397	171.582	172.0831	172.5482	172.7023
n-Heptane	214.9103	215.0775	215.2183	215.3389	215.4438	215.5363	215.566
n-Octane	250.3413	250.4134	250.4741	250.5261	250.5715	250.6114	250.6243
n-Nonane	214.7214	214.7447	214.7643	214.781	214.7956	214.8085	214.8126
n-Decane	172.6752	172.6824	172.6884	172.6935	172.6979	172.7018	172.7031
Mcyclopentane	14.85087	14.90093	14.94485	14.98366	15.01817	15.04907	15.05909
Benzene	9.892124	9.925205	9.954708	9.981224	10.0052	10.02703	10.03419
Toluene	17.5103	17.52248	17.53274	17.54151	17.54915	17.55586	17.55803
Cyclohexane	20.94137	20.98486	21.02119	21.052	21.07848	21.10155	21.10892
p-Xylene	93.56	93.58097	93.59866	93.61381	93.62701	93.63865	93.6424
Cumene	25.04069	25.04378	25.04638	25.0486	25.05054	25.05225	25.0528
Nitrogen	1.03E-14	1.16E-14	1.30E-14	1.45E-14	1.61E-14	1.78E-14	1.85E-14
CO2	7.58E-07	8.75E-07	1.00E-06	1.14E-06	1.28E-06	1.44E-06	1.50E-06
H2S	6.13E-03	7.02E-03	7.98E-03	9.00E-03	1.01E-02	1.13E-02	1.17E-02
H2O	8.63E-07	9.28E-07	9.96E-07	1.07E-06	1.14E-06	1.23E-06	1.25E-06
M-Mercaptan	0.216623	0.218144	0.219244	0.220022	0.22053	0.220847	0.220907
E-Mercaptan	5.688164	5.695276	5.696937	5.694587	5.689094	5.681384	5.67819
nPMercaptan	6.687418	6.712832	6.736374	6.758322	6.778853	6.798147	6.804589
nBMercaptan	2.297606	2.299734	2.301524	2.303055	2.304386	2.305556	2.305933
n-C11	407.7132	407.7193	407.7244	407.7288	407.7325	407.7358	407.7368
COS	0	0	0	0	0	0	0
Eglycol	3.59E-03	3.95E-03	4.32E-03	4.71E-03	5.11E-03	5.52E-03	5.67E-03
Mcyclohexane	57.04468	57.09191	57.13167	57.16569	57.19526	57.22129	57.22968
1Pentanthiol	5.088554	5.090209	5.091603	5.092796	5.093834	5.094749	5.095044
Total (kgmole/hr)	19.98449	20.02322	20.05366	20.07779	20.09678	20.11194	20.11633
Total Mass of S (kg/hr)	640.8028	642.0445	643.0207	643.7942	644.4033	644.8893	645.0301
Total Product Mass Flow (kg/hr)	229742.3	229981.5	230183.4	230356.7	230505.4	230636.8	230677.7
Sulphur Concentration (ppm)	2789.224	2791.723	2793.515	2794.771	2795.611	2796.125	2796.24

Effect of Feed Pressure (continued)

32	34	36	38	40	42	44	46	48	50
8.812	8.862	8.909	8.953	8.995	9.034	9.072	9.108	9.142	9.175
[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
2.93E-08	2.96E-08	2.99E-08	3.02E-08	3.04E-08	3.06E-08	3.08E-08	3.10E-08	3.12E-08	3.14E-08
5.33E-03	5.46E-03	5.59E-03	5.71E-03	5.83E-03	5.94E-03	6.05E-03	6.16E-03	6.26E-03	6.35E-03
33.32982	33.95194	34.52349	35.07559	35.59521	36.07646	36.54851	36.99665	37.4278	37.83475
46.95489	47.11371	47.26166	47.40198	47.53409	47.65792	47.77674	47.88948	47.99704	48.09902
111.9649	112.2635	112.542	112.8058	113.0541	113.2873	113.5105	113.7223	113.9243	114.1159
85.4598	85.58304	85.69801	85.80666	85.90891	86.00502	86.09677	86.18384	86.26677	86.34556
100.3635	100.4889	100.606	100.7166	100.8209	100.919	101.0127	101.1018	101.1867	101.2676
172.7718	172.8721	172.9643	173.0499	173.1294	173.2031	173.2724	173.3372	173.3979	173.455
215.576	215.5905	215.604	215.6166	215.6285	215.6397	215.6502	215.6601	215.6695	215.6785
250.6298	250.6377	250.6451	250.652	250.6585	250.6645	250.6703	250.6757	250.6808	250.6856
214.8149	214.8181	214.8211	214.8239	214.8265	214.829	214.8313	214.8335	214.8356	214.8375
172.704	172.7052	172.7064	172.7075	172.7085	172.7095	172.7104	172.7112	172.712	172.7128
15.06313	15.06892	15.07424	15.07916	15.08371	15.08794	15.09189	15.09558	15.09904	15.10229
10.03717	10.04148	10.04543	10.04911	10.05252	10.05569	10.05867	10.06145	10.06407	10.06653
17.55872	17.55972	17.56065	17.56153	17.56234	17.56312	17.56384	17.56453	17.56518	17.5658
21.1112	21.11449	21.11752	21.12034	21.12296	21.1254	21.1277	21.12985	21.13188	21.13379
93.64396	93.64622	93.64832	93.65028	93.65212	93.65385	93.65547	93.65701	93.65847	93.65984
25.05306	25.05345	25.0538	25.05413	25.05445	25.05474	25.05501	25.05527	25.05552	25.05575
1.84E-14	1.84E-14	1.84E-14	1.84E-14	1.84E-14	1.84E-14	1.84E-14	1.84E-14	1.83E-14	1.83E-14
1.52E-06	1.55E-06	1.58E-06	1.61E-06	1.63E-06	1.66E-06	1.68E-06	1.71E-06	1.73E-06	1.75E-06
1.19E-02	1.22E-02	1.25E-02	1.28E-02	1.31E-02	1.33E-02	1.36E-02	1.38E-02	1.41E-02	1.43E-02
1.26E-06	1.27E-06	1.28E-06	1.29E-06	1.30E-06	1.30E-06	1.31E-06	1.32E-06	1.33E-06	1.33E-06
0.22149	0.222353	0.223156	0.223918	0.224635	0.225306	0.225951	0.226563	0.227147	0.227701
5.68566	5.696635	5.706876	5.716561	5.725678	5.734243	5.742427	5.750191	5.757587	5.764607
6.807601	6.811955	6.815965	6.819701	6.823175	6.826407	6.829446	6.832297	6.834975	6.837494
2.306054	2.306229	2.306392	2.306545	2.306689	2.306824	2.306951	2.307071	2.307185	2.307293
407.7378	407.7392	407.7405	407.7417	407.7429	407.7439	407.7449	407.7459	407.7468	407.7476
0	0	0	0	0	0	0	0	0	0
5.71E-03	5.77E-03	5.82E-03	5.87E-03	5.91E-03	5.95E-03	5.99E-03	6.03E-03	6.07E-03	6.10E-03
57.2323	57.23609	57.23964	57.24295	57.24606	57.24899	57.25176	57.25437	57.25684	57.25918
5.095162	5.095331	5.095489	5.095637	5.095775	5.095905	5.096028	5.096144	5.096253	5.096357
20.12784	20.1447	20.16037	20.17515	20.18902	20.20201	20.21439	20.22611	20.23723	20.24777
645.3991	645.9397	646.4422	646.9161	647.3609	647.7775	648.1745	648.5501	648.9068	649.2446
230736	230822.7	230903.1	230979.5	231051.2	231118.2	231182.7	231243.8	231302	231357.2
2797.132	2798.423	2799.626	2800.752	2801.807	2802.797	2803.733	2804.617	2805.452	2806.244

Effect of Feed Pressure (continued)

55	60	65	70	75	80	85	90	95	100
9.252	9.32	9.379	9.399	9.364	9.329	9.292	9.257	9.22	9.184
3.188E-08	3.23E-08	3.26E-08	3.27E-08	3.251E-08	3.23E-08	3.21E-08	3.19E-08	3.17E-08	3.15E-08
0.0065935	6.80E-03	6.99E-03	7.05E-03	0.006943	0.006835	0.006716	6.61E-03	6.50E-03	6.39E-03
38.805062	39.64992	40.3941	40.64232	40.204001	39.76729	39.30097	38.86339	38.40928	37.95445
48.336045	48.54419	48.72763	48.78914	48.679971	48.57074	48.45826	48.34788	48.23541	48.12226
114.56029	114.951	115.2955	115.4112	115.20591	115.0005	114.7897	114.582	114.3709	114.1584
86.5279	86.689	86.83164	86.87982	86.794402	86.70914	86.62241	86.53661	86.44973	86.36239
101.45542	101.6226	101.7721	101.823	101.73289	101.6436	101.5534	101.4644	101.3747	101.2848
173.58317	173.6928	173.7862	173.8168	173.76204	173.7059	173.648	173.589	173.5284	173.4664
215.69877	215.7166	215.7321	215.7373	215.72803	215.7187	215.7092	215.6997	215.69	215.6802
250.69664	250.7063	250.7146	250.7174	250.71247	250.7074	250.7023	250.6971	250.6919	250.6866
214.842	214.8459	214.8493	214.8504	214.84839	214.8464	214.8443	214.8422	214.8401	214.8379
172.71451	172.716	172.7173	172.7178	172.71698	172.7162	172.7154	172.7146	172.7138	172.7129
15.10957	15.1158	15.12111	15.12285	15.119733	15.11654	15.11325	15.1099	15.10646	15.10293
10.072077	10.07684	10.08091	10.08226	10.079859	10.07741	10.07489	10.07233	10.0697	10.06703
17.567199	17.56843	17.5695	17.56986	17.569221	17.56858	17.56792	17.56726	17.56659	17.56592
21.138108	21.14184	21.14507	21.14614	21.144231	21.14229	21.14031	21.1383	21.13625	21.13416
93.66297	93.6657	93.66808	93.66888	93.667465	93.66603	93.66458	93.66311	93.66162	93.66011
25.056278	25.05674	25.05714	25.05727	25.057035	25.05679	25.05655	25.0563	25.05605	25.0558
1.831E-14	1.83E-14	1.82E-14	1.82E-14	1.826E-14	1.83E-14	1.83E-14	1.83E-14	1.83E-14	1.84E-14
1.803E-06	1.85E-06	1.89E-06	1.90E-06	1.88E-06	1.86E-06	1.83E-06	1.81E-06	1.78E-06	1.76E-06
0.0148868	1.54E-02	1.58E-02	1.60E-02	0.0157268	0.015465	0.015181	1.49E-02	1.47E-02	1.44E-02
1.347E-06	1.36E-06	1.37E-06	1.37E-06	1.368E-06	1.36E-06	1.35E-06	1.35E-06	1.34E-06	1.34E-06
0.2289881	0.230117	0.231112	0.231445	0.230854	0.230262	0.229651	0.229053	0.228443	0.227829
5.7808505	5.795148	5.807754	5.811993	5.804472	5.796942	5.789245	5.781633	5.773906	5.766128
6.8431709	6.848046	6.852213	6.853585	6.8511355	6.84863	6.846051	6.843428	6.840742	6.837998
2.307539	2.307754	2.307942	2.308005	2.3078931	2.30778	2.307666	2.30755	2.307433	2.307314
407.74952	407.7512	407.7526	407.7531	407.75225	407.7514	407.7505	407.7496	407.7487	407.7478
0	0	0	0	0	0	0	0	0	0
0.0061819	6.25E-03	6.30E-03	6.32E-03	0.0062891	0.006259	0.00622	6.19E-03	6.15E-03	6.12E-03
57.264527	57.26921	57.27329	57.27465	57.272227	57.26977	57.26728	57.26476	57.26222	57.25964
5.0965927	5.096799	5.096979	5.097039	5.0969319	5.096824	5.096714	5.096603	5.096491	5.096377
20.272028	20.29326	20.31184	20.31806	20.307013	20.2959	20.28451	20.27319	20.26167	20.25004
650.02258	650.7032	651.2992	651.4986	651.14438	650.7881	650.4227	650.0599	649.6905	649.3176
231358.19	231598.4	231697.4	231730.5	231671.76	231612.9	231551.9	231492.5	231431.6	231370.5
2809.594	2809.619	2810.991	2811.449	2810.6334	2809.809	2808.972	2808.126	2807.267	2806.398

Effect of Reboiler Temperature

Normal Reboiler T = 180 degC
MW of S = 32.065 kg/kgmole

Reboiler Temperature (degC)	100	120	140	160	165
RVP	28.95	24.68	20.21	14.23	12.7
Components					
Methane	27.39288	14.55338	1.119821	3.58E-05	4.74E-06
Ethane	47.54329	42.42374	37.04326	8.050544	1.476783
Propane	106.7536	102.5078	98.05365	87.37753	83.85913
i-Butane	56.71864	55.6458	54.52043	52.08955	51.38485
n-Butane	128.0393	126.1831	124.2359	120.0803	118.8881
i-Pentane	90.40533	89.81721	89.20006	87.85669	87.4795
n-Pentane	104.9916	104.4612	103.9043	102.62	102.2653
n-Hexane	173.789	173.7827	173.7757	173.6163	173.5567
n-Heptane	215.5662	215.5662	215.5662	215.5662	215.5661
n-Octane	250.6243	250.6243	250.6243	250.6243	250.6243
n-Nonane	214.8126	214.8126	214.8126	214.8126	214.8126
n-Decane	172.7031	172.7031	172.7031	172.7031	172.7031
Mcyclopentane	15.11165	15.11134	15.11101	15.10359	15.10082
Benzene	10.07813	10.07785	10.07754	10.07107	10.06872
Toluene	17.55803	17.55803	17.55803	17.55803	17.55803
Cyclohexane	21.12066	21.1206	21.12053	21.11907	21.1185
p-Xylene	93.6424	93.6424	93.6424	93.6424	93.6424
Cumene	25.0528	25.0528	25.0528	25.0528	25.0528
Nitrogen	0.036828	0.000336	3.81E-07	1.21E-11	1.85E-12
CO2	4.787583	3.798403	2.718757	0.003672	0.00042
H2S	11.35112	10.41972	9.441924	5.46324	2.430773
H2O	0.842641	0.693682	0.534768	0.002915	0.000353
M-Mercaptan	0.276911	0.271088	0.264979	0.251688	0.247819
E-Mercaptan	6.183058	6.12394	6.061925	5.930418	5.893179
nPMercaptan	6.8539	6.853587	6.853245	6.845778	6.843049
nBMercaptan	2.305939	2.305939	2.305939	2.305938	2.305938
n-C11	407.7368	407.7368	407.7368	407.7368	407.7368
COS	0	0	0	0	0
Eglycol	0.038673	0.036912	0.035064	0.030475	0.028872
Mcyclohexane	57.22972	57.22972	57.22972	57.22971	57.22971
1Pentanthiol	5.095044	5.095044	5.095044	5.095044	5.095044
Total (kgmole/hr)	32.06598	31.06932	30.02306	25.89211	22.8158
Total Mass of S (kg/hr)	1028.196	996.2376	962.6893	830.2304	731.5887
Total Product Mass Flow (kg/hr)	238786.5	237899	236973.4	234751	234123.2
Sulfur Concentration (ppm)	4305.92	4187.649	4062.437	3536.642	3124.802

Effect of Reboiler Temperature (continued)

170	175	180	200	220	240	260	280	300
11.31	10.03	8.779	4.803	2.251	0.9166	0.3918	<empty>	<empty>
[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
2.54E-07	7.34E-08	2.91E-08	2.7E-10	1.87E-11	2.17E-12	4.54E-13	1.56E-13	1.25E-13
0.060733	0.015161	0.005243	1.46E-05	2.73E-07	7.45E-09	5.07E-10	7.4E-11	2.53E-11
70.06354	51.16541	32.92544	0.102055	0.000768	7.26E-06	2.17E-07	1.71E-08	3.21E-09
50.05943	48.44603	46.84751	13.54194	0.098649	0.000446	7.44E-06	3.75E-07	4.32E-08
116.7663	114.2323	111.7624	72.99772	1.351421	0.004693	6.22E-05	2.64E-06	2.55E-07
86.84122	86.09572	85.37574	80.19703	51.95317	0.425387	0.00322	8.71E-05	5.37E-06
101.6565	100.9528	100.278	95.53204	76.39034	1.345516	0.008565	0.000202	1.09E-05
173.3457	173.0317	172.7023	167.6693	157.5518	109.9438	2.055588	0.028756	0.000884
215.5661	215.5661	215.566	215.3656	214.6789	199.084	100.3871	2.53776	0.048111
250.6243	250.6243	250.6243	250.5445	250.4992	249.6966	221.5831	87.4339	2.15037
214.8126	214.8126	214.8126	214.7891	214.7758	214.7429	210.1759	181.709	47.89974
172.7031	172.7031	172.7031	172.6968	172.6932	172.6844	172.6166	165.4489	136.8788
15.09088	15.07565	15.05909	14.71787	13.83845	10.69867	0.335642	0.004694	0.000134
10.06031	10.04766	10.03419	9.802654	9.244439	7.101333	0.236156	0.00315	8.27E-05
17.55803	17.55803	17.55803	17.54347	17.52409	16.5809	11.1546	0.542924	0.009635
21.1164	21.11295	21.10892	20.93557	19.85271	17.08169	1.526944	0.021441	0.00053
93.6424	93.6424	93.6424	93.61901	93.6058	93.54528	86.95718	57.3968	2.95143
25.0528	25.0528	25.0528	25.04953	25.04768	25.04314	24.53599	21.39496	6.330732
1.3E-13	4.2E-14	1.85E-14	3.91E-16	6.41E-17	1.87E-17	8.84E-18	5.97E-18	9.36E-18
1.65E-05	4.21E-06	1.5E-06	5.72E-09	1.69E-10	8.25E-12	9.37E-13	2.13E-13	1.11E-13
0.139878	0.034426	0.011667	2.5E-05	3.56E-07	8.02E-09	4.82E-10	6.94E-11	1.96E-11
1.37E-05	3.51E-06	1.25E-06	5.52E-09	1.73E-10	9.2E-12	1E-12	2.31E-13	1.13E-13
0.240149	0.230594	0.220907	0.029434	0.000163	8.71E-07	1.71E-08	1.01E-09	1.25E-10
5.828721	5.752303	5.67819	5.02052	0.819287	0.002714	2.89E-05	1.08E-06	8.28E-08
6.833416	6.81925	6.804589	6.595936	6.195071	3.969914	0.054188	0.000758	2.32E-05
2.305937	2.305935	2.305933	2.30326	2.284164	2.087687	0.794781	0.015486	0.000307
407.7368	407.7368	407.7368	407.7327	407.7302	407.7238	407.7158	406.2985	386.4849
0	0	0	0	0	0	0	0	0
0.020356	0.011391	0.005669	1.28E-05	1.19E-07	1.36E-09	4.13E-11	3.87E-12	8.19E-13
57.22971	57.22969	57.22968	57.17217	56.99768	53.2618	34.17619	1.615608	0.03083
5.095044	5.095044	5.095044	5.093178	5.092114	5.067857	4.458043	1.629701	0.037022
20.44315	20.23755	20.11633	19.04235	14.3908	11.12817	5.307041	1.645946	0.037352
655.5095	648.9171	645.0301	610.5931	461.441	356.8249	170.1703	52.77724	1.197681
233078.3	231857.9	230677.7	223719.1	213877.4	197531.5	168078.3	129648.6	87362.83
2812.4	2798.77	2796.24	2729.284	2157.502	1806.42	1012.447	407.0793	13.70928