DESIGN OF STABILIZER AT DISTILLATION UNIT FOR PRODUCTION OF LPG

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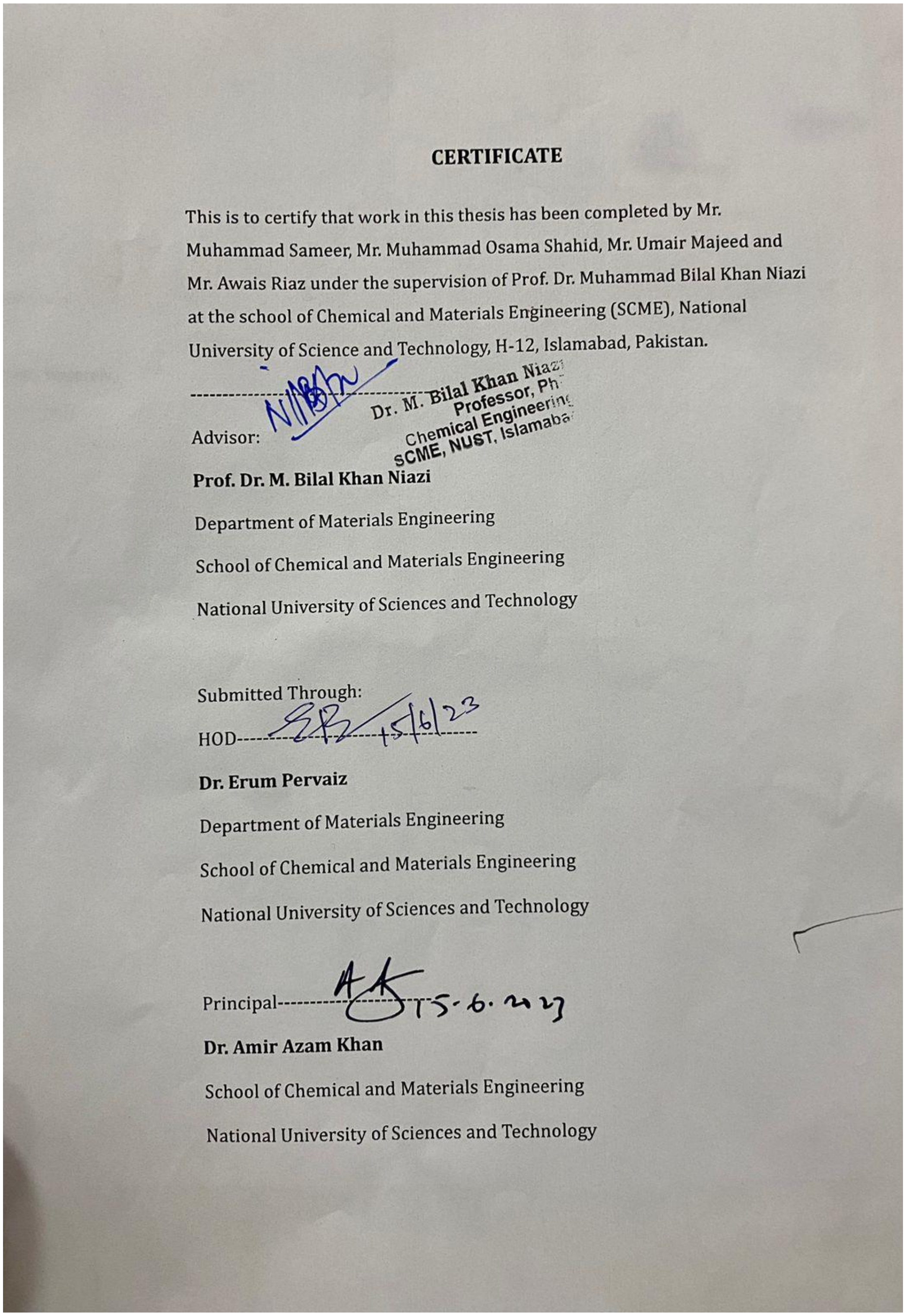

CERTIFICATE

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DEDICATION

To our families, teachers and friends for their continual support and help throughout this journey and without whom none of this would have been possible. And to SCME, the place that equipped us with everything that was needed to reach this position.

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ABSTRACT

Modern day crude oil refining at a massive scale is the backbone of the global economy. This process provides fuel for transportation, domestic use and industrial heating uses and provides feedstock for several industries like the petrochemical industry. Millions of barrels of crude oil are refined every day across the globe to keep the world moving and advancing forward.

The crux of the process lies in the distillation unit of every refinery. This unit is responsible for separating the crude oil into the different fractional streams based on their different volatilities.

In a refinery, the crude oil must first be pretreated to remove corrosion and foul-causing components such as salt and water. Environmentally harmful chemicals like Sulfur must also be removed if the crude oil is sour in nature. The pretreatment process is followed by heating to raise its temperature and decrease the viscosity of the oil before it enters the distillation unit to be separated into the various compounds.

Despite the rigorous designing of the distillation unit, it is very difficult to produce separate components effectively with high purity. This is also the case with the NAPHTHA stream of the distillation column which contains lighter valuable compounds like LPG which become increasingly difficult and expensive to separate due to the high vapor pressure of the stream. This results in the loss of valuable LPG and a less pure NAPHTHA stream which is difficult to transport or use in feedstock.

The goal is to design a stabilizer which separately extracts the LPG from the NAPHTHA stream and stabilizes it in the process. This allows us to create an additional valuable stream of LPG for commercialization and utilize a stable NAPHTHA stream for the production of gasoline and petrochemical products.

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CHAPTER 1

INTRODUCTION

Attock Oil refinery is a plant in the heart of Rawalpindi that has been running and operating for over a century now. It is in Morgah, Rawalpindi and is a subsidiary of the Attock companies.

The primary purpose of the refinery is to separate the different components of crude oil and refine them into commercially viable products. These products range from LPG, gasoline, and diesel all the way up to jet fuel, and lubricants.

The company is public and shares its profits with its investors on the PSX. Recent times have been shaky in terms of profits, but in general the company has been very profitable in its business.

We have been tasked to design a stabilizer tower for the refinery. This tower will stabilize the NAPHTHA to give us a new product stream of LPG. The need for this tower is immense since it doesn't only separate LPG, but the process of stabilizing NAPHTHA makes it very suitable for further processing into gasoline.

1.1 Crude Oil

When deceased living organisms, including animals and plants, are buried underground for thousands or millions of years under the high pressure of the earth around them, they form what we today call crude oil. It is a mixture of hydrocarbons with little amounts of oxygen, nitrogen, and sulfur. The crude oil industry began in the 19th century when the world was progressing with industrialization and our energy needs were climbing exponentially. New machinery, new industries were being set up every day which required more energy to operate than we had ever needed before. Another increase in energy demand was the advancement of the transportation industry. Cars, trains, and planes all required fuel to operate.

Crude oil is buried in very deep underground reserves in some countries and a lot of crude oil reserves also exist under the seabed. It is extracted using deep drilling and pumping equipment which has its own entire industry of petroleum engineering.

Before drilling, the oil must be located with some extent of surety and for that geological studies take place that give us an estimate of their existence. Only then are the billions of dollars invested to start drilling and extracting the crude oil. Venezuela and Saudi Arabia possess the two largest oil reserves in the world.

Once extracted, the crude oil, which is measured in barrels, is sent off to the refining process. Refineries take in crude oil as feed and convert it into multiple commercially viable products for consumer use and use in other industries. This is possible due to the large range of components contained in crude oil. The refining process relies on the distillation column, which separates the different components of crude oil based on their relative volatilities. These extracted fractions are then sent of for further processing to improve their qualities and properties. The distillation column allows us to convert the crude oil into LPG, gasoline, diesel, jet fuel, heating oils, lubricants, asphalt, and petrochemicals.

The crude oil charge being brought into the Attock refinery is referred to as ‘sweet crude oil.’ It is called sweet because it contains very low amounts of sulfur which can give the crude oil a very sour smell. The advantage of sweet crude charge is that we do not need to a lot of sulfur removal treatments which contributes to cost savings.

1.2 LPG

LPG, liquified petroleum gas, is a byproduct of the crude refining process. It is widely used as a fuel on both an industrial scale and domestic. It contains components from the C4 and C5 range, mainly Propane and Butane. Naturally LPG is a colorless gas, but it can easily be liquified for efficient storage. The biggest advantage of LPG is the high energy content it has, meaning that it can be cheaply used by the masses for cooking and other home energy intensive processes. Moreover, LPG is also regarded as a very clean burning fuel. It burns efficiently, and releases minimum emissions in the process. In Pakistan and some other countries, LPG is also often used as an alternative to gasoline and diesel by retrofitting cars with a special kit.

1.3 NAPHTHA

Naphtha is an intermediatory product of the refining process. It is not directly sold for commercial use, but instead is the feedstock for several other units and industries which convert it into a final product. It is a mixture of the lighter components of crude oil ranging from C5-C12 and contains mainly aromatics, alkenes, and alkanes. It is light yellow in color and vaporizes quickly but this is not very important since the general, non-industrial public will usually never have to see NAPHTHA. The main use of NAPHTHA is in the petrochemical industry where it helps produce ethylene and propylene. Within the refining industry, NAPHTHA is used in gasoline production, mainly as a blend to help improve the burning of the gasoline.

CHAPTER 2

LITERATURE REVIEW

2.1 Naphtha Stabilization

Naphtha Stabilization refers to the process of removing light gases from Naphtha. Basically, Naphtha is a hydrocarbon mixture of carbon molecules from carbon no. 5 to carbon no. 6. In a crude oil refinery, the crude oil is separated into different fraction which contain a range of carbon molecules, in which the naphtha fraction has lighter hydrocarbon molecules also mixed in it from carbon no.1 to carbon no.4 and their isomers. The presence of these lighter molecules makes the naphtha unstable, unsafe, and unfit for the manufacturing of gasoline and other petrochemical derivatives. This is what prompts the separation of light gases from naphtha, which is known as naphtha stabilization, a process which makes naphtha stable.

As the length of carbon chains of hydrocarbons change, there is a significant effect on the properties of the hydrocarbon mixture like boiling point, flash point, vapor pressure etc. This is the main reason why naphtha requires stabilization. The lighter molecules of C-1 to C-4 in naphtha have a very low flash point, high vapor pressure and low boiling point. Due to this these parameters of the naphtha also change. This has a very adverse effect on naphtha. Firstly, a decrease in flash point makes it very unsafe in terms of transportation because of the fact that a very low flash point means any source of ignition can cause the fire triangle to be completed at even low temperatures and cause the naphtha to combust and the tanks to even explode. Secondly, a very high vapor pressure makes the naphtha unsuitable for storage as a significant part of the naphtha will exist as a gas which will present a high risk of leakage and even storage tank explosion if the pressures are very high. Moreover, a very high vapor pressure also makes the naphtha and the gasoline made from it very unstable for use in gasoline engines as vapors burn very quickly so they would give off very less energy with the same amount as compared to liquid fuel thereby wasting a useful natural resource. Lastly, a low boiling point will also have the same effect. In essence, naphtha stabilization makes naphtha suitable for transportation, storage and as a fuel in gasoline engines.

The stabilization of naphtha not only gives us a safe fuel for further usage, but it also has another very useful advantage. Since light gases are separated from Naphtha during stabilization, these light gases can be used easily as fuels themselves. A very good example of this is the production of LPG which we are achieving in this project. LPG is a simple hydrocarbon molecule with 40% C-3 molecules and 60% C-4 molecules. LPG is a commercially valuable product as it has a very high calorific value, it is a very clean burning fuel with very less to no greenhouse gas emissions and has proven to be a very good fuel in engines. The production of LPG for any industry can result in a great financial gain from its sale.

Figure 1: Commercial Uses of LPG

2.2 Different techniques for Naphtha Stabilization

For this project, there were different techniques available to us for naphtha Stabilization. A comprehensive analysis of all of the techniques was done, based on which we chose our desired technique for this process. These techniques are described below:

2.2.1 Distillation

Distillation is a commonly used technique to separate and purify different components of a mixture based on their boiling points. In naphtha stabilization, distillation is used to remove light ends (lower boiling point hydrocarbons) and heavy components (higher boiling point hydrocarbons) from the naphtha stream, thereby improving its stability.

2.2.2 Hydrotreating

Hydrotreating, also known as hydrofining, involves the treatment of naphtha with hydrogen in the presence of a catalyst at elevated temperatures and pressures. This process helps to remove impurities such as sulfur, nitrogen, and oxygen compounds, which can contribute to instability and corrosion. Hydrotreating also helps to reduce the concentration of olefins (unsaturated hydrocarbons) in naphtha, which can polymerize and form gums.

2.2.3 Clay treatment

Clay treatment, or clay contacting, is a technique where naphtha is treated with adsorbent clays, such as bentonite or Fullers earth. These clays have a high surface area and can selectively adsorb impurities like polar compounds, sulfur compounds, and color bodies, thereby improving the stability and color of the naphtha.

2.2.4 Sweetening

Sweetening involves the removal of sulfur compounds from naphtha, which can cause instability and corrosion. Various methods are employed for sweetening, including chemical treatment with caustic soda (sodium hydroxide) to convert sulfur compounds into water-soluble salts that can be removed by subsequent water washing.

2.2.5 Solvent extraction

Solvent extraction, also known as liquid-liquid extraction, is a process where naphtha is mixed with a selective solvent to extract specific impurities. For example, phenolic compounds, which can contribute to instability, can be removed by extracting them with a selective solvent like caustic soda or a phenol-specific solvent.

2.2.6 Adsorption

Adsorption is a process where naphtha is passed through adsorbent materials, such as activated carbon or molecular sieves. These materials have a high affinity for certain impurities, including sulfur compounds and other polar compounds, which are adsorbed onto their surfaces. The treated naphtha is then separated from the adsorbent material, resulting in improved stability.

2.3 Technique Chosen for this project

After doing a comprehensive cost benefit analysis of the above-mentioned techniques, we came up with the conclusion that the distillation method for stabilization of naphtha would be more suitable. The pros of this technique significantly outweighed the pros of the other techniques. Some of these are mentioned below:

1. As an already established plant with seven other oil refining sub-sections having distillation and separation equipment, like distillation towers, stripping and rectifying towers, adsorption columns, heat exchangers, furnaces etc., the technique of distillation aligned with the current operation and the technical expertise of the plant perfectly. This

meant that no additional human resource needed to be hired, no new extra land or facility needed to be set up for the stabilization process as the tower can be easily installed inside the light crude distillation unit.

2. Being an oil refinery itself, adjusting a distillation tower for stabilization was the least energy intensive option as energy from other streams can be easily utilized as source or sink for the new equipment.
3. No new and specialized equipment needed to be bought except a distillation tower as other auxiliaries like condensers, reboilers and cooling towers are already installed in the plant.
4. No new processing materials and reagents needed to be bought like solvents, catalysts, etc., as the distillation process does not use solvents in the case of naphtha refining.
5. Preliminary energy and material balance on all of these techniques also yielded that this process gives the highest percentage yield of LPG as compared to the other techniques which would result in increased revenue and profitability.
6. All of these reasons concluded that distillation is the most economical, safe, and relatively easy technique to be implemented at the light crude distillation of Attock Refinery Limited.

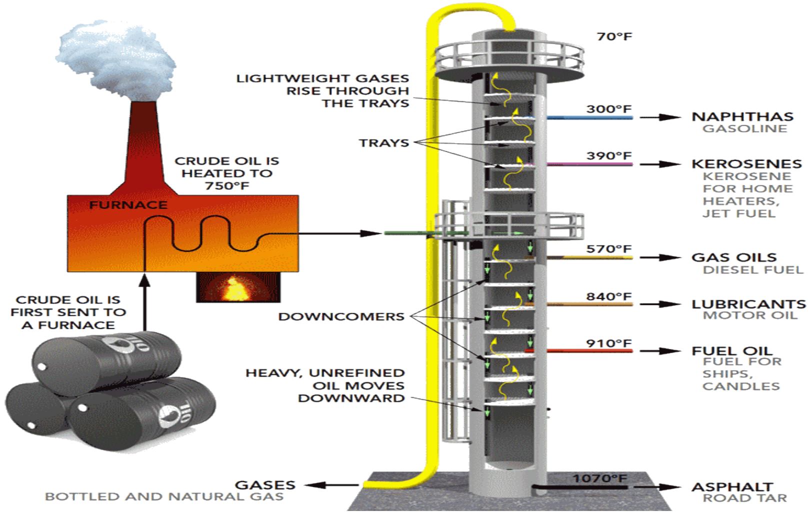


Figure 2:Distillation Column Schematic

2.4 Type of Distillation Column used (atmospheric column)

The crude oil refining process is the separation of crude oil into various fractions or straight-run cuts by distillation in atmospheric and vacuum towers. The main fractions or "cuts" obtained have specific boiling-point ranges and can be classified in order of decreasing volatility into gases, light distillates, middle distillates, gas oils, and residuum. The desalted crude feedstock is preheated using recovered process heat. The feedstock then flows to a direct-fired crude charge heater then into the vertical distillation column just above the bottom, at pressures slightly above atmospheric and at temperatures ranging from 340-370°C (above these temperatures undesirable thermal cracking may occur). All but the heaviest fractions flash into vapor. As the hot vapor rises in the tower, its temperature is reduced. Heavy fuel oil or asphalt residue is taken from the bottom. At successively higher points on the tower, the various major products including lubricating oil, heating oil, kerosene, gasoline, and uncondensed gases (which condense at lower temperatures) are drawn off.

Heat exchange with product and reflux streams raises the temperature of crude oil to around 288°C. It is then charged to the flash zone of the atmospheric fractionators after being heated to roughly 399°C in a furnace. The temperature of the furnace discharge is high enough (343 to 399°C) to cause vaporization of all goods extracted above the flash zone, plus roughly 10% to 20% of the bottoms product. Condensing the tower's upper vapors and returning a part of the liquid to the top of the tower provide reflux. Lower in the tower, pump-around and pump-back streams.

The amount of reflux below the point of draw off reduces with each side stream product removed from the tower. Removing all heat at the top of the tower achieves maximum reflux and fractionation, but this results in an inverted cone-type liquid loading, which necessitates a very wide diameter at the top of the tower.

Intermediate heat-removal streams are utilized to induce reflux below the side stream removal sites, reducing the top diameter of the tower and levelling out the liquid loading across its length. To do so, liquid is withdrawn from the tower, cooled in a heat exchanger, and then returned to the tower, or a portion of the cooled side stream may be returned to the tower. This chilly stream condenses more of the vapors rising from the bottom, increasing reflux below that point. Although reboilers are not typically used in crude towers, many trays are usually incorporated below the flash zone, and steam is supplied below the bottom tray to strip any remaining gas oil from the liquid in the flash zone and generate high flash-point bottoms. In most cases, the atmospheric fractionator has 30 to 50 fractionation trays. Separating complex combinations in crude oils is quite simple, and each side stream product typically requires five to eight trays, plus the same number above and below the feed plate. As a result, a crude oil atmospheric fractionation tower with four liquid side stream draw offs will necessitate between 30 and 42 trays. The extracted liquid side stream from the tower will contain components with a low boiling point and a lower flashpoint Each side stream's "light ends" are stripped in a separate small stripping tower with four to ten trays and steam introduced under the bottom tray. Above the appropriate side-draw tray, the steam and stripped light ends are vented back into the atmospheric fractionator's vapor zone. The pentane and heavier percentage of the vapors that pass out of the top of the atmospheric tower are condensed by the overhead condenser on the atmospheric tower. This is the light gasoline section of the tower overhead vapor, containing some propane and butanes as well as practically all of the higher-boiling components. Some of the condensate is returned to the top of the tower as reflux, while the rest is routed to the refinery gas plant's stabilization section, where the butanes and propane are separated from the C5-180°F (C5-82°C) condensate. LSR is a kind of gasoline.

Stabilization is a slightly different kind of distillation where our entering feed is unstabilized light naphtha containing light gases from C-1 to C-4. So, in the case of stabilization, essentially two product streams are separated unlike a normal atmospheric distillation column where side streams or cuts are obtained at intermediate plates. We can extract side streams just for the purpose of extracting heat from them to be used in other processes as the same principle of crude oil refining is being applied which causes intermediate fractions to be vaporized which have a lot of sensible and latent heat stored in them. By utilizing this principle in the distillation tower, we can obtain a top product stream of pure LPG and off gases of C-1 and C-2 which can then be sent to a two phase separator (accumulator) to be separated.

CHAPTER 3

PROCESS DESCRIPTION

3.1 Description

The main operations that take place in our plant are desalting, pre flash separation, stabilizer and then water removal before storage. The crude oil enters the plant at normal temperatures and pressures and so it first must be pumped into a heater to raise both its temperature and pressure. After the initial heating the, the oil enters an electric desalter. The purpose of this desalter is to remove salt content from the oil by adding water and charging it using electricity. The polarity allows us to effectively separate the salty water from the crude oil.

After the desalter, the crude oil is once again heated to match the pre flash tower's entering temperature. The pre flash tower allows us to separate our crude oil into heavy and light components. Heavy products exit out of the bottom from where they head to the main distillation unit while the products from the top are lighter (C1-C12) and head for the stabilizer. After the exit, the top draw is condensed at which point the accumulator separates off gases from the draw. Reflux is provided and the rest move forward.

At this stage, we have un-stabilized NAPHTHA, it is once again heated to achieve entering temperature requirements and then enters the stabilizer column. This is the second distillation column in our process, and it used to help separate LPG from NAPHTHA to create LPG product stream and a stabilized NAPHTHA stream. From the bottom of the tower, we get our stabilized NAPHTHA which can be sent to other units and from the top we get our LPG. Our top draw is condensed, followed by the removal of off gases before reflux can be provided. The remaining product heads over to the salt drier which allows us to remove any excess water in our product stream.

Afterwards our commercially viable LPG can be safely sent to product storage or for further processing like adding artificial odors for extra safety.

3.2 Process Flow Diagram

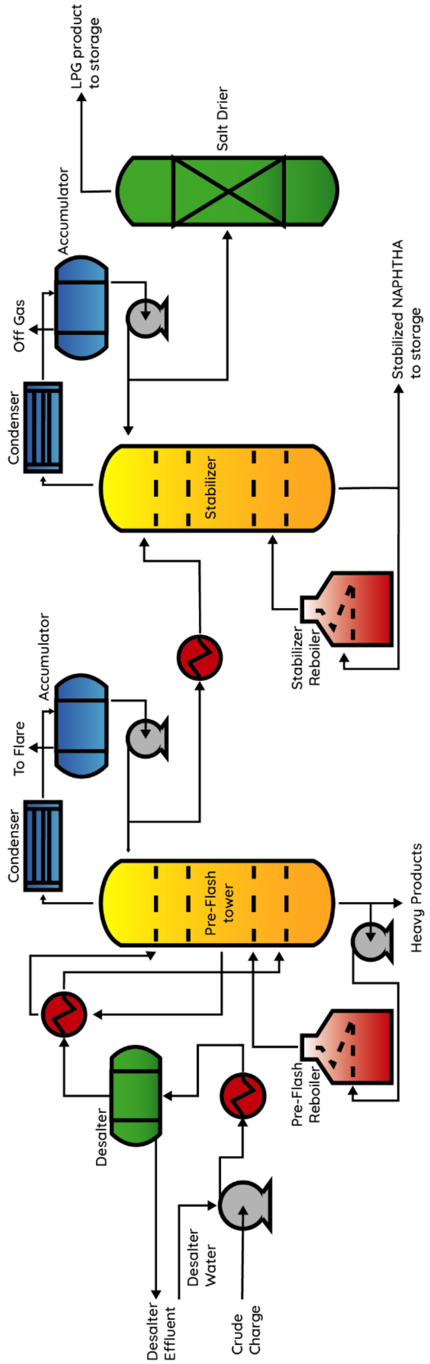


Figure 3:Process Flow Diagram

CHAPTER 4

MATERIAL BALANCE

The material balance in the proposed design was conducted by applying the law of conservation of mass to each of the significant components involved. Since the equipment utilized in the design were unit operations rather than involving chemical reactions, the task of material balance was simplified.

In the material balance calculations, all hypothetical components derived from the petroleum assay data were considered. The balances encompassed individual component analyses unless explicitly stated otherwise.

The true boiling point (TBP) analysis of the crude oil feed employed in the design is as follows:

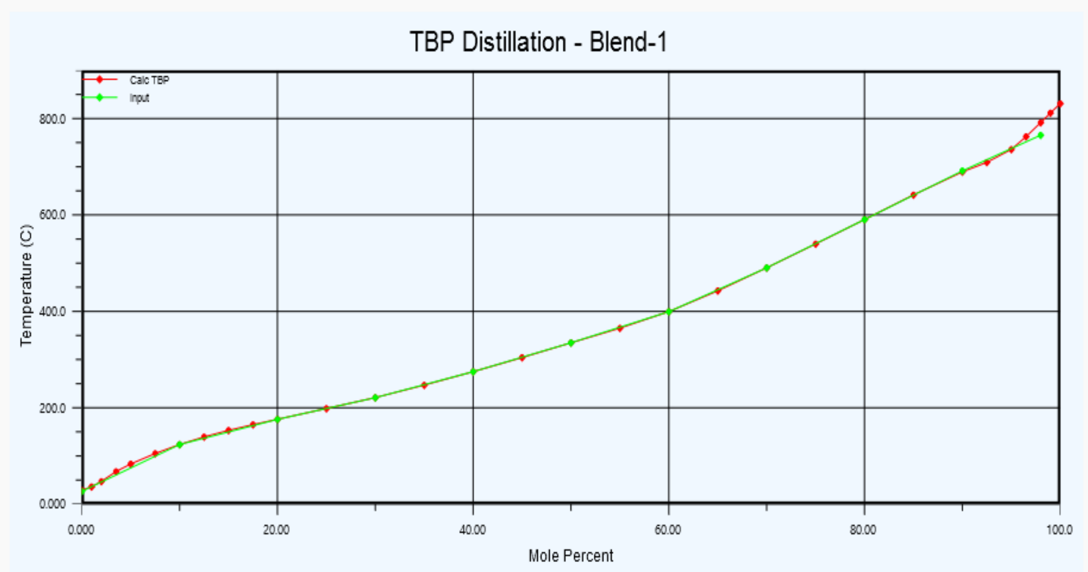


Figure 4: True boiling point (TBP) analysis of the crude oil feed

By considering the TBP analysis along with the conservation of mass principles, a comprehensive material balance was established for the proposed design. This enabled a thorough understanding of the distribution and transformation of components throughout the system, serving as a foundation for subsequent analyses and evaluations.

Furthermore, the material balance calculations incorporated the calculated crude oil composition and properties, which were obtained from Table 6, the output blend of the analysis. This data provided a comprehensive understanding of the component distribution, properties, and quantities within the proposed design

4.1 Splitter (Tee)

In the case of splitting a crude oil stream into two equal flow rates, the material balance ensures that half of the incoming crude oil mass is allocated to each resulting stream. This balance is crucial for maintaining process efficiency, as it allows for consistent and controlled flow rates in downstream equipment or processes.

Furthermore, the two split streams are sent to separate heat exchanger trains, maintaining a constant flow rate throughout the exchangers is essential. This ensures uniform heat transfer and helps optimize the performance of the heat exchanger system. The material balance ensures that the flow rates of the split streams remain constant, enabling efficient heat exchange and achieving the desired heating objectives within each exchanger.

Table 1: Mass Balance across splitter

Splitter			
Streams	**IN**	**Out**	
	Crude Charge	**Stream A**	**Stream B**
Vapour Fraction	0.00	0.00	0.00
Temperature C	26.70	118.00	265.60
Pressure(bar)	1.01	11.72	8.15
Mass Flow(kg/hr)	171146.61	85573.31	85573.31
Components			
H2O	1711.47	855.73	855.73
Salt	698.28	349.14	349.14
Methane	0.00	0.00	0.00

Table 1 (cont'd)

Ethane	770.16	385.08	385.08
Propane	2858.15	1429.07	1429.07
n-butane	3388.70	1694.35	1694.35
i-butane	3200.44	1600.22	1600.22
C5-C6	2145.10	1072.55	1072.55
C7-C9	3594.08	1797.04	1797.04
C10-C12	35685.06	17842.53	17842.53
C13-C16	46709.33	23354.67	23354.67
C17-C-21	25678.78	12839.39	12839.39
C-22-C-28	51857.42	25928.71	25928.71
Total	178296.97	89148.49	89148.49
Subtotal	**178296.97**	**178296.97**	

4.2 Heat Exchanger

According to the law of conservation of mass, the mass flow rate of components entering the heat exchanger is equal to the mass flow rate of components exiting the heat exchanger, ensuring mass conservation throughout the process. This can be summarized as:

IN = OUT

This principle highlights that the mass flow rates of the components entering and exiting the heat exchanger remain balanced, reflecting the conservation of mass within the system.

Table 2: Mass Balance across Heat Exchanger

Streams	**Shell Side**		**Tube Side**	
	In	**Out**	**IN**	**OUT**
Vapour Fraction	0.00	0.00	0.00	0.00
Temperature C	66.67	135.00	185.00	117.22
Pressure(bar)	10.76	9.65	11.44	11.33

Table 2 (cont'd)

Mass Flow(kg/hr)	40191.91	40191.91	37316.14	37316.14
Specific Heat Capacity	2.46		2.67	
Q(Heat)(MMBTU/hr)	-75.92	-69.52	-58.00	-64.40
Q=mcpdt(kJ/hr)	6752357.00		6752357.00	
Components				
H2O	100.48	100.48	0.00	0.00
Salt	0.00	0.00	0.00	0.00
Methane	0.00	0.00	0.00	0.00
Ethane	430.05	430.05	0.00	0.00
Propane	1591.60	1591.60	0.00	0.00
n-butane	1872.94	1872.94	806.03	806.03
i-butane	1772.46	1772.46	253.75	253.75
C5-C6	12057.57	12057.57	36256.36	36256.36
C7-C9	10007.79	10007.79		
C10-C12	8661.36	8661.36		
C13-C16	3617.27	3617.27		
C17-C-21	40.19	40.19		
C-22-C-28	40.19	40.19		
Total	**40191.91**		**37316.14**	

4.3 Desalter:

The material balance of a desalter is based on the principle of conservation of mass, which states that the total mass entering the desalter must equal the total mass exiting the unit. This balance ensures that the process operates efficiently and that the desired separation of water and impurities from the crude oil is achieved.

Table 3: Mass Balance across Desalter Section

Desalter				
	IN		**OUT**	
	Crude Oil	**Water**	**Desalted Crude**	**Water + Salt**
Vapor Fraction	0.00	0.00	0.00	0
Temperature (C)	137.20	117.80	137.20	135.00
Pressure (bar)	11.72	11.72	11.72	10.34
Mass Flow (Kg/hr)	171146.61	12700.58	171146.61	12700.58
Component (kg/hr)				
H2O	1711.47	12700.58	4484.04	9928.00
SALT	698.28	0.00	0.00	698.28
Methane	0.00	0.00	0.00	0.00
Ethane	770.16	0.00	770.16	0.00
Propane	2858.15	0.00	2858.15	0.00
n-Butane	3388.70	0.00	3388.70	0.00
i-Butane	3200.44	0.00	3200.44	0.00
C5-C6	2105.10	0.00	2105.10	0
C7-C9	3594.08	0.00	3594.08	0
C10-12	35085.06	0.00	35085.06	0
C13-C16	46709.33	0.00	46709.33	0
C17-C21	19168.42	0.00	19168.42	0
C22-C28	51857.42	0.00	51857.42	0.00
	171146.61	12700.58	173220.91	10626.28
Total	**183847.19**		**183847.19**	

4.4 PreFlash Tower

The material balance of a preflash tower is based on the principle of conservation of mass, which states that the total mass entering the tower must equal the total mass exiting the unit. This balance ensures that the separation process is operating effectively, with the desired fractions being collected at the appropriate points in the tower.

Table 4: Mass Balance across Preflash Tower

PreFlash Tower					
	IN			**OUT**	
	Desalted Crude Oil Feed	**Reflux**	**Boil-Up**	**Tops**	**Bottoms**
Vapor Fraction	0.55	0.00	0.43	1.00	0
Temperature (C)	231.11	69.00	309.44	136.11	275.00
Pressure (bar)	3.01	4.35	3.83	2.48	150.00
Mass Flow (Kg/hr)	171682.30	31072.41	120196.89	71778.67	246137.16
Component(kg/hr)					
H2O	4498.08	77.68	0.00	4572.68	0.00
SALT	0.00	0.00	0.00	0.00	0.00
Methane	0.00	0.00	0.00	0.00	0.00
Ethane	755.40	332.47	0.00	1086.50	0.00
Propane	2798.42	1230.47	0.00	2749.12	0.00
n-Butane	3313.47	1447.97	12.02	3237.22	24.61
i-Butane	3124.62	1370.29	12.02	3064.95	24.61
C5-C6	2111.69	9321.72	1442.37	20169.81	2953.65
C7-C9	3605.33	7737.03	2632.31	16121.49	5390.40
C10-12	30422.10	6696.10	3113.10	15791.31	6374.95

Table 4 (cont'd)

C13-C16	37255.06	2796.52	24039.38	7185.04	49227.43
C17-C21	40877.56	31.07	42068.91	71.78	86148.01
C22-C28	42920.58	31.07	46876.79	71.78	95993.49
Subtotal	171682.30	31072.41	120196.89	71778.67	246137.16
Total	**376606.12**			**376605.64**	

4.4.1 Accumulator PreFlash Tower

Table 5: Mass Balance across Accumulator of Preflash Tower

Accumulator of Preflash				
	IN		**OUT**	
	From Condenser	**Direct Over Head**	**Light Naphtha to stabilizer**	**Water**
Vapor Fraction	0.00	1.00	0	0
Temperature (C)	37.78	135.56	170.56	65.00
Pressure (bar)	2.14	2.14	3.83	2.14
Mass Flow (Kg/hr)	63683.41	8095.26	71264.75	2391.87
Component Mass Flow (Kg/hr)				
H2O	2279.87	289.81	178.16	2391.87
SALT	0.00	0.00	0.00	0.00
Methane	0.00	0.00	0.00	0.00
Ethane	662.31	84.19	762.53	0.00
Propane	2439.07	310.05	2822.08	0.00
n-Butane	2872.12	365.10	3320.94	0.00
i-Butane	2719.28	345.67	3142.78	0.00
C5-C6	17895.04	2274.77	21379.42	0.00
C7-C9	14303.29	1818.19	17744.92	0.00
C10-12	14010.35	1780.96	15357.55	0.00

Table 5 (cont'd)

C13-C16	6374.71	810.34	6413.83	0.00
C17-C21	63.68	8.10	71.26	0.00
C22-C28	63.68	8.10	71.26	0.00
Subtotal	63683.41	8095.26	71264.75	513.92
Total	**71778.67**		**71778.67**	

4.5 Stabilizer

Table 6: Mass Balance across Naphtha Stabilizer

Stabilizer (T-200)					
	IN			**Out**	
	Feed	**Reboil**	**Reflux**	**Bottoms**	**Tops**
Vapour Fraction	0.00	1.00	0.00	0.00	1.00
Temperature C	135.00	185.00	37.80	173.30	65.00
Pressure(bar)	9.65	11.45	12.48	11.45	11.10
Mass Flow(kg/hr)	40191.91	25526.36	6969.45	62891.49	9796.23
Components					
H2O	100.48	0.00	6.27	0.00	106.63
Salt	0.00	0.00	0.00	0.00	0.00
Methane	0.00	0.00	0.00	0.00	0.00
Ethane	430.05	0.00	634.22	0.00	1064.56
Propane	1591.60	0.00	2336.86	0.00	3266.06
n-butane	1872.94	1605.61	1625.97	2471.64	2271.75
i-butane	1772.46	579.45	2251.13	855.32	3145.57
C5-C6	12057.57	23341.31	115.00	37207.86	160.66
C7-C9	10007.79	0.00	0.00	9999.75	0.00
C10-C12	8661.36	0.00	0.00	8660.16	0.00
C13-C16	3617.27	0.00	0.00	3616.26	0.00

Table 6 (cont’d)

C17-C-21	40.19	0.00	0.00	40.25	0.00
C-22-C-28	40.19	0.00	0.00	40.25	0.00
Subtotal	40191.91	25526.36	6969.45	62891.49	9796.23
Total	**72687.72**			**72687.72**	

4.5.1 Accumulator Stabilizer Tower

Table 7: Mass Balance across Accumulator of Naphtha Stabilization Tower

Stabilizer Accumulator			
	IN	**Out**	
	Tops	**LP Tops**	**Off Gas**
Vapour Fraction	1.00	0.00	0.00
Temperature C	37.80	37.80	37.80
Pressure(bar)	11.10	10.90	10.90
Mass Flow(kg/hr)	9796.23	9776.28	19.95
Components (kg/hr)			
H2O	65.63	65.63	0.00
Salt	0.00	0.00	0.00
Methane	19.95	0.00	19.95
Ethane	886.56	889.64	0.00
Propane	3266.06	3277.99	0.00
n-butane	2271.75	2280.81	0.00
i-butane	3145.57	3157.74	0.00
C5-C6	160.66	161.31	0.00
C7-C9	0.00	0.00	0.00
C10-C12	0.00	0.00	0.00
C13-C16	0.00	0.00	0.00
C17-C-21	0.00	0.00	0.00
C-22-C-28	0.00	0.00	0.00
Subtotal	9796.23	9776.28	19.95
Total	**9796.23**	**9796.23**	

4.6 Overall Material Balance:

Table 8: Overall Material Balance

Overall Material Balance								
Streams	**IN**		**Out**					
	Crude Charge	**Desalter Water In**	**Residue From Preflash**	**Salt+Water from desalter**	**Heavy Naphtha to storage**	**Stabilized Naphtha to Storage**	**LPG**	**Water**
Vapour Fraction	0.00	0.00	0.00	0.00	0.00	0.00		0.00
Temperature C	26.70	118.00	265.60	135.00	37.80	185.00	37.80	37.80
Pressure(bar)	1.01	11.72	8.15	11.72	4.76	11.44	15.90	15.90
Mass Flow(kg/hr)	171146.61	12700.00	125940.38	10626.30	9834.80	37316.14	2804.30	2.53
Components								
H_2O	1711.47	12700.00	0.00	9928.00	3097.00	0.00	0.00	1383.00
Salt	698.28	0.00	0.00	698.28	0.00	0.00	0.00	0.00
Methane	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ethane	770.16	0.00	0.00	0.00	0.98	0.00	255.42	0.00
Propane	2858.15	0.00	0.00	0.00	8.85	0.00	941.13	0.00
n-butane	3388.70	0.00	12.59	0.00	21.64	806.03	654.83	0.00

Table 8 (cont’d)

i-butane	3200.44	0.00	12.59	0.00	16.72	250.02	906.61	0.00
C5-C6	2145.10	0.00	1511.28	0.00	3048.79	15784.73	46.31	0.00
C7-C9	3594.08	0.00	2758.09	0.00	3147.14	8582.71	0.00	0.00
C10-C12	35685.06	0.00	3261.86	0.00	1868.61	9217.09	0.00	0.00
C13-C16	46709.33	0.00	25188.08	0.00	1180.18	891.86	0.00	0.00
C17-C-21	25678.78	0.00	44079.13	0.00	339.30	891.86	0.00	0.00
C-22-C-28	51857.42	0.00	49116.75	0.00	196.70	891.86	0.00	0.00
Subtotal	178296.97	12700.00	125940.38	10626.28	12925.90	37316.14	2804.30	1383.00
Total	**190996.97**		**190995.99**					

CHAPTER 5

ENERGY BALANCE

Energy balance is a calculation that determines whether, energetically, the process being designed is practical or not as it uses the laws of conservation of energy and thermodynamics to see if there is any irregularity in the energetics of the entire process. It also determines the energy requirements of a new process and helps make the decision of which energy resources to buy and which can be fulfilled using in plant extra energy sources.

Material balance was carried out previously which gave the preliminary data for energy balance to be based on. Energy balance was only applied to processes which involved changes in temperature, pressure, or any other energy of the fluid. Fluid enthalpy data was also utilized.

5.1 Pumps

There were a total of four pumps utilized for the process. These are mentioned below:

1. Crude Charge Pump
2. Pre flash Reboiler Pump
3. Pre flash Accumulator Pump
4. Stabilizer Reboiler Pump

The Bernoulli's equation was modified and used for the power requirements used for the pumps.

The Bernoulli's equation is shown below:

$$Q_{p1} = \dot{m}\left(\Delta Z + \frac{\Delta P}{\rho g} + \Delta \frac{v^2}{2g}\right) + Q_{losses}$$

Ignoring elevations, changes in velocity and losses, and incorporating efficiency of pump, the equation is reduced to:

$$\eta Q_{p1} = \dot{m}\left(\frac{\Delta P}{\rho}\right)$$

The required data and corresponding duties of all the pumps are shown in the table below.

Table 9: Duty for Pumps

Pumps				
Property	**1**	**2**	**3**	**4**
Temperature/◦C	26.7	264.4	65	37.8
Pressure$_{in}$/kPa	0.0	303	234	1090
Pressure$_{out}$/kPa	1,634	938	607	1611
ΔP/kPa	1,634	635	373	521
Density/(kg/m^3)	790	626	662	514
Mass Flowrate/ (kg/hr)	171,147	120,197	71,265	9,776
Isentropic Efficiency	0.75	0.75	0.75	0.75
Pump Duty (kW)	131.1	45.2	15.0	3.7
Total Power Required by Pumps	195kW			

Hence, the total electrical power required for all four pumps is calculated to be **195 kW.**

5.2 Heat Exchangers

Energy balance was used to calculate the heat duty of each heat exchanger i.e., the heat energy required to bring our fluid to the desired temperature by balancing it exactly with the amount of heat provided by the cold fluid.

There were essentially three heat exchangers used in our process.

1. Desalter Heat Exchanger
2. Pre flash heat exchanger
3. Stabilizer heat exchanger

The basic equations used for all heat exchangers were:

1. Equation of sensible heat transfer:

$$Q_{sensible} = \dot{m} c_p \Delta T$$

2. Equation of latent heat transfer:

$$Q_{latent} = \dot{m}(\Delta H)_{lv}$$

The calculations for these heat exchangers are shown below.

Table 10: Energy Balance on Desalter Heat Exchanger

Desalter Heat Exchanger		**Shell Side**		**Tube Side**	
Parameters	Units	Crude In	Crude Out	Steam In	Steam Out
Vapor Fraction		0	0	1	1
Temperature	◦C	27.2	137.2	427	204
Pressure	kPa	1634	1172	1200	1025
Mass Flow	Kg/hr	171,147		102,084	
Specific Heat Capacity	kJ/kg. ◦C	2.26		1.87	
Duty, Q	kW	11,825		11,825	

Table 11: Energy Balance on Pre flash Heat Exchanger

Pre flash Heat Exchanger		**Shell Side**		**Tube Side**	
Parameters	Units	Crude In	Crude Out	Residue In	Residue Out
Vapor Fraction		0	0	0	0
Temperature	◦C	137.2	231.1	265	146
Pressure	kPa	1034	310	1200	1025
Mass Flow	Kg/hr	171,147		120,197	
Specific Heat Capacity	kJ/kg. ◦C	2.26		2.54	
Duty, Q	kW	10,090		10,090	

Table 12: Energy Balance on Stabilizer Heat Exchanger

Stabilizer Heat Exchanger		**Shell Side**		**Tube Side**	
Parameters	Units	Crude In	Crude Out	Steam In	Steam Out
Vapor Fraction		0	0	0	0
Temperature	◦C	66.7	135	350	225
Pressure	kPa	1076	965	1200	1025
Mass Flow	Kg/hr	40,192		24,195	
Specific Heat Capacity	kJ/kg. ◦C	2.06		1.87	
Duty, Q	kW	1,571		1,571	

5.3 Condensors

For condensers, the total heat energy released is calculated using the equations of sensible heat and latent heat mentioned in section 5.3.

There are two condensers:

1. Pre flash condenser.
2. Stabilizer condenser.

The calculations are tabulated below for the condensers mentioned above.

Table 13: Pre flash condenser

Pre flash condenser		**Shell Side**		**Tube Side**	
Parameters	Units	Tops In	Tops Out	Cooling water in	Cooling water out
Vapor Fraction		1	0	0	0
Temperature	◦C	136.1	65	25	89
Pressure	kPa	248	234	101	100
Mass Flow	Kg/hr	71,780		36,595	
Specific Heat Capacity	kJ/kg. ◦C	1.92		4.184	
Duty, Q	kW	2,722		2,722	

Table 14: Stabilizer condenser

Stabilizer condenser		**Shell Side**		**Tube Side**	
Parameters	Units	Tops In	Tops Out	Cooling water in	Cooling water out
Vapor Fraction		1	0	0	0
Temperature	◦C	65	37.8	25	42
Pressure	kPa	1,110	1,248	101	100

Table 14: Stabilizer condenser (cont'd)

Mass Flow	Kg/hr	9,796	6,818
Specific Heat Capacity	kJ/kg. ◦C	1.85	4.184
Duty, Q	kW	134.7	134.7

5.4 Furnaces

For the energy balance of the furnace, the heat energy required to obtain a specific temperature rise in the crude oil was first found. The calorific value of the fuel used in the furnace was then divided from the heat requirement to obtain the amount of fuel required.

There are two furnaces in our plant.

1. Pre flash furnace.
2. Stabilizer furnace.

The equation used for furnace heat required is:

$$Q_{sensible} = \dot{m} c_p \Delta T$$

The equation used for fuel amount is:

$$Q = \dot{m} \times Calorific\ value\ of\ fuel$$

Table 15: Energy balance on Pre flash furnace

Pre flash Furnace	Units	In	Out
Mass flow	kg/h	120,197	120,197
Heat Capacity	kJ/kg˚C	4.26	4.26
Temperature	˚C	265	309.4
Duty	kW	6,315	

Table 16: Amount of fuel of Pre flash furnace

Natural Gas		
Q	kJ/s	6,315
Calorific Value	kJ/kg	54,300
Mass	kg/hr	443

Table 17: Energy balance on Stabilizer furnace

Pre flash Furnace	**Units**	**In**	**Out**
Mass flow	kg/h	25,523	25,523
Heat Capacity	kJ/kg˚C	4.26	4.26
Temperature	˚C	173.3	185
Duty	kW	7,688	

Table 18: Amount of fuel of Stabilizer furnace

Natural Gas		
Q	kJ/s	7,688
Calorific Value	kJ/kg	54,300
Mass	kg/hr	510

5.5 Summary of Energy Balance

Table 19: Total heat energy required

Equipment	**Heat Energy**
Heat Exchangers	23,486 kW
Furnaces	14,003 kW
Total	37.5 MW

Table 20: Total Heat Energy Released

Equipment	**Energy**
Condenser	2,857 kW
Total	2.857 MW

Table 21: Total Electrical Energy required

Equipment	**Energy**
Pumps	195 kW
Total	195 kW

Conclusively, the utilities required as per the energy balance are steam, cooling water, fuel (natural gas) and electricity for this plant. Instead of steam and cooling water, other streams from the plant can be used as the source or sink of heat energy, but in this calculation, these were used to give us the maximum possible cost estimates.

CHAPTER 6

EQUIPMENT DESIGN

6.1 Heat exchanger

6.1.1 Heat exchanger stream specifications

The specification of our inlets streams to the heat exchanger are as shown in the table below:

Table 22: Heat Exchanger Stream Specifications

Heat Exchanger Stream Specifications		**Shell Side**	**Tube Side**
	Units	Naphtha	Steam
Inlet Temperature	^{o}C	82	426
Outlet Temperature	^{o}C	135	204
Mass Flow Rate	kg/hr	40,190	11,420
Specific Heat Capacity	kJ/kg.^{o}C	2.39	2.01
Thermal Conductivity	W/m.^{o}C	0.14	0.042
Viscosity	cP	0.17	0.02
Friction Factor	-	0.0015	0.00015

6.1.2 Calculating Heat Flow

Our primary step, after noting down a0ll the given stream information from the industry and all the required specs taken from their respective literature, is to calculate the heat flow of our shell stream. We do this to help us understand how much steam will be required to achieve our desired heat flow.

Heat flow, the transfer of heat energy, is represented by the letter "Q," has the units kJ/hr and is calculated using the formula:

$$Q = \dot{m}c\Delta T$$

$$Q = (40{,}190)(2.39)(135 - 82) = 5{,}099{,}703 kJ/hr$$

6.1.3 Tube Side Mass Flow Rate

Now that we have our required heat flow to heat approximately forty thousand kilograms of Naphtha from 82^{o}C to 135^{o}C, we will use this value to calculate the mass flow rate of steam required to achieve this heat transfer.

We will be using the same formula that was used to calculate our heat flow rate; however, we will change the formula's subject from mass flow to heat flow:

$$\dot{m} = \frac{Q}{c\Delta T}$$

$$\dot{M} = \frac{5099703}{(2.01)(426 - 204)} = 11{,}420 kg/hr$$

6.1.4 LMTD Calculations

The Log Mean Temperature Difference is a calculation that allows to find the mean temperature inside our heat exchanger. We use the log mean variation instead of a regular mean calculation because the heat exchange inside the heat exchanger is not linear in nature. Hence, our mean temperature calculations will not reflect the true mean temperature. The LMTD allows us to linearize the heat exchange and then find a mean temperature which we can use.

The formula used to calculate the LMTD is as follows:

$$LMTD = \frac{\Delta T_H - \Delta T_C}{ln\left(\frac{\Delta T_H}{\Delta Tc}\right)}$$

Table 23: Stream Temperatures

Hot (^{o}C)	**Cold (^{o}C)**	**Difference (^{o}C)**
426	135	274
204	82	105
204	35	151

We find our LMTD to be:

LMTD = 177.3 ^{o}C

The LMTD that we have just found is uncorrected and needs to be corrected using a correction factor. We first calculate the variable R and S and use them against the correction factor figure to get our factor.

The formulas for R and S are as follows:

$$R = \frac{T_1 - T_2}{t_2 - t_1} \text{ and } S = \frac{t_2 - t_1}{T_1 - t_1}$$

R = 4.18 and S = 0.15

We plot this against graph as such:

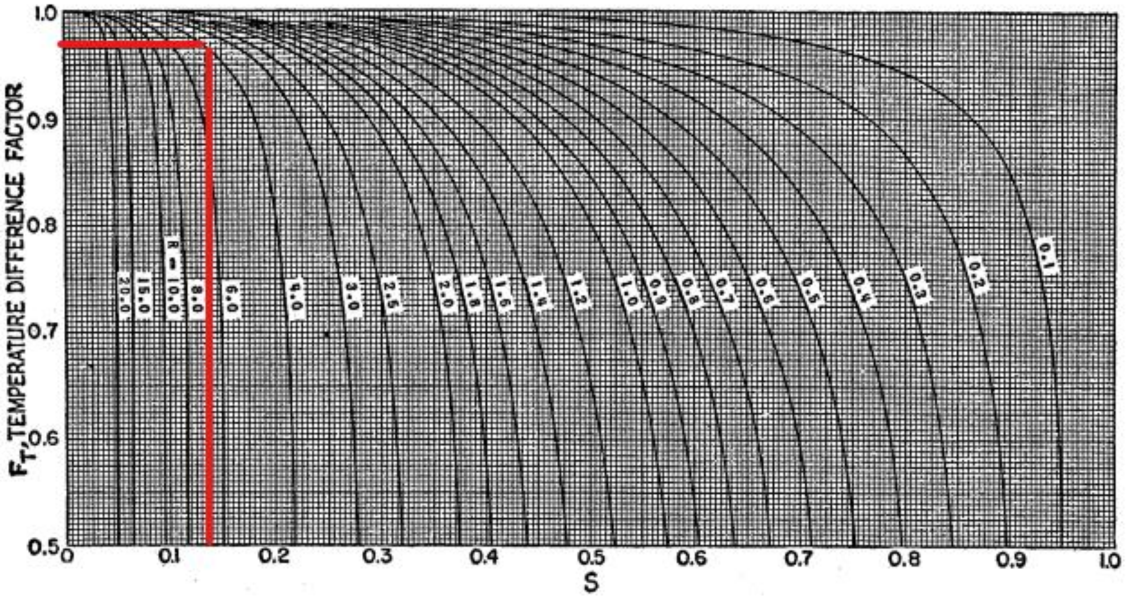


Figure 5: Temperature difference correction factor

We get a correction factor of:

$$Ft = 0.94$$

We use this to calculate our true LMTD which is:

$$\text{True LMTD} = 177.3 \times 0.94 = 161.1\ ^{o}C$$

6.1.5 Calorific Temperature Calculations

Next, we will calculate our calorific temperature which will be used in our heat coefficient calculations. For this we will calculate the factors needed and use them to obtain our calorific temperature factor from a graph.

Our first step is to get the ratio of the change in cold stream temperature and the hot stream temperature.

$$\frac{\Delta t_c}{\Delta t_H} = 0.42$$

Following this, we use the figure 3 to find our K_c value and our calorific temperature factor, F_c.

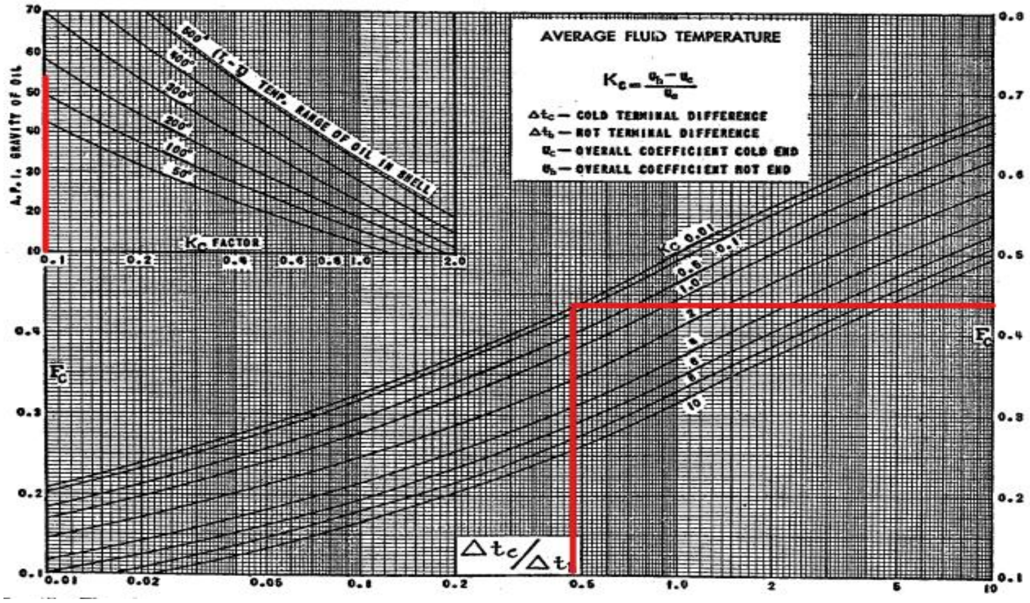


Figure 6: Calorific temperature Factor, F_c

We get a factor of Fc:

Fc = 0.45

We will now use this to calculate our hot side calorific temperature using the formula:

$$Tc = T_2 + F_c(T_1 - T_2) = 304^0C$$

Cold side calorific temperature:

$$Tc = t_1 + F_c(t_2 - t_1) = 106^0C$$

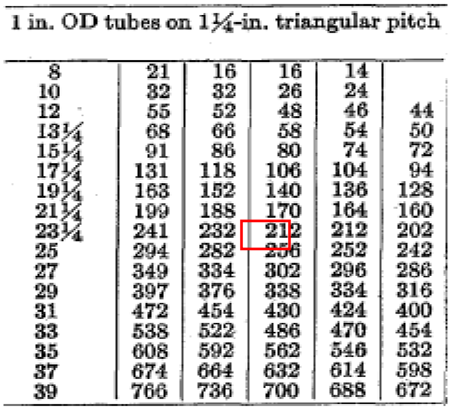
This is a reasonable point in our calculation to assume a heat exchanger design to allow us to proceed with its verification.

6.1.6 Heat Exchanger Design

Table 24: Heat Exchanger Design Data

Exchanger Data			
Shell Side		**Tube Side**	
ID (m)	0.54	Number & Length (m)	188, 6.10
Baffle Space (m)	0.15	OD (m), BWG, & Pitch (m)	0.025, 14, 0.032
Passes	1	Passes	2

We procured our number of tubes using literature as follows.



1 in. OD tubes on 1¼-in. triangular pitch

8	21	16	16	14	
10	32	32	26	24	
12	55	52	48	46	44
13¼	68	66	58	54	50
15¼	91	86	80	74	72
17¼	131	118	106	104	94
19¼	163	152	140	136	128
21¼	199	188	170	164	160
23¼	241	232	212	212	202
25	294	282	256	252	242
27	349	334	302	296	286
29	397	376	338	334	316
31	472	454	430	424	400
33	538	522	486	470	454
35	608	592	562	546	532
37	674	664	632	614	598
39	766	736	700	688	672

Figure 7: Tube Sheet Count

6.1.7 Shell side calculations for heat coefficient

We will begin by first calculating our flow area, α_s. The formula used is shown followed by the calculations.

$$\alpha_s = ID(C)(B) \,/\, 144P_T$$

$$\alpha_s = 21.25(0.25)(6)/144(1.25) = 0.177 = 0.0016\text{m}^2$$

Next, we will be calculating our mass velocity, G_s.

$$G_s = \frac{W}{\alpha_s}$$

$$G_s = \frac{40190}{00016} = 2{,}424{,}887\text{kg/hr.m}^2$$

Next, we will calculate our Reynold's Number. For this we will first need to find our equivalent diameter and our viscosity.

Our viscosity is:

$$\mu = 0.17\text{cP}$$

And our equivalent diameter is:

$$De = 0.018\text{m}$$

We will now find our Reynold's Number using the calculation as shown below.

$$Re = \frac{D_e G_S}{\mu} = 71{,}000$$

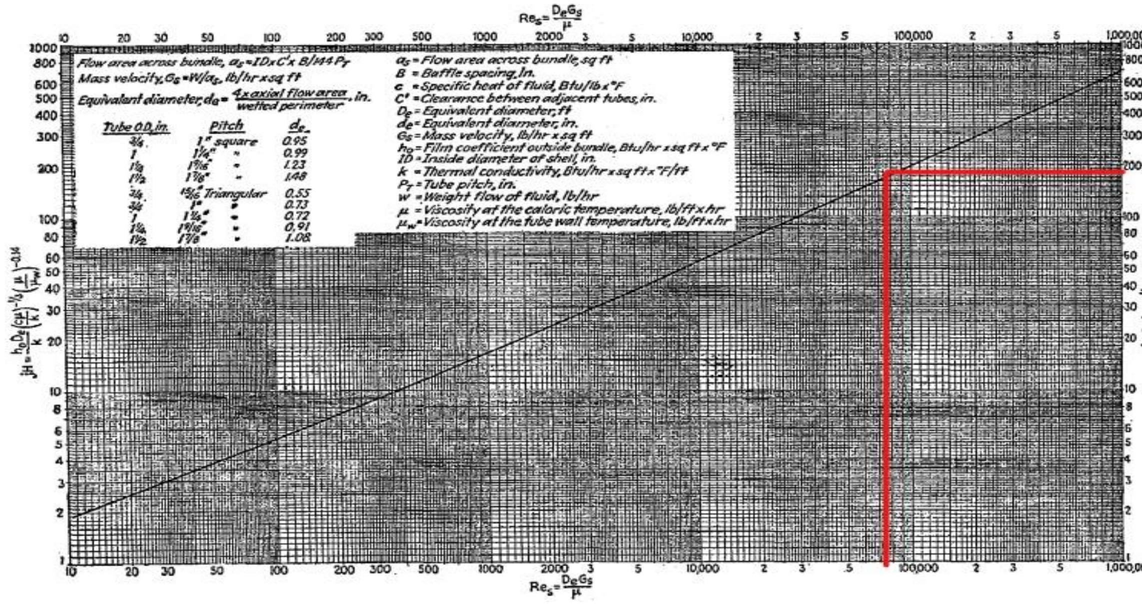


Figure 8: j_H against Reynold's Number Graph

We get a value of:

$$j_H = 160$$

Next, we will get our thermal conductivity and specific heat values.

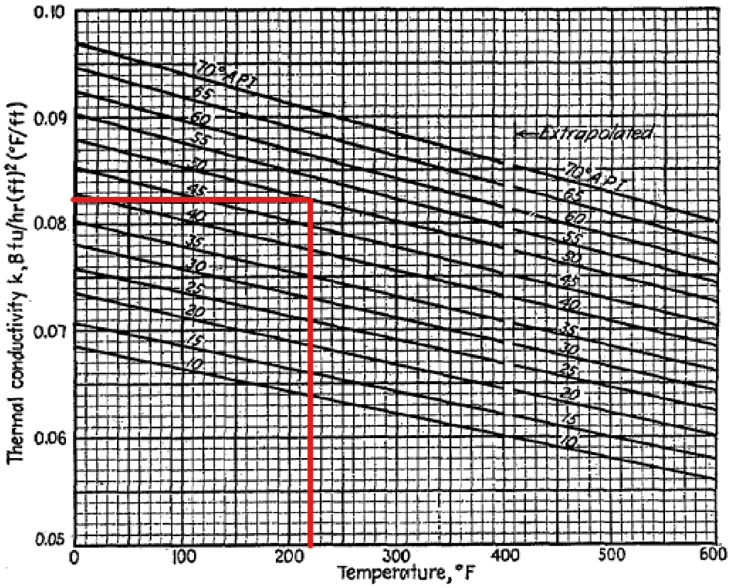


Figure 9: Thermal conductivity of hydrocarbons

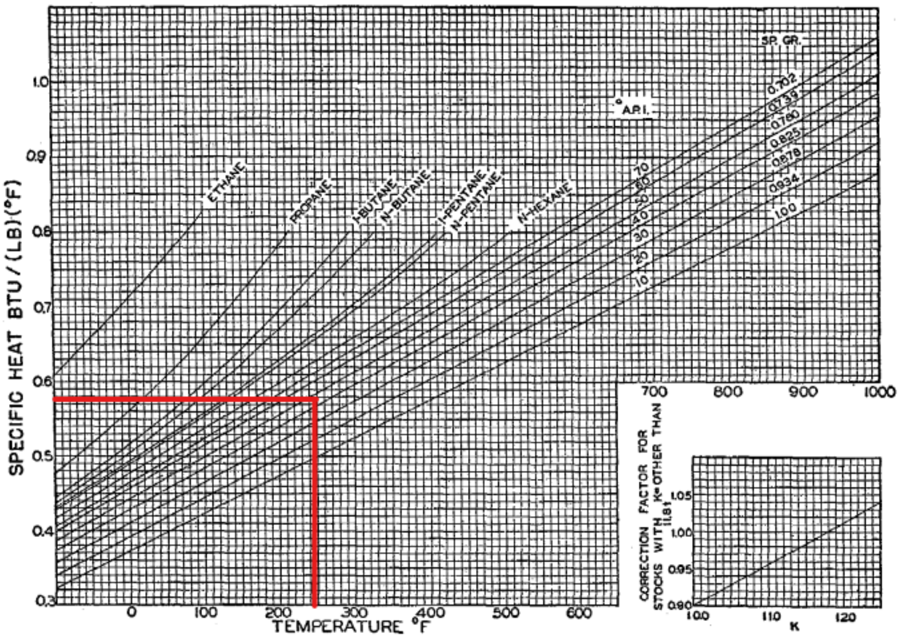


Figure 10: Specific heat capacity of hydrocarbons

We will now employ these values to calculate our ratio between the heat coefficient and the wall coefficient.

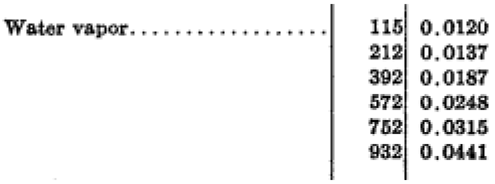
$$\frac{h_0}{\phi_s} = J_H \times \frac{k}{C} \times \left(\frac{C\mu}{k}\right)^{1/3} = 316$$

6.1.8 Tube side calculations for heat coefficient

Since we are using steam on our tube side, we do not need to calculate the heat coefficient since literature advises us to use a value of 1500 always.

Ho = 1500

We find our thermal conductivity value and our specific heat using these tables and figures:

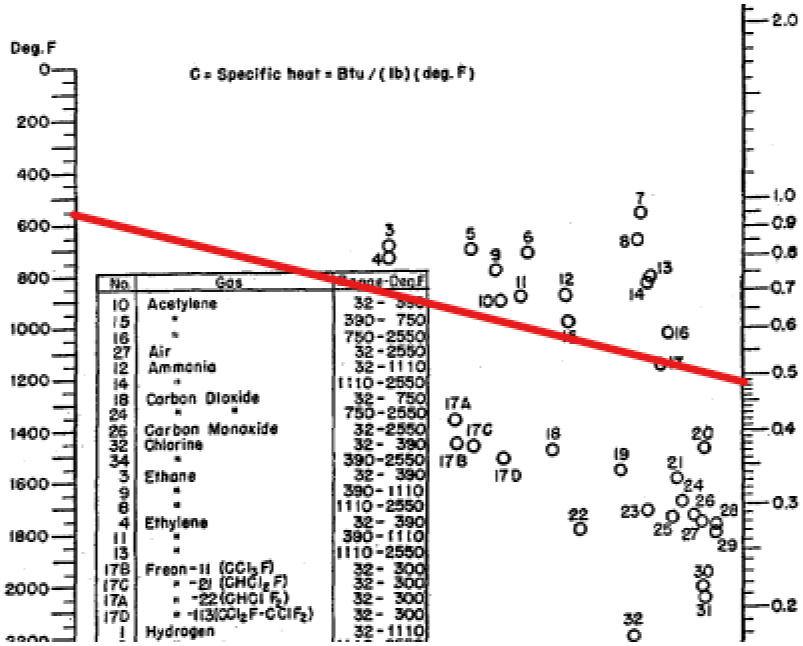


Water vapor...................	115	0.0120
	212	0.0137
	392	0.0187
	572	0.0248
	752	0.0315
	932	0.0441

Figure 11: Thermal Conductivity of Steam

We extrapolate using this data to get a value of:

$$k = 0.042\ \text{W/hr.m}^2$$



No.	Gas	Range-Deg.F
10	Acetylene	32 - 390
15	"	390 - 750
16	"	750 - 2550
27	Air	32 - 2550
12	Ammonia	32 - 1110
14	"	1110 - 2550
18	Carbon Dioxide	32 - 750
24	" "	750 - 2550
26	Carbon Monoxide	32 - 2550
32	Chlorine	32 - 390
34	"	390 - 2550
3	Ethane	32 - 390
9	"	390 - 1110
8	"	1110 - 2550
4	Ethylene	32 - 390
11	"	390 - 1110
13	"	1110 - 2550
17B	Freon-11 (CCl_3F)	32 - 300
17C	" -21 ($CHCl_2F$)	32 - 300
17A	" -22 ($CHClF_2$)	32 - 300
17D	" -113 (CCl_2F-$CClF_2$)	32 - 300
1	Hydrogen	32 - 1110

Figure 12: Specific Heat Capacity of Steam

Specific Heat Capacity = 2.01W/kg.^{o}C

From here on we calculate the value of wall coefficient.

$$\emptyset = 27.22$$

6.1.9 Wall Temperature Calculation

To calculate the wall temperature, we are going to use the formula shown below.

$$t_w = t_c + \frac{\frac{h_o}{\emptyset_s}}{\frac{h_{io}}{\emptyset_t} + \frac{h_o}{\emptyset_s}}(T_c - t_c)$$

$$t_w = 279^o\text{C}$$

Next, we calculate our shell side wall coefficient using wall viscosity.

$$\mu=0.018$$

$$\emptyset_s = (\frac{h_o}{\mu_w})^{0.14}=1.55$$

$$h_{io} = \frac{h_o}{\emptyset_s}\emptyset_s=489.8$$

6.1.10 Overall Coefficient

We now calculate our overall coefficients. Starting with the clean coefficient.

$$U_c = \frac{h_{io}h_o}{h_{io}+h_o} = 2093\text{W/m}^2.^{\text{o}}\text{C}$$

From here on we find our design coefficient.

$$U_D = \frac{Q}{A\Delta t} = 40.40\ \text{W/m}^2.^{\text{o}}\text{C}$$

6.1.11 Shell Side Pressure Drop

We first start by finding our shell side friction factor from literature.

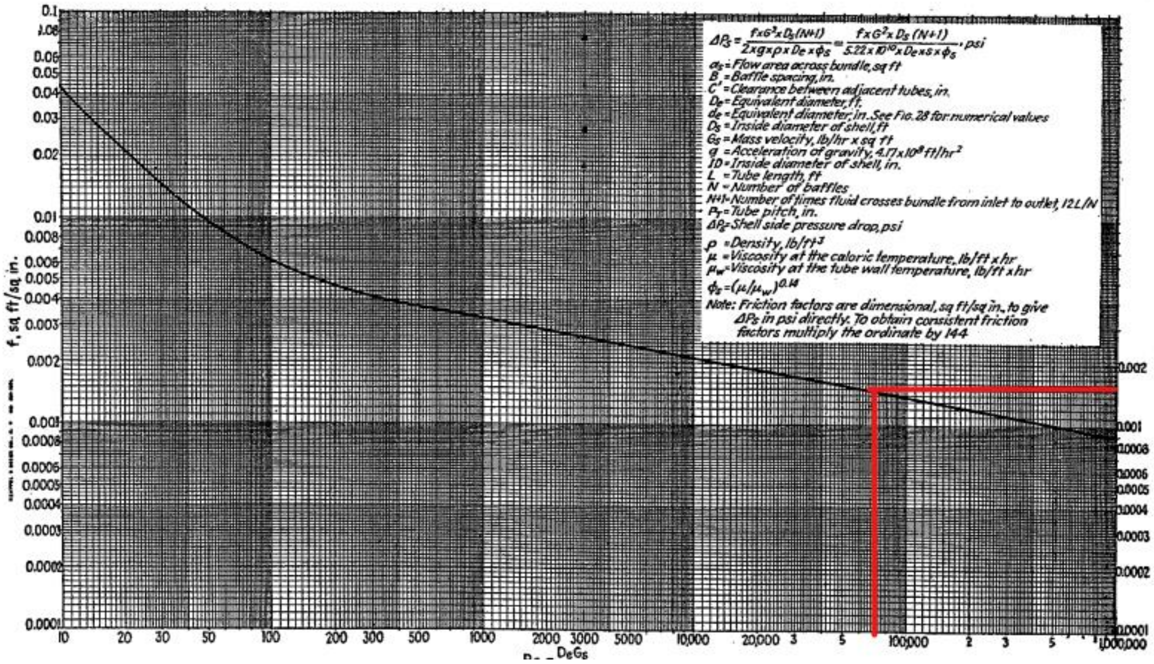


Figure 13: Shell Side Friction Factor

$$f = 0.0015$$

Similarly, we find our specific gravity from literature.

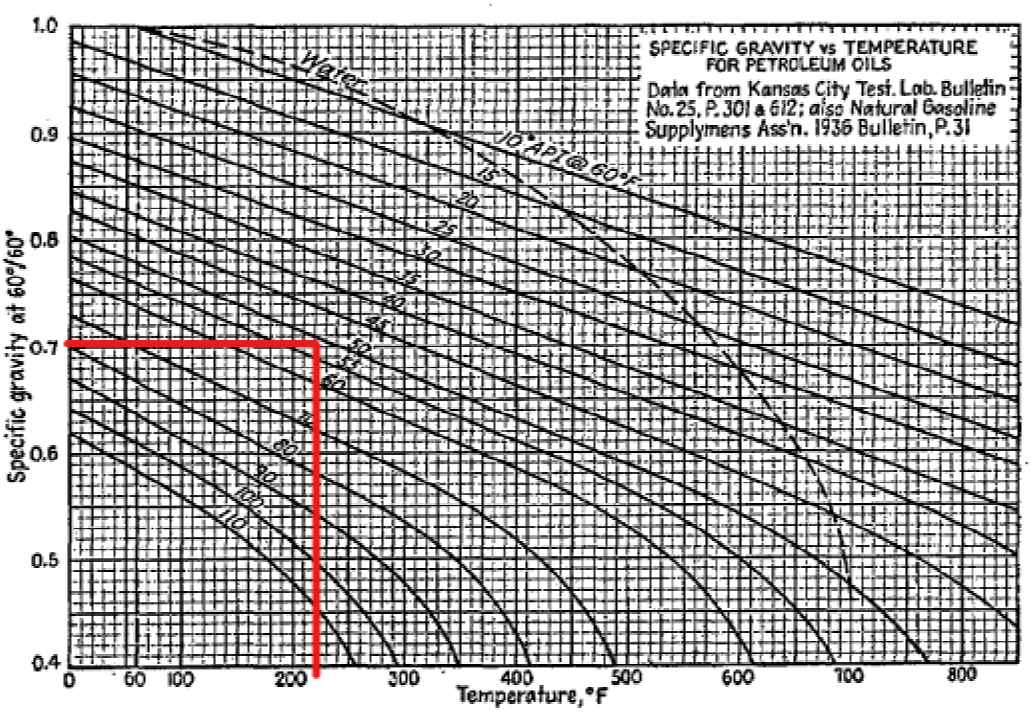


Figure 14: Specific Gravity of Hydrocarbons

$$s = 0.7$$

Next, we calculate the pressure drop using the formula and calculations shown below.

$$\Delta P_s = \frac{f G_s^2 D_s (N+1)}{5.22\times10^{10} D_e s \emptyset_s} = 9.37\text{psi}$$

Allowable pressure drop is 10psi.

6.1.11 Tube Side Pressure Drop

We first start by finding our tube side friction factor from literature.

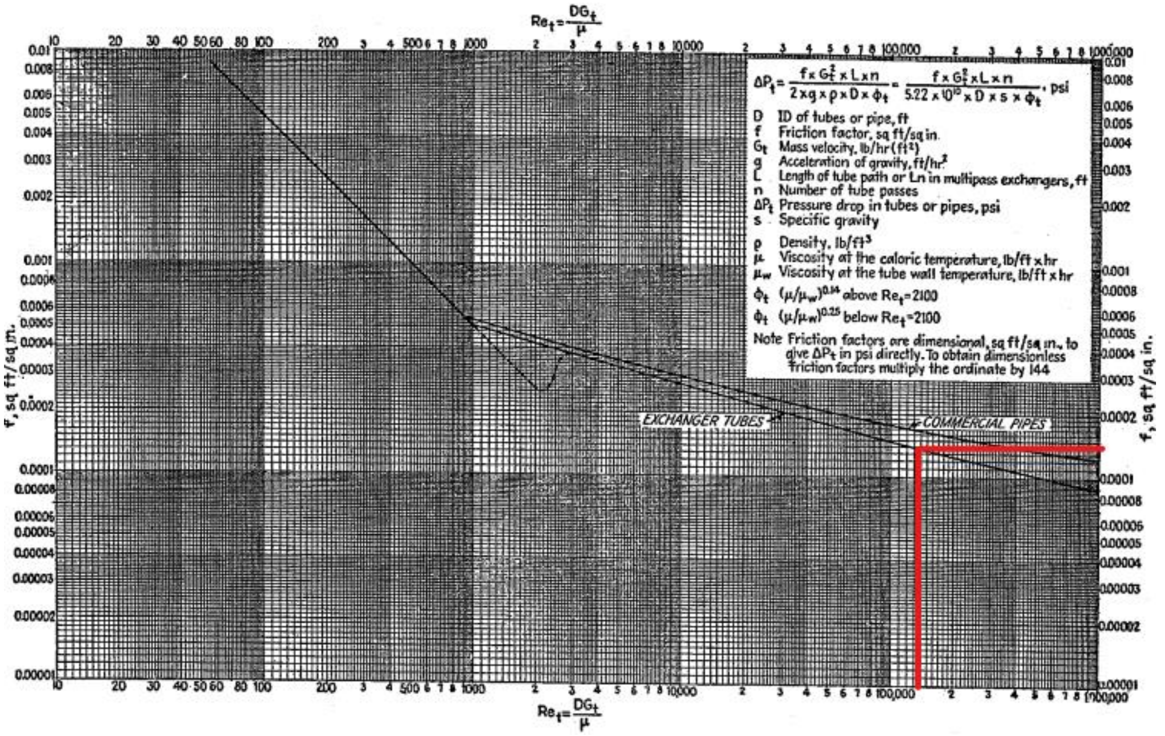


Figure 15: Tube Side Friction Factor

From our steam tables we get a specific gravity of steam of:

$$s = 8.55 \times 10^{-4}$$

Once again, we use the same pressure drop formula to get our drop.

$$\Delta P_t = \frac{f G_S^2 D_S (N+1)}{5.22 \times 10^{10} D_e s \emptyset_S} = 0.35\text{psi}$$

$$\Delta P_T = \Delta P_t + 0.5\Delta P_t = 0.53\text{psi}$$

The allowable pressure drop is 1psi.

6.2 Stabilizer

The stabilizer tower, crux of our project, is the second distillation column in our PFD. It is called a stabilizer tower because it helps us stabilize our NAPHTHA. Stabilizing NAPHTHA refers to the process of removing the lighter ends from our mix before sending the product off to storage or commercial use. The Lighter ends being removed are C1-C4 which will also include our LPG.

6.2.1 Operating Pressure, Bubble point and Dew Point

Before we can proceed with our calculations, it is necessary to calculate our operating pressure, Bubble point, the first temperature at which our liquid starts to vaporize, and our Dew point, the first point at which our fluid starts to condense.

We must estimate operating pressure using the condensing temperatures of our top products.

$$P = 1.2MPa$$

The formula used to calculate our Bubble Point is as follows:

$$\sum y_i = \sum K_i x_i = 1$$

Bubble Point = 278°C

The formula used for Dew Point calculations is as followed:

$$\sum x_i = \sum \frac{y_i}{K_i} = 1$$

Dew Point = 50°C

6.2.3 Minimum Number of Stages

To calculate the minimum number of stages of our Stabilizer, we utilize the Fenske's Equation.

$$N_m = \frac{\log\left(\frac{x_{lk}}{x_{hk}}\right)_D \times \left(\frac{x_{hk}}{x_{lk}}\right)_B}{\log \alpha_{LK}}$$

Nm = 32

6.2.4 Minimum Reflux Ratio

Our next step is to find the minimum reflux ratio of our tower. This is done using the combination of the Underwoods equation.

$$\sum\frac{\alpha_i x_i f}{\alpha_i-\theta} = 1 - q \text{ and } \sum\frac{\alpha_i x_i d}{\alpha_i-\theta} = R_m + 1$$

$$R_m = 2$$

6.2.5 Suitable reflux ratio and number of plates

We will now find the suitable number of plates for our stabilizer along with the suitable reflux ratio. To do so, we will be using the Erbar Maddox method.

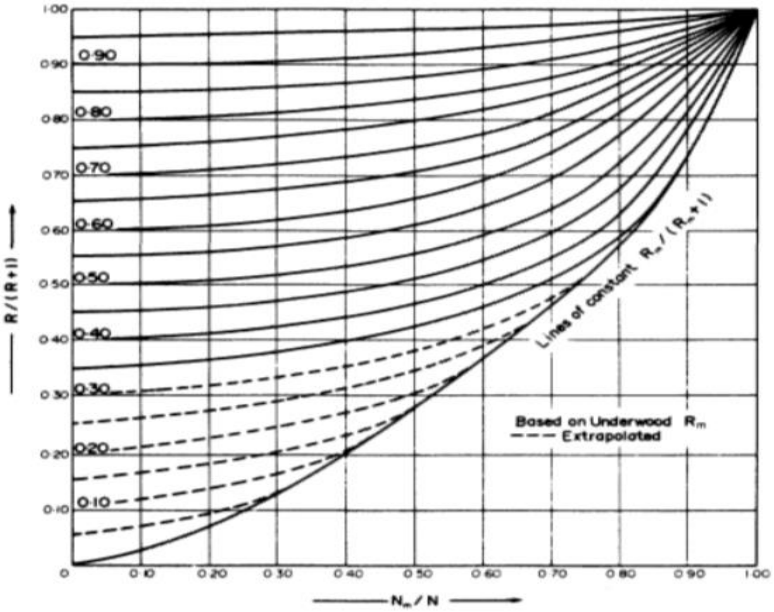


Figure 16: Erbar Maddox diagram

6.2.6 Feed Point Location

Our next goal is to find the location at which the feed enters the stabilizer column. This point is calculated using the Kirkbride Empirical Relation.

$$log\frac{NR}{NS} = 0.206\log\left[\frac{x_{HF}}{x_{LF}} \times \frac{B}{D} \times \left(\frac{x_{LB}}{x_{HD}}\right)^2\right]$$

$$\frac{N_R}{N_S} = 3.53$$

This concluded to us that the best feed point would be the 6th plate from the top of the tower. The entrance is higher up, making our rectifying section shorter than the stripping section. This is ideal since products expected to leave the top are far fewer than the products from the bottom.

6.2.7 Plate efficiency

We will now find the plate efficiency using the O'Connell correlation.

$$\mu \times \alpha_{LK} = 2 \times 0.25 = 0.5$$

Our ideal number of stages are 28, we can now use our efficiency to find the actual number of stages required.

$$\frac{(28-1)}{0.62} = 44$$

6.2.8 Diameter and Height

The information that we do have is:

Plate spacing = 0.5m

Vapour density = 25.38kg/m^3

Liquid Density = 989kg/m^3

We can now find Vapor Superficial Velocity.

$$u_i = (0.171(l_t)^2 + 0.27l_t - 0.047)\left[\frac{(p_L - p_v)}{p_v}\right]^{0.5} = \frac{0.18m}{s}$$

Column diameter.

$$D_c = \sqrt{\frac{4V_w}{\pi p_v U_v}}$$

V$_w$ is 0.83kg/s. hence diameter =2m and height 24m.

6.3 Accumulator

6.3.1 Design pressure

An accumulator is basically a pressure vessel; hence, the pressure is found similar to a pressure vessel. We first find the gauge pressure using the input stream pressure.

Gauge Pressure = 1.21MPa

The absolute pressure is the gauge pressure plus the atmospheric pressure.

Absolute Pressure = 1.21 + 0.1 = 1.31MPa

Operating Pressure = 1.31 + 1.25bar = 1.44MPa

Our design pressure is finally ten percent over the operating pressure.

Design Pressure = 1.44MPa + 10% = 1.58MPa

6.3.2 Length to Diameter Ratio

The L/D ratio of a horizontal separator with a diameter greater than 1m is taken as 3.

6.3.3 Vapor Volumetric Flowrate

First, we get our fraction of total cross-sectional area occupied by liquid. This has to be taken as an educated and literature consulted assumption which we have taken to be 0.5.

Some more data needed for calculations:

Height of Liquid = D/2

Mass Flow of Liquid = 2807kg/hr

Mass Flow of Vapor = 227.3kg/hr

Density of liquid = 499.3kg/m^3

Density of Vapor = 25.38kg/m^3

Tangential Mass Flow = 0.3m^3/s

Actual Mass Flow = 0.045m^3/s

We can now calculate our Vapor Volumetric Flow Rate.

Formula = 0.03 m^3/s

6.3.4 Vapor Residence Time

We will now find our vapor residence time.

$$\frac{h}{u} = \frac{0.5D}{0.0454}$$

Actual Residence time is equal to the vapor velocity.

$$\frac{Lv}{U_v} = \frac{3D}{0.078D^{-2}}$$

Residence time is equal to actual residence time.

$$11.02D = 38.46D^3$$

6.3.5 Design Specifications

Solving these equations together gives us our diameter.

Diameter = 0.535m

Length of Accumulator = 1.85m

Holdup Volume = 0.27m3

Holdup Time = 3 minutes

CHAPTER 7

SIMULATION

7.1 Properties:

This chapter focuses on the simulation of a Naphtha Stabilization Unit, an integral part of petroleum refining processes. Through advanced simulation techniques, this research aims to model and optimize the Naphtha Stabilization Unit, providing valuable insights for process design, operation, and performance improvement.

7.1.1 Fluid Package Used: Peng Robinson EOS

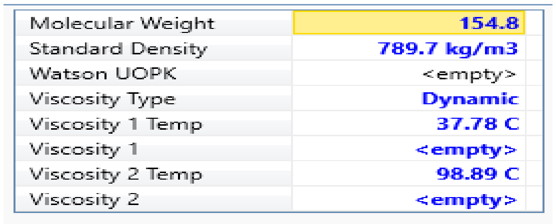
To accurately simulate the behavior of the Naphtha Stabilization Unit, an appropriate fluid package must be selected. In this study, the Peng Robinson equation of state (EOS) has been chosen as the fluid package for the simulation.

7.2 Assay Input

To generate a representative petroleum assay for the naphtha stream in the simulation of the Naphtha Stabilization Unit, we can provide an example set of input data. Please note that these values are fictional and should be replaced with actual data or representative values obtained from laboratory analysis or industry sources for your specific naphtha stream.

7.2.1 Input Properties:

7.2.1.1 Bulk Properties:



Molecular Weight	154.8
Standard Density	789.7 kg/m3
Watson UOPK	<empty>
Viscosity Type	Dynamic
Viscosity 1 Temp	37.78 C
Viscosity 1	<empty>
Viscosity 2 Temp	98.89 C
Viscosity 2	<empty>

Figure 17: Bulk Properties of Crude

7.2.1.2 Light Ends:

These values represent a sample set of mole fractions for various hydrocarbon components in the light end. The percentages indicate the relative abundance of each component within the stream.

Table 25: Light End Properties of Crude Oil

Components	**Mole Fraction**	**Normal Boiling Point (C)**
Methane	0.000	-161.52
Ethane	4.50e-003	-88.60
Propane	1.67e-002	-11.73
i-Butane	1.87e-002	-42.10
n-Butane	1.98e-002	-0.50
H2O	0.000	99.99
NaCl	0.000	326.8

7.2.1.3 Distillation Curve Data:

Assay Basis: Molar

Table 26 True Boiling Point Curve Data

Assay Percent	Temperature [C]
00.00	26.67
10.00	123.9
20.00	176.1
30.00	221.1
40.00	275.0
50.00	335.0
60.00	399.4
70.00	290.6
80.00	590.6
90.00	691.7
98.00	765.6

7.2.1.4 Molecular Weight Curve Data:

Table 27 Molecular Weight Curve Data

Assay Percent	Molecular Wt.
00.00	68.00
10.00	119.0
20.00	150.0
30.00	182.0
40.00	225.0
50.00	282.0
60.00	350.0
70.00	456.0
80.00	585.0
90.00	713.0
98.00	838.0

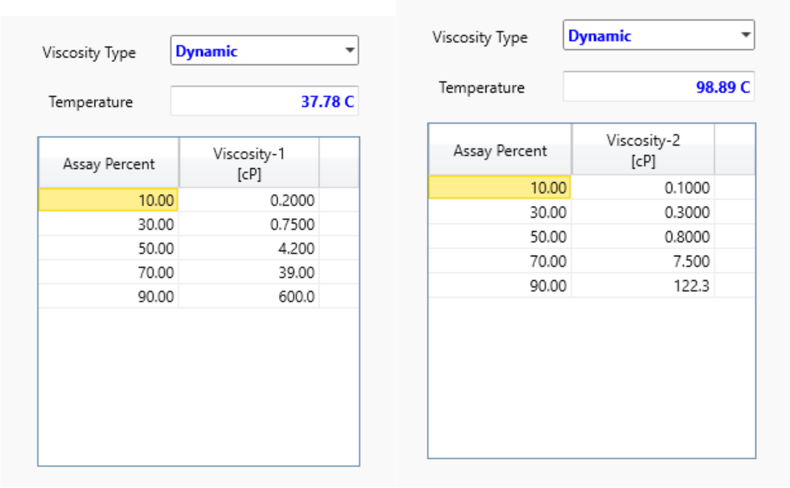
7.2.1.5 Density Curve Data:

Table 28 Density Curve Data

Assay Percent	Density [kg/m3]
13.00	68.00
33.00	119.0
57.00	150.0
74.00	182.0
91.00	225.0

7.2.1.6 Viscosity Curves Data:

Table 29 Viscosities Data at different Assay percentages



Viscosity Type: Dynamic

Temperature: 37.78 C

Assay Percent	Viscosity-1 [cP]
10.00	0.2000
30.00	0.7500
50.00	4.200
70.00	39.00
90.00	600.0

Viscosity Type: Dynamic

Temperature: 98.89 C

Assay Percent	Viscosity-2 [cP]
10.00	0.1000
30.00	0.3000
50.00	0.8000
70.00	7.500
90.00	122.3

7.3 Output Blend:

After simulating the provided input data and the selected fluid package (Peng Robinson), the calculated output blend can be analyzed. The output blend refers to the resulting composition of the naphtha stream after undergoing the stabilization process.

7.3.1 Calculated Composition and Properties of Crude Oil:

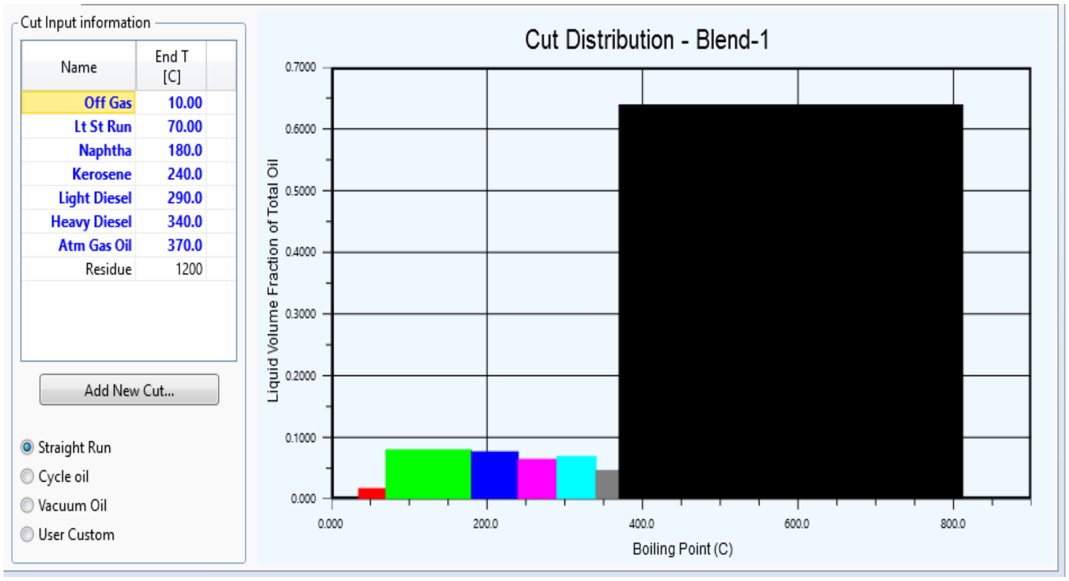
Table 30 The Calculated Crude Oil Composition and Properties

Comp. Name	NBP [C]	Molecular Weight	Density [kg/m3]	Viscosity 1 [cP]	Viscosity 2 [cP]
NBP_34	34.03	53.15	653.89	0.09	0.06
NBP_47	46.96	53.48	677.82	0.10	0.06
NBP_61	61.37	54.20	679.45	0.11	0.07
NBP_75	75.28	55.27	681.36	0.12	0.07
NBP_89	88.97	56.73	683.70	0.13	0.08
NBP_103	102.79	58.27	686.26	0.15	0.08
NBP_117	116.61	60.02	689.13	0.18	0.09
NBP_130	130.44	62.08	692.36	0.22	0.11
NBP_144	144.26	64.56	696.07	0.26	0.12

Table 30 (cont'd)

NBP_158	158.08	67.67	700.33	0.31	0.14
NBP_172	171.64	71.47	705.07	0.38	0.17
NBP_185	185.39	75.71	709.97	0.46	0.20
NBP_199	199.23	80.30	714.85	0.55	0.23
NBP_213	212.88	85.27	719.65	0.67	0.27
NBP_227	226.67	90.29	724.16	0.81	0.32
NBP_241	240.60	95.20	728.21	0.99	0.35
NBP_254	254.38	100.07	731.89	1.21	0.39
NBP_268	268.20	104.75	735.45	1.49	0.44
NBP_282	282.01	109.40	738.93	1.83	0.49
NBP_296	295.82	114.30	742.38	2.27	0.54
NBP_310	309.61	119.44	745.81	2.81	0.61
NBP_323	323.49	125.02	749.24	3.49	0.70
NBP_337	337.29	131.44	752.77	4.37	0.83
NBP_351	351.02	138.61	756.43	5.51	1.08
NBP_365	364.79	146.00	760.12	6.95	1.34
NBP_379	378.56	153.40	763.70	8.71	1.71
NBP_392	392.29	160.60	766.97	10.81	2.18
NBP_406	406.15	167.41	770.01	13.20	2.68
NBP_420	420.02	173.85	772.94	15.91	3.20
NBP_441	440.70	183.08	777.30	20.77	4.13
NBP_468	468.50	195.20	783.37	29.55	5.76
NBP_496	496.30	207.05	789.82	41.98	8.06
NBP_524	524.01	218.38	796.87	59.93	11.51
NBP_552	551.79	229.35	804.79	85.95	16.57
NBP_580	579.52	240.32	813.49	123.90	24.06
NBP_607	607.29	251.34	822.79	179.78	35.26
NBP_635	635.07	262.68	832.79	263.28	52.24
NBP_682	682.34	282.97	851.72	521.01	105.72
NBP_739	739.04	310.62	879.68	1329.48	277.35

7.3.2 Cut Distribution:



Cut Input information

Name	End T [C]
Off Gas	10.00
Lt St Run	70.00
Naphtha	180.0
Kerosene	240.0
Light Diesel	290.0
Heavy Diesel	340.0
Atm Gas Oil	370.0
Residue	1200

● Straight Run
○ Cycle oil
○ Vacuum Oil
○ User Custom

Figure 18 Out Put Blend Cut Distribution

7.4 Simulation of Complete Plant:

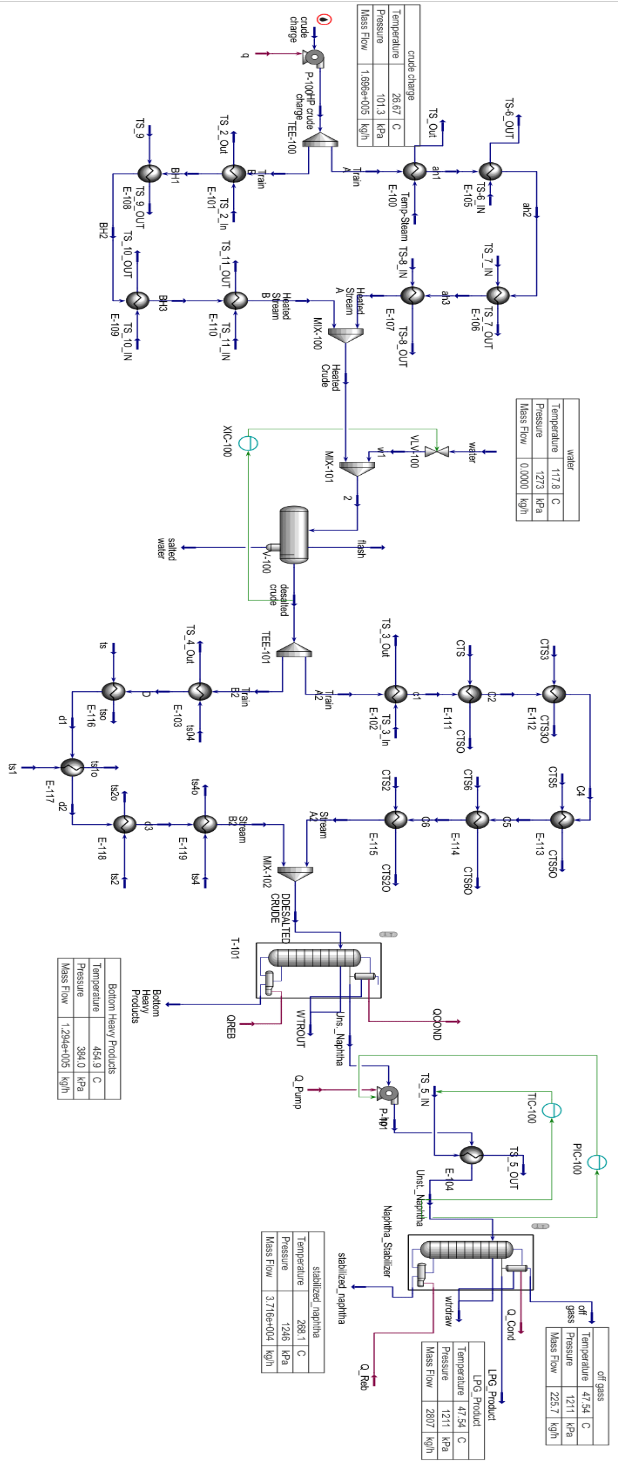


Figure 19: Simulation of Complete Plant

7.4.1 Desalter(V-100):

The simulation focused on the implementation of a three-phase separator configuration for the desalter. The three-phase separator was employed to facilitate the efficient separation of crude oil, water, and salts. By modeling the electrostatic desalter and incorporating the three-phase separator, the simulation aimed to optimize the desalting process, ensuring effective removal of impurities and enhancing the quality of the crude oil product. The simulation results provide insights into the performance and efficiency of the desalter, contributing to the optimization and improvement of crude oil desalting operations in the petroleum refining industry.

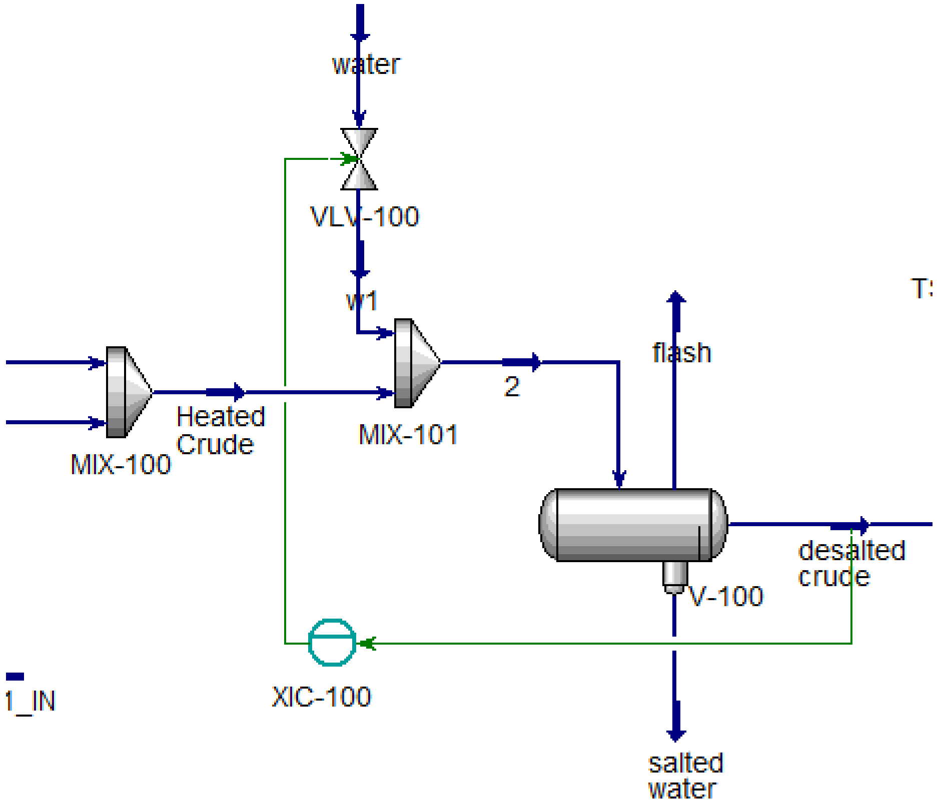
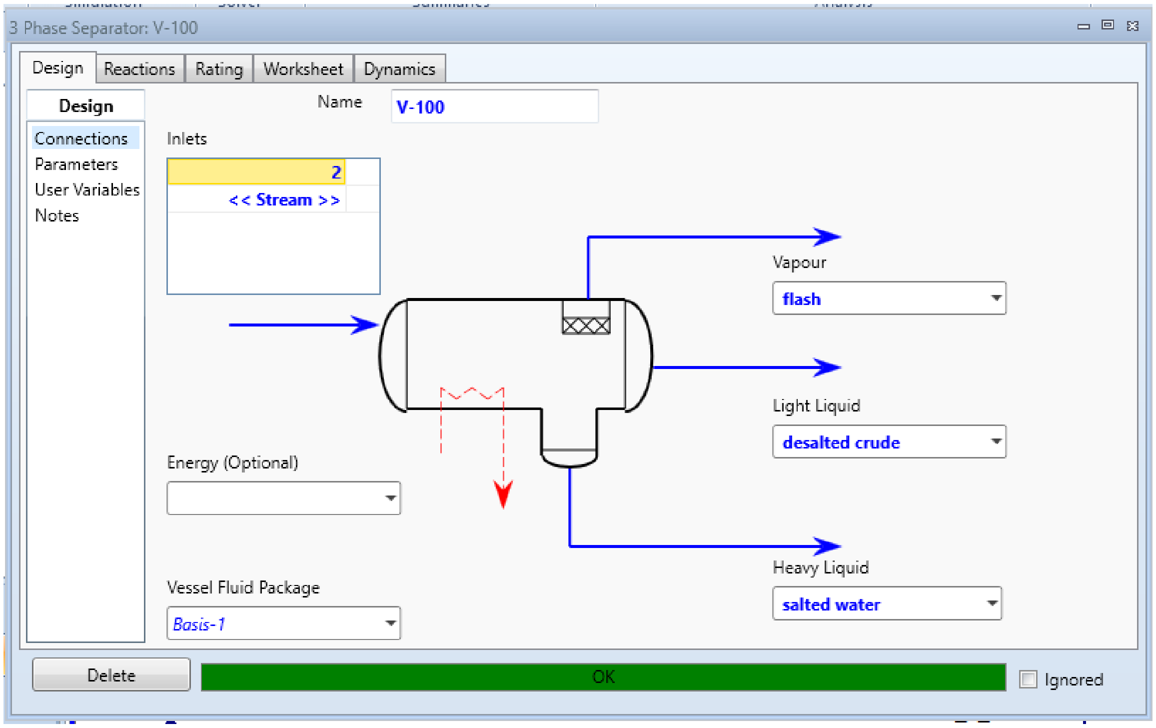


Figure 20: Desalter Simulation



Name: V-100

Inlets: 2

Vapour: flash

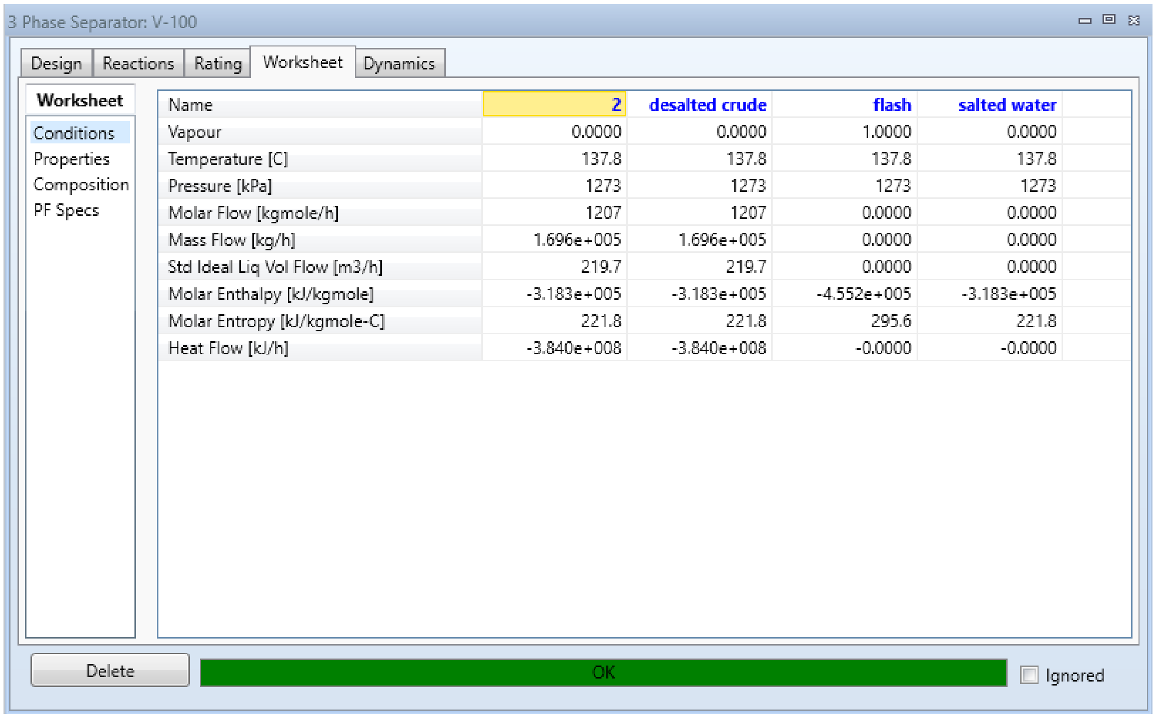
Light Liquid: desalted crude

Heavy Liquid: salted water

Energy (Optional):

Vessel Fluid Package: Basis-1

Figure 21: Desalter Connections

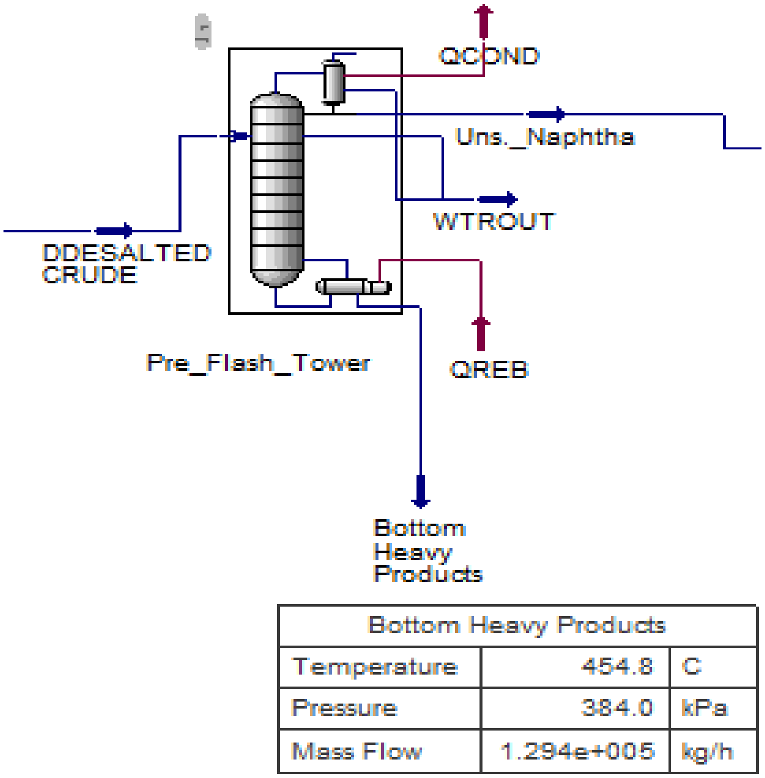


Name	2	desalted crude	flash	salted water
Vapour	0.0000	0.0000	1.0000	0.0000
Temperature [C]	137.8	137.8	137.8	137.8
Pressure [kPa]	1273	1273	1273	1273
Molar Flow [kgmole/h]	1207	1207	0.0000	0.0000
Mass Flow [kg/h]	1.696e+005	1.696e+005	0.0000	0.0000
Std Ideal Liq Vol Flow [m3/h]	219.7	219.7	0.0000	0.0000
Molar Enthalpy [kJ/kgmole]	-3.183e+005	-3.183e+005	-4.552e+005	-3.183e+005
Molar Entropy [kJ/kgmole-C]	221.8	221.8	295.6	221.8
Heat Flow [kJ/h]	-3.840e+008	-3.840e+008	-0.0000	-0.0000

Figure 22: Desalter Worksheet

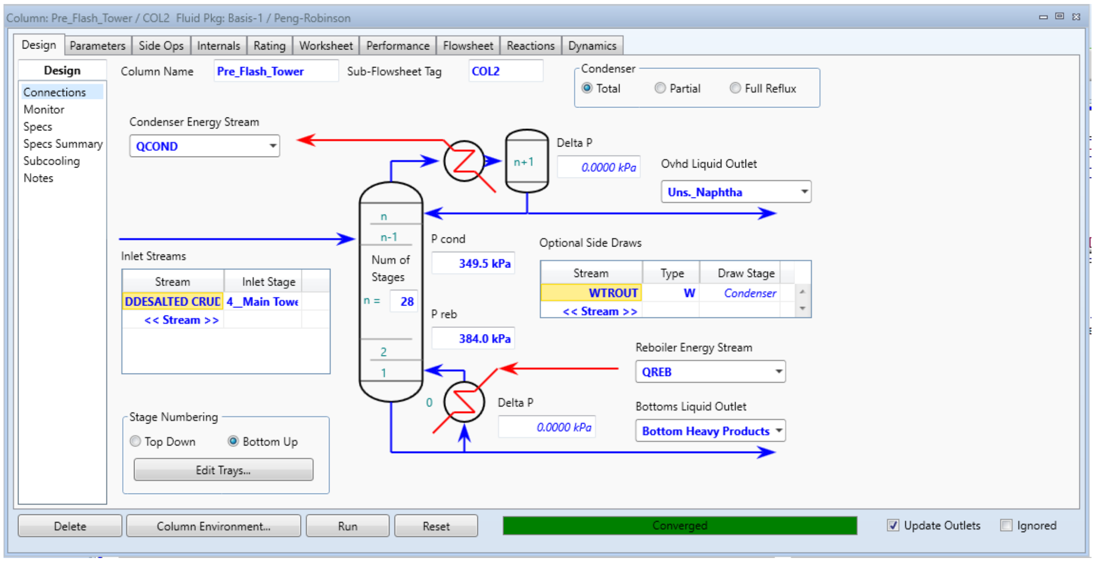
7.4.2 Pre-Flash Tower:

The preflash tower serves as a crucial unit for the separation of lighter components from the crude oil feedstock before entering the main distillation column. The simulation focused on modeling the preflash tower and its associated equipment, such as heat exchangers and reflux systems. By accurately representing the thermodynamics and unit operations within the preflash tower, the simulation aimed to optimize the separation process, ensuring the efficient removal of volatile components from the crude oil. The simulation results provide valuable insights into the performance and efficiency of the preflash tower, aiding in the design and operation of this critical unit in petroleum refining processes.



Bottom Heavy Products		
Temperature	454.8	C
Pressure	384.0	kPa
Mass Flow	1.294e+005	kg/h

Figure 23: Simulation Pre-Flash Tower



Column: Pre_Flash_Tower / COL2 Fluid Pkg: Basis-1 / Peng-Robinson

Column Name: Pre_Flash_Tower
Sub-Flowsheet Tag: COL2
Condenser: ● Total ○ Partial ○ Full Reflux
Condenser Energy Stream: QCOND
Delta P: 0.0000 kPa
Ovhd Liquid Outlet: Uns._Naphtha
Num of Stages: n = 28
P cond: 349.5 kPa
P reb: 384.0 kPa
Reboiler Energy Stream: QREB
Delta P: 0.0000 kPa
Bottoms Liquid Outlet: Bottom Heavy Products
Stage Numbering: ○ Top Down ● Bottom Up
☑ Update Outlets ☐ Ignored

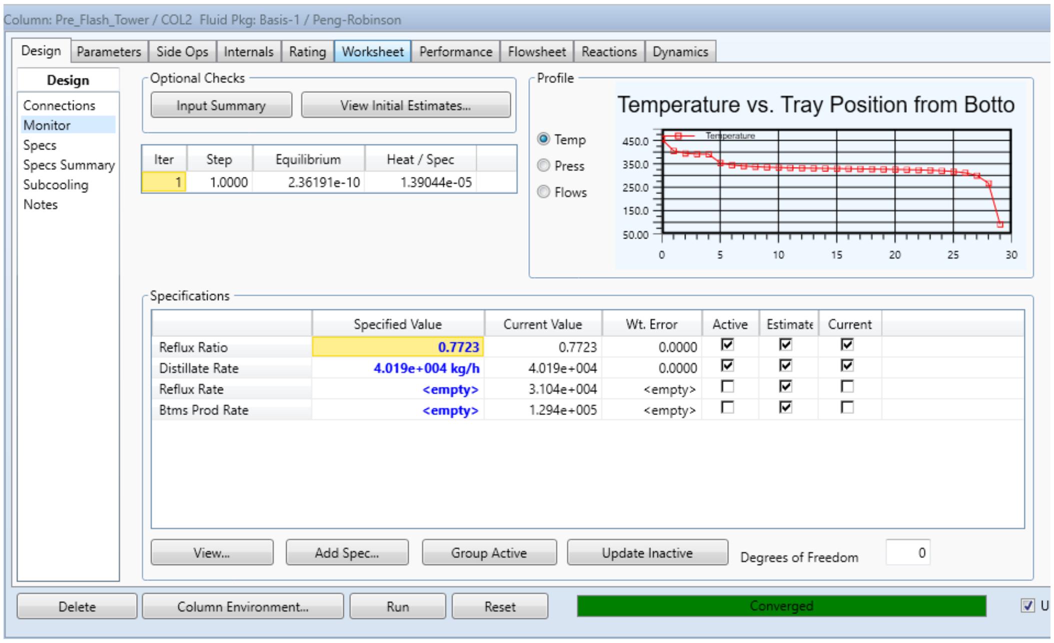
Inlet Streams

Stream	Inlet Stage
DDESALTED CRUD	4__Main Towe
<< Stream >>	

Optional Side Draws

Stream	Type	Draw Stage
WTROUT	W	Condenser
<< Stream >>		

Figure 24: Connections of Pre-Flash Tower



Column: Pre_Flash_Tower / COL2 Fluid Pkg: Basis-1 / Peng-Robinson

Profile: ● Temp ○ Press ○ Flows

Iter	Step	Equilibrium	Heat / Spec
1	1.0000	2.36191e-10	1.39044e-05

Specifications

	Specified Value	Current Value	Wt. Error	Active	Estimate	Current
Reflux Ratio	0.7723	0.7723	0.0000	☑	☑	☑
Distillate Rate	4.019e+004 kg/h	4.019e+004	0.0000	☑	☑	☑
Reflux Rate	<empty>	3.104e+004	<empty>	☐	☑	☐
Btms Prod Rate	<empty>	1.294e+005	<empty>	☐	☑	☐

Degrees of Freedom: 0

Figure 25: Specifications of Pre-Flash Tower

7.4.3 Naphtha Stabilizer:

The Naphtha Stabilization Unit simulation involved the modeling of a distillation column, reboiler, condenser, and associated heat exchangers. The distillation column was designed to separate the volatile components, such as light hydrocarbons, from the naphtha stream, thereby stabilizing it to meet the desired specifications.

By accurately representing the fluid properties and employing appropriate thermodynamic models, such as the Peng Robinson equation of state, the simulation aimed to predict the phase behavior, vapor-liquid equilibrium, and temperature profiles within the Naphtha Stabilization Unit. Additionally, the simulation accounted for factors such as column pressure, feed composition, and reflux ratio to optimize the separation efficiency and product quality.

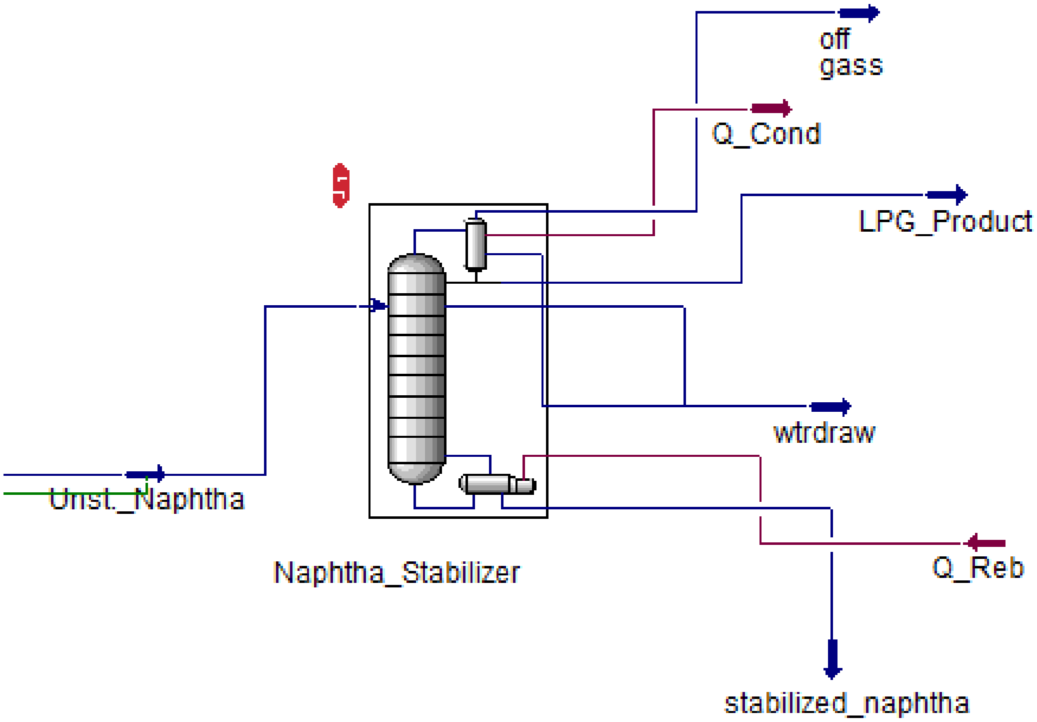
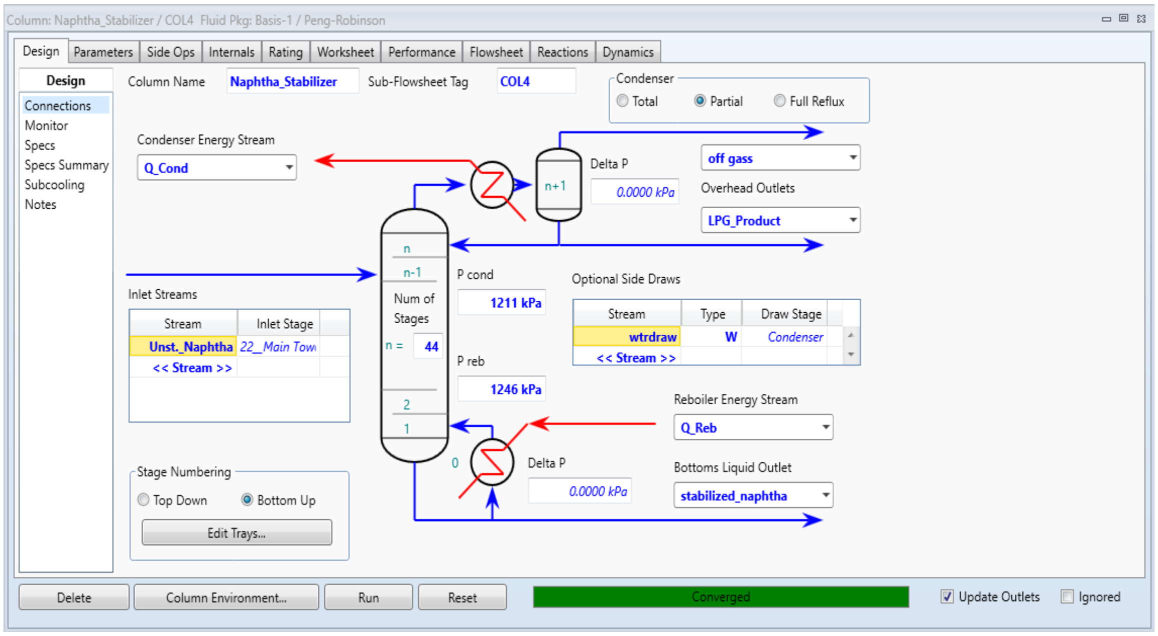


Figure 26: Simulation Naphtha Stabilizer



Column: Naphtha_Stabilizer / COL4 Fluid Pkg: Basis-1 / Peng-Robinson

Column Name: Naphtha_Stabilizer

Sub-Flowsheet Tag: COL4

Condenser: ○ Total ● Partial ○ Full Reflux

Condenser Energy Stream: Q_Cond

Delta P: 0.0000 kPa

Overhead Outlets: off gass, LPG_Product

P cond: 1211 kPa

n = 44

P reb: 1246 kPa

Inlet Streams

Stream	Inlet Stage
Unst._Naphtha	22__Main Tow
<< Stream >>	

Optional Side Draws

Stream	Type	Draw Stage
wtrdraw	W	Condenser
<< Stream >>		

Reboiler Energy Stream: Q_Reb

Delta P: 0.0000 kPa

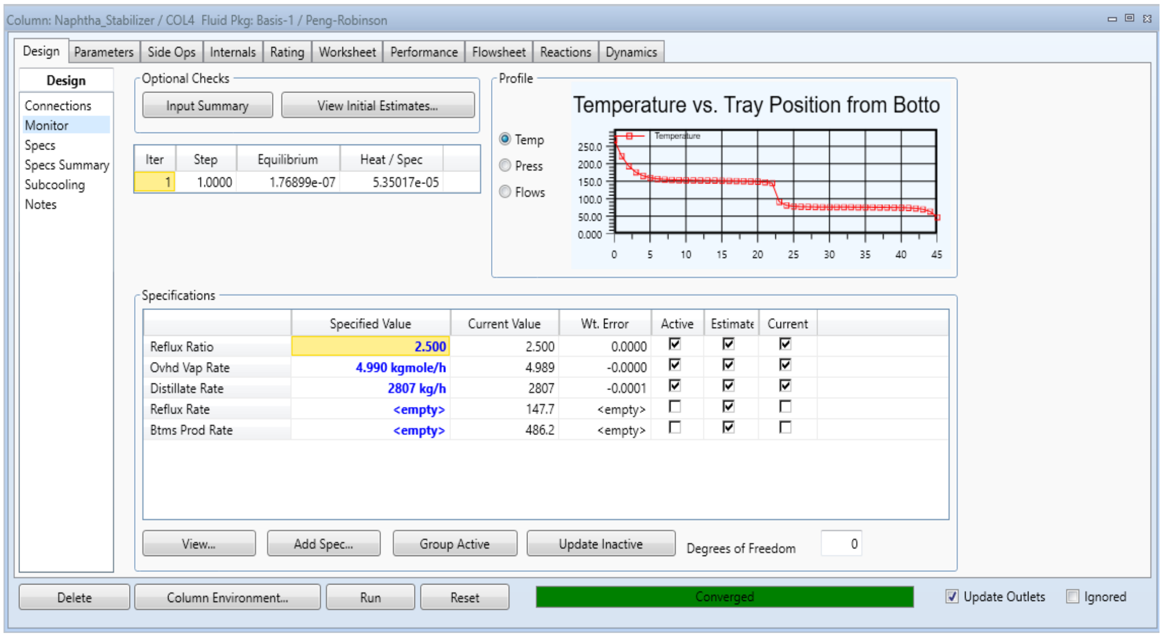
Bottoms Liquid Outlet: stabilized_naphtha

Stage Numbering: ○ Top Down ● Bottom Up

Converged

☑ Update Outlets ☐ Ignored

Figure 27: Connections of Naphtha Stabilizer



Column: Naphtha_Stabilizer / COL4 Fluid Pkg: Basis-1 / Peng-Robinson

Iter	Step	Equilibrium	Heat / Spec
1	1.0000	1.76899e-07	5.35017e-05

Profile: ● Temp ○ Press ○ Flows

Specifications

	Specified Value	Current Value	Wt. Error	Active	Estimate	Current
Reflux Ratio	2.500	2.500	0.0000	☑	☑	☑
Ovhd Vap Rate	4.990 kgmole/h	4.989	-0.0000	☑	☑	☑
Distillate Rate	2807 kg/h	2807	-0.0001	☑	☑	☑
Reflux Rate	<empty>	147.7	<empty>	☐	☑	☐
Btms Prod Rate	<empty>	486.2	<empty>	☐	☑	☐

Degrees of Freedom: 0

Converged

☑ Update Outlets ☐ Ignored

Figure 28: Specifications of Naphtha Stabilizer

7.4.4 TEE(s) Splitter:

In the simulation, tees were employed to split the process stream into two equal flow rates. This approach aimed to ensure efficient heating and optimize the performance of the heat exchangers, taking into account their sizes and the required heat energy (duty). By achieving a balanced flow distribution, the tees facilitated effective utilization of the heat transfer area, preventing uneven heating and maximizing heat transfer efficiency. This stream splitting technique played a crucial role in enhancing the overall performance of the heat exchangers and promoting efficient heating in the Naphtha Stabilization Unit.

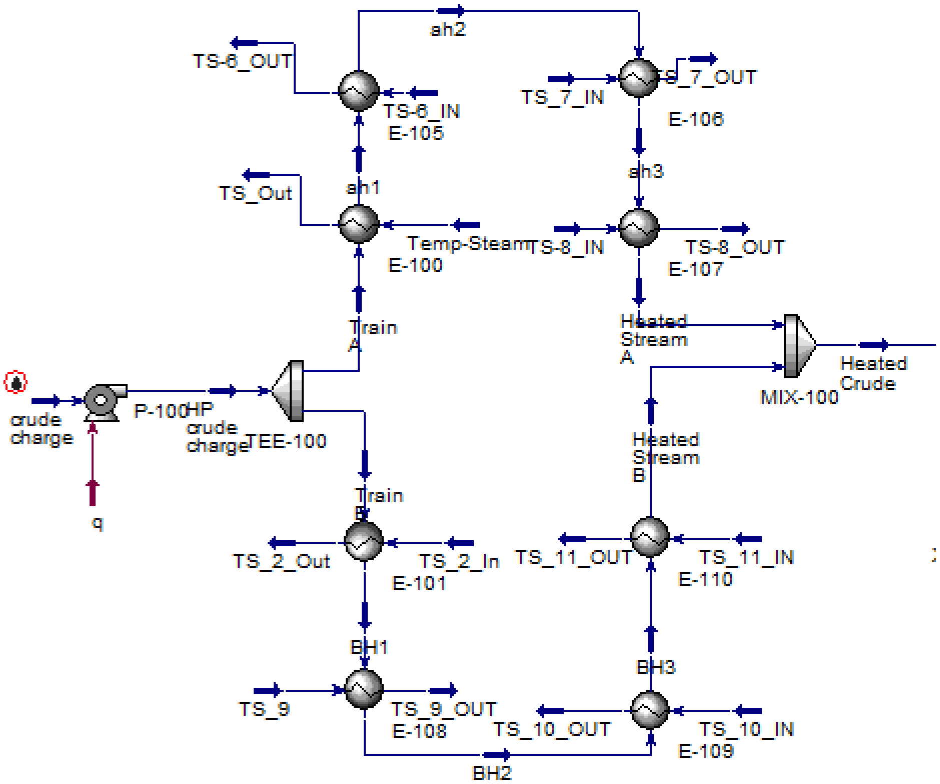
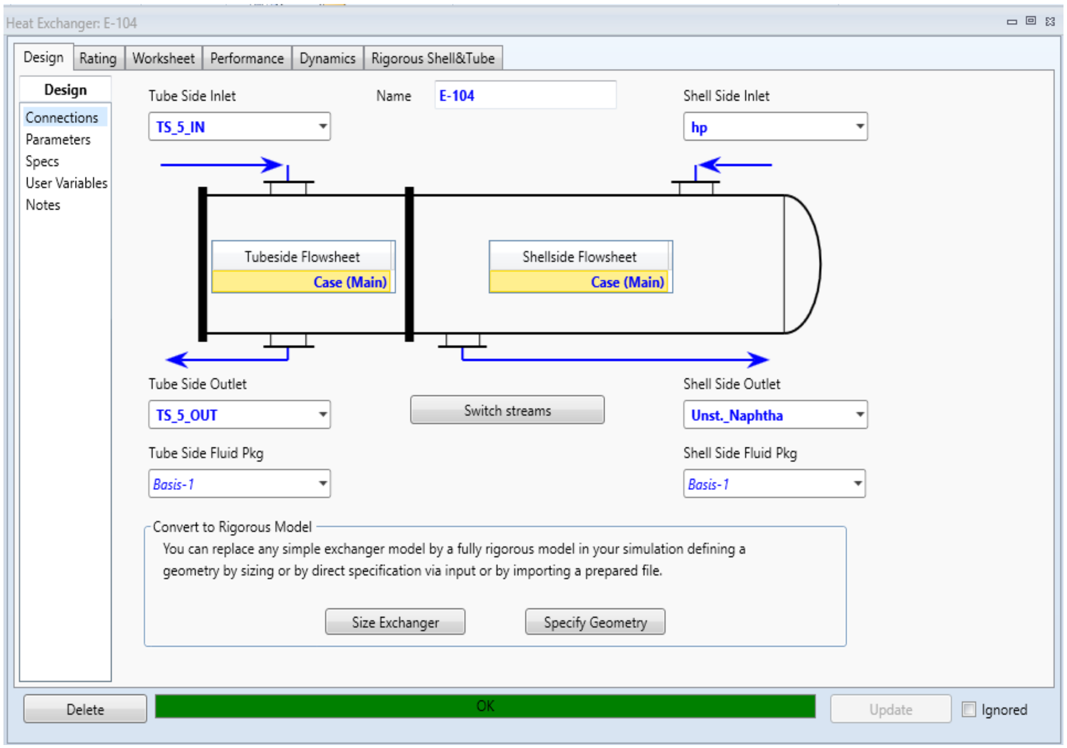


Figure 29: TEE (Splitting of Process Stream)

7.4.5 Heat Exchangers:

In the simulation, TEMA type shell and tube heat exchangers were utilized for efficient heat transfer. These exchangers feature a shell that houses a bundle of tubes, allowing effective heat exchange between process streams. Temporary steam was used as the heat source, which can be replaced by other process streams from different refinery units for heating purposes. The simulation optimized heat transfer, thermal efficiency, and energy utilization within the unit, contributing to improved heat integration and energy-saving opportunities in the refinery.

A typical heat exchanger's parameters and summary are shown below, providing an overview of its key characteristics and performance:



Heat Exchanger: E-104

Tube Side Inlet: TS_5_IN

Name: E-104

Shell Side Inlet: hp

Tube Side Outlet: TS_5_OUT

Shell Side Outlet: Unst._Naphtha

Tube Side Fluid Pkg: Basis-1

Shell Side Fluid Pkg: Basis-1

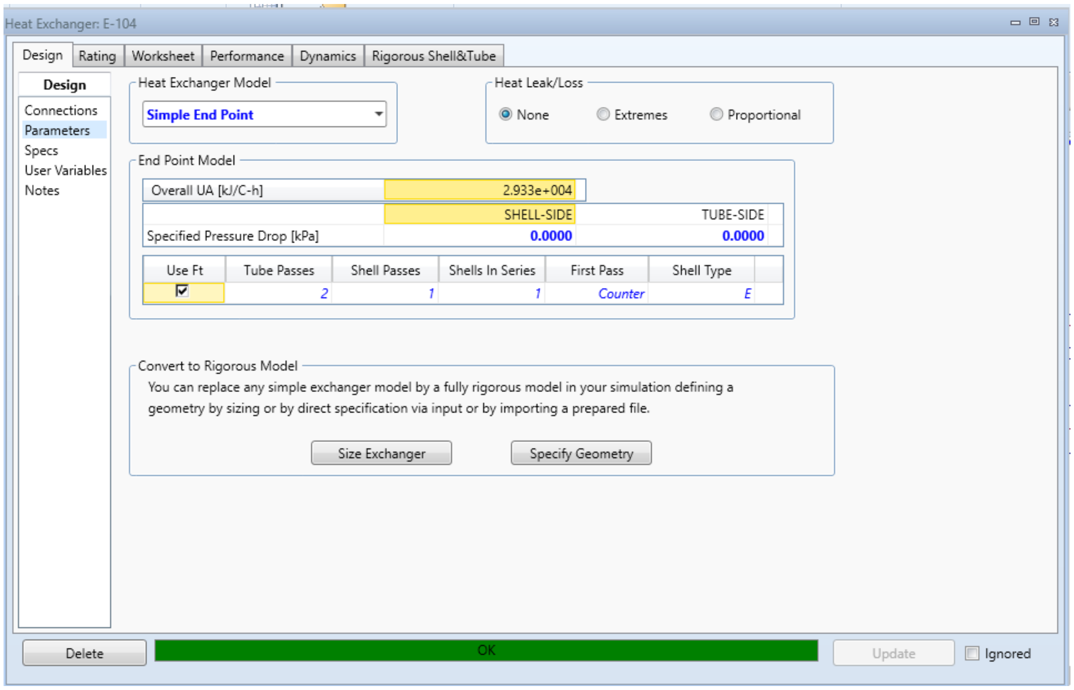
Convert to Rigorous Model

You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

OK

☐ Ignored

Figure 30: Heat Exchanger Connections



Heat Exchanger Model: Simple End Point

Heat Leak/Loss: ● None ○ Extremes ○ Proportional

End Point Model

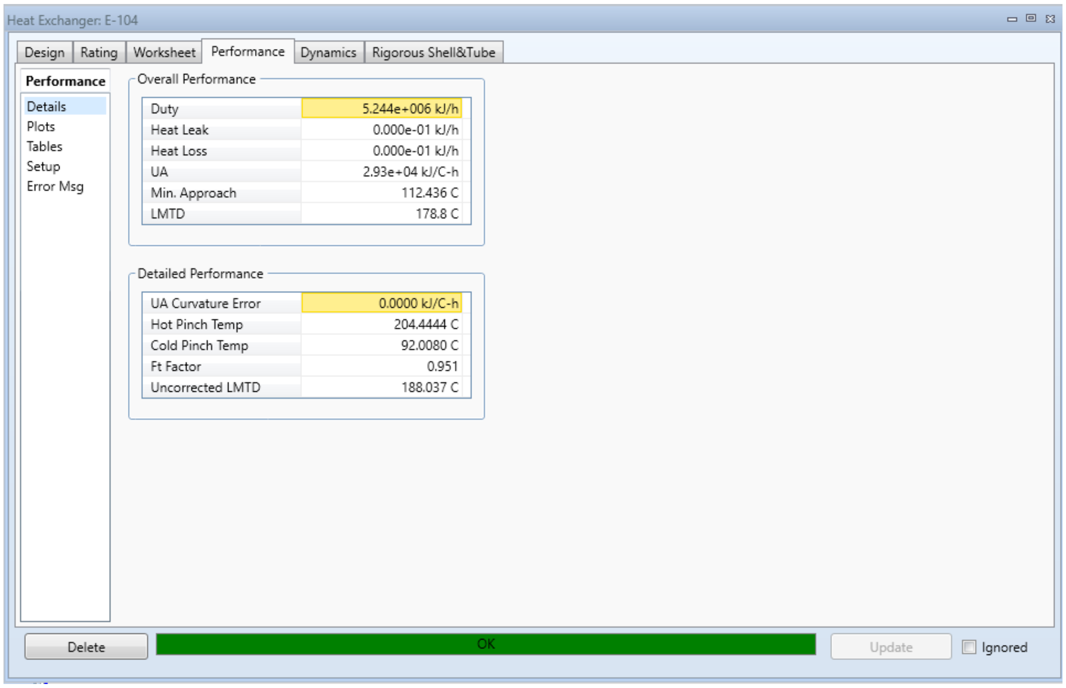
Overall UA [kJ/C-h]	2.933e+004	
	SHELL-SIDE	TUBE-SIDE
Specified Pressure Drop [kPa]	0.0000	0.0000

Use Ft	Tube Passes	Shell Passes	Shells In Series	First Pass	Shell Type
☑	2	1	1	Counter	E

Convert to Rigorous Model

You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

Figure 31: Parameters of Heat Exchanger



Overall Performance

Duty	5.244e+006 kJ/h
Heat Leak	0.000e-01 kJ/h
Heat Loss	0.000e-01 kJ/h
UA	2.93e+04 kJ/C-h
Min. Approach	112.436 C
LMTD	178.8 C

Detailed Performance

UA Curvature Error	0.0000 kJ/C-h
Hot Pinch Temp	204.4444 C
Cold Pinch Temp	92.0080 C
Ft Factor	0.951
Uncorrected LMTD	188.037 C

Figure 32: Performance Details of Heat Exchanger

CHAPTER 8

ECONOMIC ANALYSIS

8.1 Costing and Evaluation

Costing and Evaluation is an integral part of the preliminary work done before determining whether a new project or process is feasible or not. In costing and evaluation, all implications related to the financials of the project or process are identified, they are reported and analyzed to give recommendations regarding the project. In this process, a very detailed and comprehensive assessment of firstly, various cost factors are done like the total purchase cost of equipment, physical plant cost and using these the total capital investment required is calculated. Then, operating costs are calculated by summing fixed operating costs like maintenance, labor, insurance etc., and variable costs like Raw Material costs, utilities costs and other miscellaneous costs. Moving further, the revenues are calculated by adding the revenues from different products sold by the plant and these are then subtracted from the annual operating costs to calculate the net profit. Finally, the total investment required for the project is divided by the net profit to calculate the payback period which is itself one of the most important factors while determining the feasibility and value in terms of money for the project. Many different business decisions are based on costing and evaluation for instance amount and timeframe for any loans required, human resource required for the project, selling prices of the products, salary packages for employees, stockholder dividends (if necessary) etc. Overall, since nearly every industry and organization are looking to earn profit, costing and evaluation becomes essential in determining whether a new process, product, project, or service is viable in terms of financials or not.

In our Project, we utilized the Coulson and Richardson method with a time base of Mid-2004 to cost and evaluate our project for the determination of its economic feasibility.

Firstly, the costs of all equipment were calculated separately, then factors for other different costs were included to give the fixed capital required, then annual operating cost was calculated and used with the revenues that were calculated, which finally yielded us with the payback period.

8.2 Costing of Stabilizer

Table 31: Design Specifications of the Stabilizer

Parameters	Value
Diameter/m	2
Height/m	24
Operating Pressure/ bar	12
No. of trays	44
Tray Type	Valve

8.2.1 Bare Vessel Cost

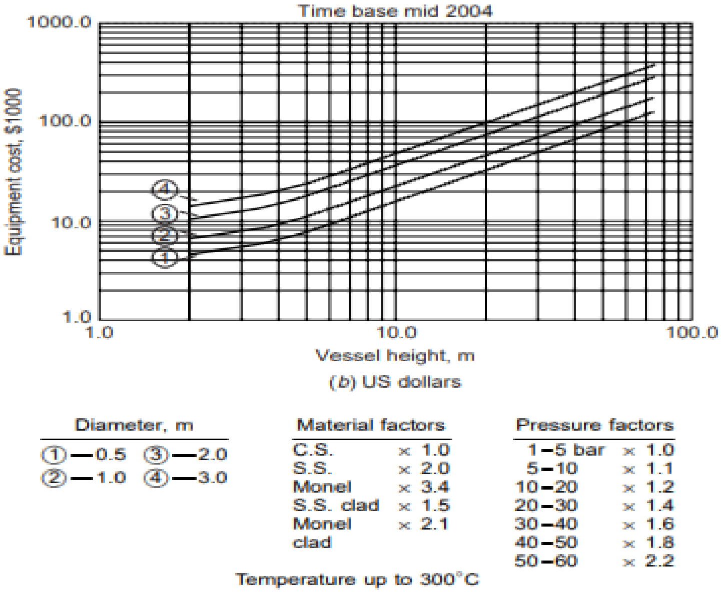


Figure 33: Vessel Cost for Vertical Pressure Vessels

1. For a vessel height of 24m the vessel cost was calculated using figure 8.1 above by extrapolating the height of the vessel where it crossed the curve 3 for 2m diameter. The bare cost of the vessel came out to be \$80,000.
2. The material chosen for this vessel was carbon steel with a material factor of 1.0.
3. Since the operating pressure is 12 bar, the pressure factor of 1.2 would be applicable.
4. Total cost of the vessel can be calculated using:

$$\begin{aligned}\text{Total Cost of Vessel} \\ &= \text{Bare Vessel Cost } \times \text{Material Factor } \times \text{Pressure Factor} \\ &= 80000 \times 1.0 \times 1.2 = \$\mathbf{96000}\end{aligned}$$

8.2.2 Total Cost of Plates

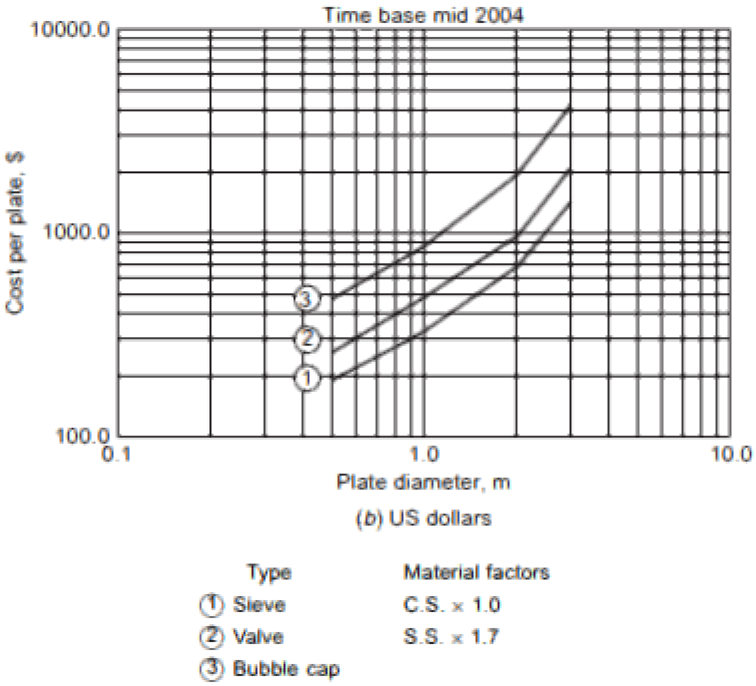


Figure 34: Cost of a distillation column plate

1. For a valved type of plate, the cost per plate was calculated by using a plate diameter of 2m and extrapolating it where it crosses curve 2 on figure 8.2 to the cost per plate axis. The cost per plate came out to be $950.
2. For a carbon steel plate, the material factor as shown in figure 8.2 is 1.0.
3. Total cost of all plates was calculated using the formula:

$$\text{Total cost of all plates} = (\text{cost of one plate } \times \text{material factor}) \times \text{total no. of plates}$$

$$= (950 \times 1.0) \times 44 = \$\mathbf{44950}$$

8.2.3 Total Cost of Stabilizer

The total cost of stabilizer is the sum of the cost of the vessel and cost of all plates:

$$= 96000 + 44950 = \$\mathbf{124,950}$$

8.2.4 Indexing from 2004 to 2023

Index in 2004 = 188

Index in 2023 = 303

Price in 2023 = $\frac{\text{Index in 2023}}{\text{Index in 2004}} \times \text{Price in 2004} = \frac{303}{188} \times 124950 = \mathbf{\$201,382}$

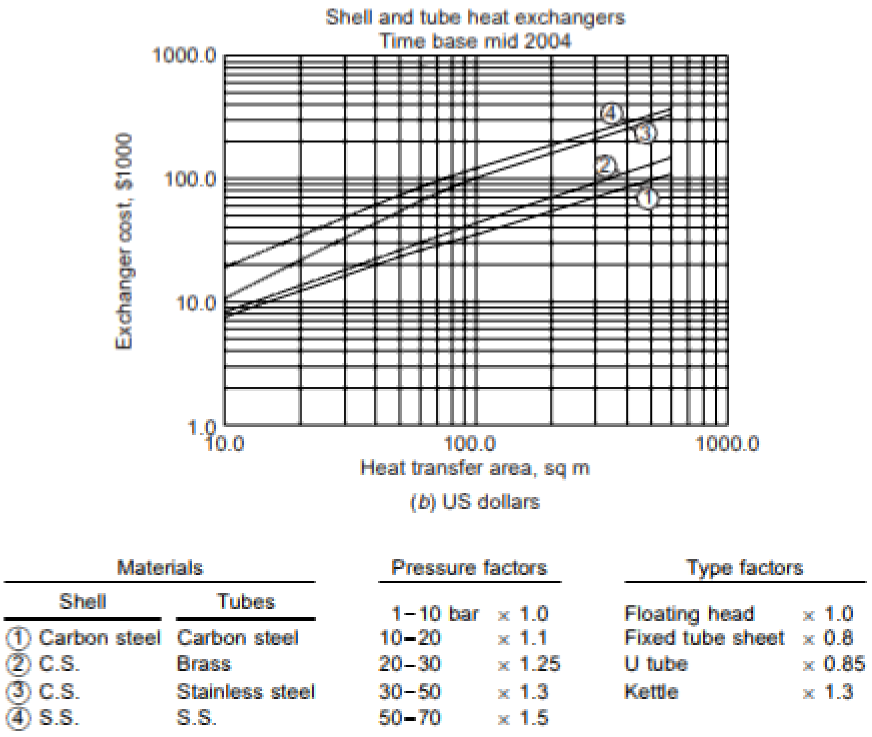
8.3 Costing of Shell and Tube Heat Exchanger

8.3.1 Total Cost of Heat Exchanger

Table 32: Design Specification of the shell and tube heat exchanger

Parameters	Value
Type	Fixed Tube Sheet
Passes	2
Area/m2	190.6
Material	Stainless Steel
Operating Pressure/bar	15.5

The specifications above will be used to calculate the cost of the shell and tube heat exchanger.



Materials		
	Shell	Tubes
①	Carbon steel	Carbon steel
②	C.S.	Brass
③	C.S.	Stainless steel
④	S.S.	S.S.

Pressure factors	
1–10 bar	× 1.0
10–20	× 1.1
20–30	× 1.25
30–50	× 1.3
50–70	× 1.5

Type factors	
Floating head	× 1.0
Fixed tube sheet	× 0.8
U tube	× 0.85
Kettle	× 1.3

Figure 35: Cost of a shell and tube heat exchanger

1. For a heat transfer area of 190.6 m^2 and stainless-steel shell and tube, the area as extrapolated to the curve 4 from where it was stretched to the cost axis which gave a bare vessel cost of $108,000.
2. The pressure factor of the heat exchanger is 1.1 owing to an operating pressure of 1.1.
3. The type of heat exchanger is a fixed tube type so the type factor of the heat exchanger is 0.80.
4. The total cost of the heat exchanger will be calculated by:

$$\text{Total Cost of Heat Exchanger} = \text{bare vessel cost} \times \text{pressure factor} \times \text{type factor}$$

$$= 108000 \times 1.1 \times 0.8 = \$\mathbf{95,040}$$

8.3.2 Indexing

Index in 2004 = 188

Index in 2023 = 303

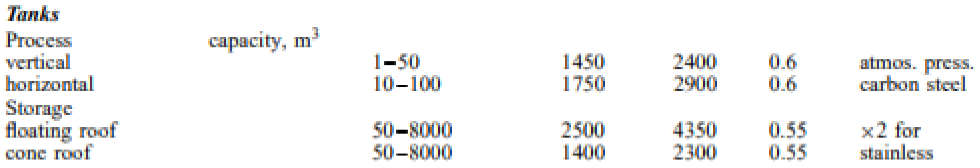
$$\text{Price in 2023} = \frac{\text{Index in 2023}}{\text{Index in 2004}} \times \text{Price in 2004} = \frac{303}{188} \times 95040 = \mathbf{\$153{,}176}$$

8.4 Costing of Accumulator

8.4.1 Total Cost of Accumulator

Table 33: Design Specifications of Accumulator

Parameter	Value
Capacity/m^3	2
Material	Carbon Steel



Tanks						
Process	capacity, m^3					
vertical		1–50	1450	2400	0.6	atmos. press.
horizontal		10–100	1750	2900	0.6	carbon steel
Storage						
floating roof		50–8000	2500	4350	0.55	×2 for
cone roof		50–8000	1400	2300	0.55	stainless

Figure 36: Costing of Vessels

1. Using a capacity of $2m^3$ of the vessel, the bare vessel cost is $2,400.
2. The index, n, for the carbon steel material at atmospheric pressure is 0.66.
3. Following equation would be used to calculate the cost of this equipment:

$$C(\$) = CS^n$$

$$C(\$) = 2400 \times 2^{0.6}$$

$$\mathbf{C = \$3,638}$$

8.4.2 Indexing

Index in 2004 = 188

Index in 2023 = 303

$$\text{Price in 2023} = \frac{\text{Index in 2023}}{\text{Index in 2004}} \times \text{Price in 2004} = \frac{303}{188} \times 3638 = \mathbf{\$5,863}$$

8.5 Purchased Cost of Equipment

8.5.1 Cost of Equipment, Time base Mid-2004

The cost of the other equipment in the plant was obtained from the economic analysis section of the simulation.

Table 34: Cost of Equipment from Simulation

Equipment	Cost/$
Heat Exchangers Train 1 and 2	1,778,300
Crude Charge Pump	97,100

Table 34 (cont'd)

Desalter	59,200
Stabilizer Pump	23,800
Pre flash tower with Reboiler and condenser	608,600

8.5.2 Indexing

Index in 2017 = 244

Index in 2023 = 303

Using Price in 2023 = $\frac{\text{Index in 2023}}{\text{Index in 2017}} \times$ Price in 2017, the cost of these equipment in 2023 is calculated below.

Table 35: Cost of Equipment from Simulation indexed to 2023

Equipment	**Cost/$**
Heat Exchangers Train 1 and 2	2,208,300
Crude Charge Pump	120,560
Desalter	73,515
Stabilizer Pump	29,555
Pre flash tower with Reboiler and condenser	755,760

8.5.3 Total purchased cost of all equipment

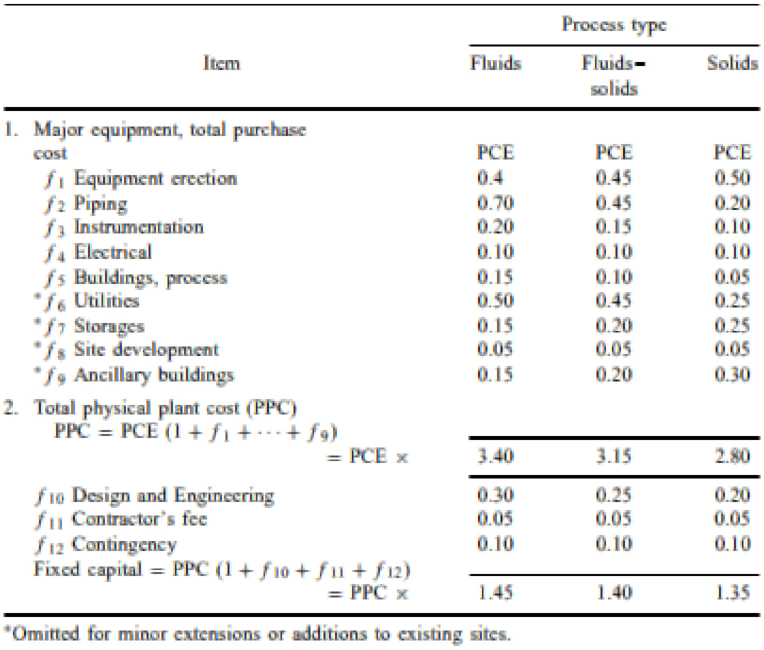
The total purchased cost of all equipment is calculated by adding all the costs of equipment as shown on the next page.

Table 36: Purchased Cost of Equipment

Equipment	Cost/$
Heat Exchangers Train 1 and 2	2,208,300
Crude Charge Pump	120,560
Desalter	73,515
Stabilizer Pump	29,555
Pre flash tower with Reboiler and condenser	755,760
Stabilizer	201,382
Heat Exchanger	153,176
Accumulator	5,863
PCE	3,548,111

8.6 Total Investment Required for Project

Table 37: Costing f factors



Item	Process type		
	Fluids	Fluids-solids	Solids
1. Major equipment, total purchase cost	PCE	PCE	PCE
f_1 Equipment erection	0.4	0.45	0.50
f_2 Piping	0.70	0.45	0.20
f_3 Instrumentation	0.20	0.15	0.10
f_4 Electrical	0.10	0.10	0.10
f_5 Buildings, process	0.15	0.10	0.05
*f_6 Utilities	0.50	0.45	0.25
*f_7 Storages	0.15	0.20	0.25
*f_8 Site development	0.05	0.05	0.05
*f_9 Ancillary buildings	0.15	0.20	0.30
2. Total physical plant cost (PPC) $PPC = PCE (1 + f_1 + \cdots + f_9)$ $= PCE \times$	3.40	3.15	2.80
f_{10} Design and Engineering	0.30	0.25	0.20
f_{11} Contractor's fee	0.05	0.05	0.05
f_{12} Contingency	0.10	0.10	0.10
Fixed capital $= PPC (1 + f_{10} + f_{11} + f_{12})$ $= PPC \times$	1.45	1.40	1.35

*Omitted for minor extensions or additions to existing sites.

8.6.1 Physical Plant Cost

The total physical plant cost is calculated using the formula:

$$PPC = PCE(1 + f_1 + \cdots + f_9)$$

Physical Plant cost is calculated using the factors below.

Our plant is a fluids processing plant.

Table 38: Factors for our project

Factor	Value
Equipment Erection	0.40
Piping	0.70
Instrumentation	0.20
Electrical	0.10
Buildings, Process	None Required
Utilities	Not Applicable
Storage	Provided in PCE
Site Development	Not Applicable
Ancillary Buildings	None Required
Physical Plant Cost/$	8,515,466

8.6.2 Fixed Capital Cost

Table 39: Factors for fixed capital

Factor	Value
Design and Engineering	0.3
Contractor’s fee	Nor required
Contingency	0.1
Fixed Capital/$	11,921,652

8.6.3 Total Investment required for project

Working capital required for initial charge = 5% of fixed capital

Working Capital = $596,083

Total Investment Required for project = 11,921,652 + 596,083

= $12,517,735

8.7 Annual Operating Cost

Table 40: Summary of Production Costs

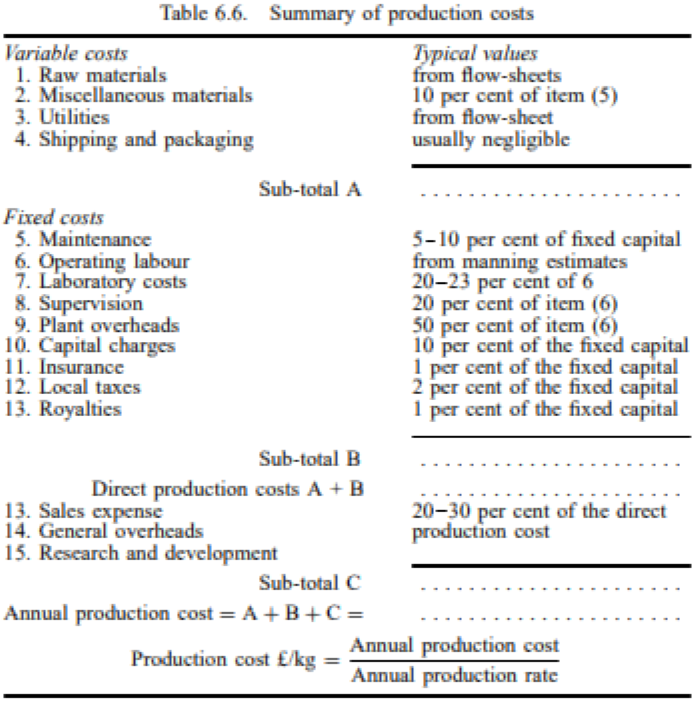


Table 6.6. Summary of production costs

Variable costs	*Typical values*
1. Raw materials	from flow-sheets
2. Miscellaneous materials	10 per cent of item (5)
3. Utilities	from flow-sheet
4. Shipping and packaging	usually negligible
Sub-total A	....
Fixed costs	
5. Maintenance	5–10 per cent of fixed capital
6. Operating labour	from manning estimates
7. Laboratory costs	20–23 per cent of 6
8. Supervision	20 per cent of item (6)
9. Plant overheads	50 per cent of item (6)
10. Capital charges	10 per cent of the fixed capital
11. Insurance	1 per cent of the fixed capital
12. Local taxes	2 per cent of the fixed capital
13. Royalties	1 per cent of the fixed capital
Sub-total B	....
Direct production costs A + B	....
13. Sales expense	20–30 per cent of the direct production cost
14. General overheads	
15. Research and development	
Sub-total C	....
Annual production cost = A + B + C =	....

$$\text{Production cost £/kg} = \frac{\text{Annual production cost}}{\text{Annual production rate}}$$

8.7.1 Plant Attainment

To calculate the annual operating cost, the plant attainment i.e., the number of days for which the plant operates in a year is calculated first.

Plant Attainment = 95%

For 1 year, no. of days = 347 days/ year

No. of hours = 8328 hrs./year

8.7.3 Variable operating costs

The variable costs associated with this project are outlined below in reference to table 8.10 above. The time base is mid-2004, which is then indexed to 2023.

Table 41: Variable Costs

Description	Cost/$
Maintenance at 5% of Fixed Capital	595,988
Miscellaneous Materials @ 10% of Fixed Capital	59,599
Raw Materials	Not Applicable
Utilities	
Electricity ($/hr)	30
Total Cost of Electricity ($/yr)	249,840
Cooling Water cost ($/hr)	66
Total cost of cooling water ($/yr)	549,648
Steam cost ($/hr)	74
Total cost of steam	616,272
Shipping and packaging	N/A
Total variable costs	1,475,359

8.7.3 Fixed operating cost

The fixed costs associated with this project are outlined below in reference to table 8.10 above. The time base is mid-2004, which is then indexed to 2023.

Table 42: Variable Costs

Description	Cost/$
Maintenance at 5% of Fixed Capital	595,988
Operating Labor	30,000
Plant Overheads @ 50% of operating labor	15,000
Lab @ 30% of operating labor	9,000
Capital charges @ 6% of Fixed Capital	715,186
Insurance @ 1% of Fixed Capital	119,198
Total Fixed operating cost	1,484,372

8.7.4 Total Annual operating cost

The total annual operating cost is the sum of variable operating cost and fixed operating cost and is given below.

Total annual operating cost = Fixed cost + Variable cost

= 1484372 + 1475359

= **$2,959,731**

8.8 Revenue, Profit, and Payback period

The revenue is calculated by determining an acceptable market selling price of the product and multiplying it with the amount of product sold in the selected time-basis.

The profit is then calculated by subtracting the revenue from the annual operating cost.

The payback period, lastly, is calculated by dividing the total investment required for the project with the profit per year.

All of these calculations are shown in the table.

Table 43: Revenue from selling LPG, profit and Payback period

Description	Value
LPG selling price ($/liter)	0.235
Volumetric flowrate of LPG (liters/hr)	4500
Total production of LPG (liters/year)	37,476,000
Annual Revenue from LPG ($/year)	8,806,860
Net Profit ($/year)	5,847,129
Payback period (years)	2.14

The Payback period is calculated to be 2.14 years. In literature, for a project to be feasible, the payback period should be between 2 to 5 years. Comparing our value with the literature, we can safely say that in terms of finances, our project is highly feasible.

CHAPTER 9

CONTROL LOOPS

9.1 Temperature Control Loop

9.1.1 Heat Exchanger

Process Variable

The process variable is the temperature of the feed stream entering the stabilizer unit. The objective is to heat the naphtha to a specific temperature set point.

Controlled Variable

The controlled variable is the flow rate of steam used for heating the naphtha.

Disturbances

Several disturbances can impact the outlet temperature of the naphtha in the stabilizer:

1. Changes in the flow rate of the naphtha feed stream.
2. Changes in the inlet temperature of the naphtha feed stream.
3. Variations in the flow rate of the heating steam.

Control Objective:

The main objective of the control system is to maintain the outlet temperature of the naphtha, T1Out, at the desired set point, regardless of disturbances. This is achieved by manipulating the flow rate of the heating steam.

Feedback Control Loop:

The control scheme utilized in this simulation is a feedback control loop. The process variable, which is the measured temperature Tout, is fed into a PID (Proportional-Integral-Derivative) based feedback temperature controller (TC). The TC compares the process variable with the desired temperature set point and calculates the necessary control action. It then adjusts the flow rate of the heating steam by either opening or closing the steam control valve accordingly to maintain the desired temperature.

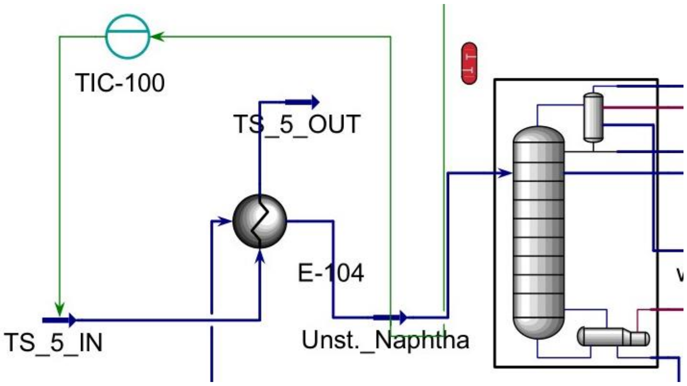
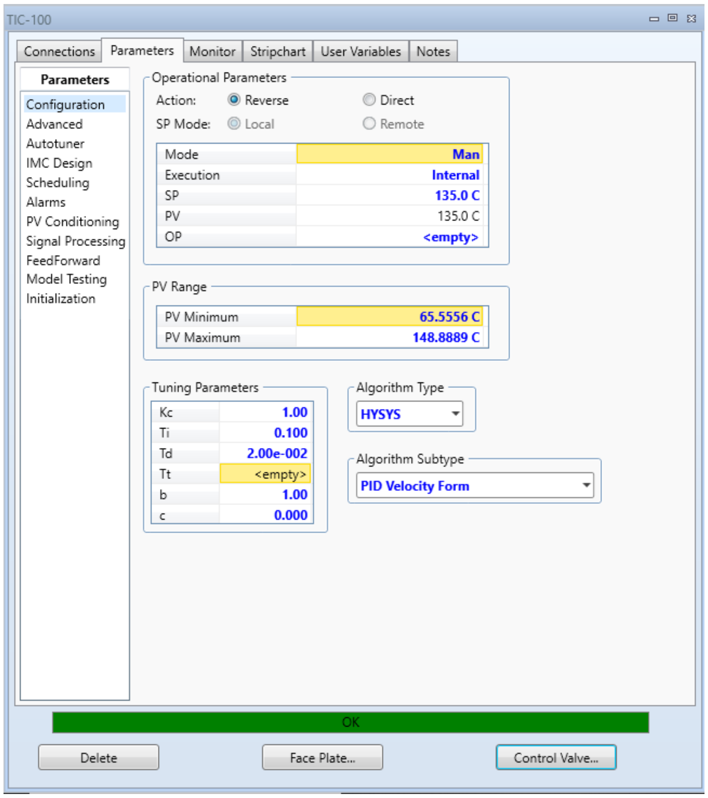


Figure 37: Temperature Control Loop on Stabilizer Feed



Operational Parameters

Action: ● Reverse ○ Direct

SP Mode: ○ Local ○ Remote

Mode	Man
Execution	Internal
SP	135.0 C
PV	135.0 C
OP	<empty>

PV Range

PV Minimum	65.5556 C
PV Maximum	148.8889 C

Tuning Parameters

Kc	1.00
Ti	0.100
Td	2.00e-002
Tt	<empty>
b	1.00
c	0.000

Algorithm Type: HYSYS

Algorithm Subtype: PID Velocity Form

Figure 38: Parameters of the Temperature Controller

By implementing the feedback control loop, the naphtha stabilizer system can effectively regulate the flow rate of the heating steam and ensure that the naphtha reaches and maintains the desired temperature, even in the presence of disturbances.

9.2 Pressure Control Loop

Process Variable

The process variable is the pressure of the feed stream entering the stabilizer unit. The objective is to maintain the pressure at an optimum condition.

Controlled Variable

The controlled variable is the speed of the impeller in the pump used to control the feed stream. The controller adjusts the impeller speed to regulate the pressure of the naphtha stabilizer.

Disturbances

Several disturbances can affect the pressure of the naphtha feed stream in the stabilizer:

1. Changes in the flow rate of the naphtha feed stream.
2. Variations in the properties of the incoming naphtha, such as density or viscosity.
3. Changes in the system's operating conditions, such as fluctuations in the downstream pressure.

Control Objective

The main objective of the control system is to maintain the pressure of the naphtha feed stream at the optimum set point, despite disturbances. The control loop ensures that the pressure remains stable by manipulating the speed of the impeller in the pump.

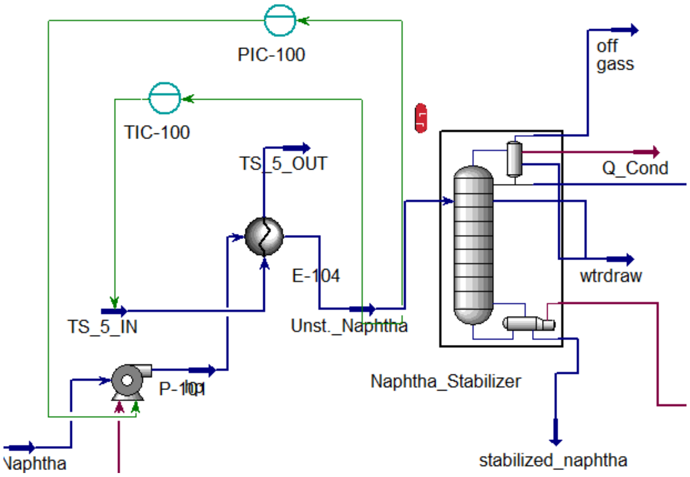


Figure 39: Pressure Control Loop on Stabilizer Feed

By utilizing the feedback control loop, the pressure controller effectively regulates the impeller speed, ensuring that the pressure of the naphtha feed stream remains within the desired range, even in the presence of disturbances.

9.3 Concentration Control Loop

Process Variable

The process variable in the desalter section of the plant is the concentration of salt in the crude oil leaving the desalter unit. It serves as a key parameter to monitor the effectiveness of salt removal from crude oil.

Controlled Variable

The controlled variable is the flow rate of wash water to the desalter. By adjusting the flow rate, the concentration controller regulates the amount of water used for washing the crude oil and controlling the salt concentration.

Concentration Sensor

A conductivity sensor is installed downstream of the desalter to measure the salt concentration in the crude oil accurately. This type of sensor utilizes the electrical conductivity of the fluid to determine the salt concentration. The conductivity sensor provides real-time feedback to the concentration controller, allowing for precise control of the wash water flow rate.

Control Objective

The main objective of the concentration control system is to maintain the salt concentration in the crude oil leaving the desalter within a desired range. This ensures efficient removal of salts from the crude oil and enhances the quality of the refined products.

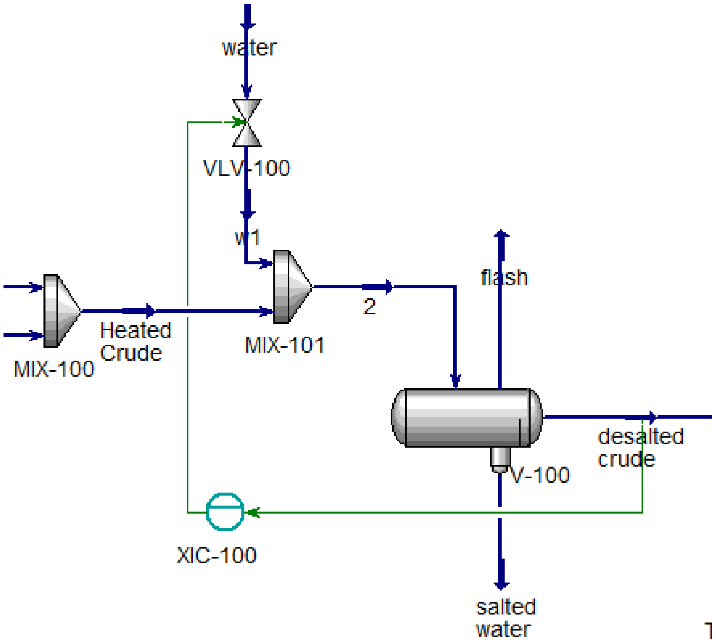
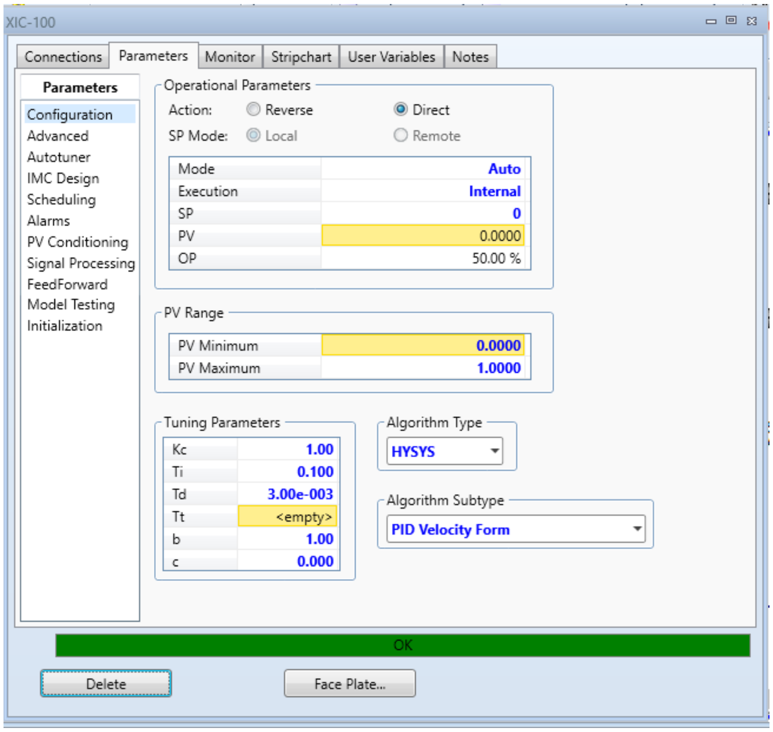


Figure 40: Salt Concentration Control Loop on Desalter

Control Strategy

The concentration controller continuously monitors the salt concentration through the conductivity sensor. If the measured salt concentration exceeds the set point, indicating a higher salt content, the concentration controller increases the flow rate of wash water. This intensifies the washing process and enhances the removal of salts from the crude oil.



Operational Parameters

Action: ○ Reverse ● Direct

SP Mode: ○ Local ○ Remote

Mode	Auto
Execution	Internal
SP	0
PV	0.0000
OP	50.00 %

PV Range

PV Minimum	0.0000
PV Maximum	1.0000

Tuning Parameters

Kc	1.00
Ti	0.100
Td	3.00e-003
Tt	<empty>
b	1.00
c	0.000

Algorithm Type: HYSYS

Algorithm Subtype: PID Velocity Form

Figure 41: Parameters of the Concentration Controller

Conversely, if the salt concentration is below the set point, the concentration controller decreases the flow rate of wash water to avoid excessive water usage. This control strategy ensures that the salt concentration remains within the desired range.

CHAPTER 10

HAZOP ANALYSIS

A hazard and operability study is used for systematic, critical, examination of operability of a process. It can be applied to a process design or operating plant to find out the potential hazards that may arise from deviations that occur instead of intended design conditions. This safety analysis finds its application in identifying potential risks and hazards surrounding an industrial process. HAZOP analysis deals with minimizing accidents at plants. Following are the goals achieved by HAZOP hazard analysis:

1. Identification of hazard.
2. Extent of the damage.
3. Guidelines to solve the problem.

10.1 HAZOP Guide Words

HAZOP utilizes a set of Guide words that identify probable deviations that can occur during a chemical process. These guide words prove their worth by allowing a systematic and consistent procedure through brainstorming. Some of these guide words which are as follows:

Table 44: Guide words and their meaning in HAZOP analysis:

Sr. No	Guide Word	Meaning
1	No	Complete contradiction of original design
2	More	Increase quantitatively
3	Less	Decrease quantitatively
4	Other Than	Entire design

Table 45: Hazop Study at Distillation Column

Guide Words	Parameters	Possible Causes	Possible Consequences	Actions Required
More	Temperature	Excessive heat input, Incorrect temperature control settings.	Thermal degradation Equipment damage Product quality issues	Optimize temperature control, implement safety interlocks, and relief devices, monitor temperature limits.
Less		Insufficient heat input, Inadequate temperature control.	Inadequate vaporization Poor separation efficiency	calibrate temperature control systems, maintain heat transfer surfaces.
More	Pressure	Excessive pressure	Overpressure Equipment damage	Implement redundant pressure control systems
Less		Insufficient pressure	Inadequate vapor-liquid contact, Poor separation efficiency	Ensure sufficient pressure, calibrate pressure control systems.
No	Liquid Flowrate	Blocked or clogged liquid lines, Pump failure.	Inadequate liquid-vapor contact, Reduced efficiency.	Regularly inspect and maintain liquid lines, install backup pumps
More		Excessive liquid flow rate, Incorrect control settings	Increased pressure drops, weeping, channeling, flooding	Optimize liquid flow control, adjust flow rates,
Other Than		Inadequate liquid distribution Improper tray/packing design	Uneven liquid distribution Reduced separation efficiency	Optimize tray/packing design, ensure proper liquid distribution, maintain tray/packing.

Table 45 (cont'd)

No	Level	Instrument failure, Blockage in level sensing lines	Reduced separation efficiency, Potential equipment damage	Implement backup level control measures, monitor level indicators, conduct regular maintenance.
More		Inlet valve malfunction	Risk of liquid carryover, potential flooding, reduced vapor-liquid contact efficiency	Optimize level control, ensure safety measures are in place, monitor level limits.
Less		Outlet valve malfunction	Reduced separation efficiency, potential product loss	Calibrate level control systems, maintain equipment integrity.

Table 46: Hazop Study at Condenser

Guide Words	Parameters	Possible Causes	Possible Consequences	Actions Required
More	Corrosion	Corrosive feedstock,	Material degradation, loss of structural integrity,	Analyze feedstock composition, consider corrosion-resistant materials, optimize corrosion protection measures

Table 46 (cont'd)

More	Temperature	Excessive cooling	Corrosion, reduced condensation efficiency,	Optimize temperature control, implement safety interlocks and relief devices,
Less		Insufficient cooling	Inadequate condensation, reduced product recovery	Ensure sufficient cooling, calibrate temperature control systems, maintain heat transfer surfaces.
More	Pressure	Excessive pressure	Increased pressure drops, equipment damage, product quality issues	Optimize pressure control, implement safety interlocks and relief devices, monitor pressure limits.
Less		Insufficient pressure	Reduced condensation efficiency, inadequate separation	Ensure sufficient pressure, calibrate pressure control systems, maintain condenser internals.

Table 47: Hazop Study at Reboiler

Guide Words	Parameters	Possible Causes	Possible Consequences	Actions Required
More	Heating Medium	Excessive heating medium flow Failure in control mechanisms	Risk of overpressure, potential equipment damage or failure, increased energy consumption	Optimize flow control, ensure safety relief devices are in place, monitor pressure limits.
Less		Insufficient heating medium flow Blockage in heating medium lines	Reduced vaporization, inefficient separation	Clear blockages, maintain heating medium lines.

CONCLUSION

In this project, a comprehensive cost benefit analysis of different techniques for stabilization was done in literature review after which stabilization by distillation was found to be the most economic and feasible technique for our project. The selection and design of equipment was done based on the characteristics of the raw materials we were processing and the desired products through which we also obtained our process flow diagram. Material and energy balance calculations were done to see the feasibility of this process in terms of raw materials required, products made, and utilities required. Costing and evaluation of the process was then done which gave a very reasonable payback period. Simulation of the whole process on aspen hysys was also conducted to see different scenarios and conditions to select final conditions. Lastly, HAZOP analysis was conducted to determine the safest possible way in which our desired operation could be conducted. Conclusively, stabilization of naphtha by adding a new distillation column in the light crude distillation unit of Attock Refinery Limited and producing LPG is a highly feasible solution in terms of economics, safety, practical implementation and technology.

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