DESIGN AND SIMULATION OF PRE-REFORMER UNIT IN REFINERY

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2024

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A THESIS

Submitted to

National University of Sciences and Technology

in partial fulfilment of the requirements for the degree of

School of Chemical and Materials Engineering (SCME)

National University of Sciences and Technology (NUST)

June 2024

CERTIFICATE

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DEDICATION

This work is dedicated to individuals who believed in the potential of this group, especially, our beloved families, friends, and teachers who played a significant role in bringing us this far.

ACKNOWLEDGEMENTS

Thanks to Almighty Allah for His countless blessings, He gave us the courage to finish this project successfully.

A very special recognition to our supervisor, **Dr. Nouman Ahmad** for supervising this project and for his support throughout the year. Because of his experience and knowledge, we are able to complete this project.

We are able to achieve this milestone due to love and prayers of our beloved Parents.

ABSTRACT

The presence of contaminants in naphtha, including sulfur, oxygen, nitrogen, and organometallic compounds, poses a substantial obstacle to the refining process of crude oil into useful products. The effectiveness and durability of the reformer's catalyst, which is essential and expensive, are threatened by these contaminants. This project's goal was to model and create a pre-reformer unit that would produce "sweet" naphtha devoid of these harmful contaminants.

The study began with a thorough analysis of the literature to understand the issues and current approaches in naphtha purification. After providing a brief overview of the project, we meticulously calculated the energy and mass balances of every piece of equipment. A comprehensive evaluation of the energy and material balance was then carried out to guarantee the process's feasibility.

A thorough simulation was run using the potent simulation tool HYSYS, which made it possible to assess several operational factors and setups. In order to ensure the safety and dependability of the suggested design, a Hazard and Operability (HAZOP) study was also carried out on the primary equipment to detect and minimize any risks.

In addition, a thorough cost analysis was carried out for every part of the pre-reformer unit, offering information on investment considerations and economic viability. The project's design of a pre-reformer unit using Co-Mo catalyst, which showed encouraging results and supported its possible integration into refinery operations, was its climate.

The study's conclusions not only provide a sustainable way to produce sweet naphtha, but they also improve refinery operations by making the best use of available resources and reducing their negative effects on the environment.

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CHAPTER 1

INTRODUCTION

1.1 Introduction to Refinery Operations

Refineries, often known as oil refineries or petroleum refineries, are commercial establishments that process crude oil to produce petrochemicals, petrol, diesel, jet fuel, heating oil and other valuable products. It is essential to the oil industry's downstream sector because it transforms crude oil into refined products that satisfy consumer demand. Crude oil is broken down into its component parts by refineries using a number of intricate procedures, which are subsequently reassembled into more valuable products.

1.2 Refinery Processes

Primary and secondary processes are the two types of refinery processes. Through distillation, crude oil is separated into several fractions in primary processes, and these fractions are then further refined to increase their quality and produce valuable products in secondary processes.

1.2.1 Distillation:

Distillation towers are used to heat crude oil and divide it into several parts according to their respective boiling points. Heavy fractions like diesel and leftover fuel oil fall to the bottom, while lighter fractions like gasses and naphtha rise to the top.

1.2.2 Conversion:

Cracking, reforming, and hydro-processing are examples of secondary processes that are used to change heavier fractions into lighter, higher-value products. High-octane gasoline is created by reforming, which reorganizes the molecular structure of bigger hydrocarbon molecules into smaller ones.

1.2.3 Treatment:

To meet environmental criteria and enhance product quality, refineries employ treatment techniques including hydrotreating, desulfurization, and hydrocracking to eliminate pollutants like sulfur, nitrogen, and metals.

1.3 Hydrotreating Section

An essential refinery procedure called hydrotreating is used to improve the quality of petroleum products by eliminating contaminants like nitrogen and sulfur. In this process, feedstock—such as gasoline or diesel—is passed over a catalyst at high pressures and temperatures while hydrogen gas is present. Hydrogenation, desulfurization, denitrification, and desulfurization are the usual reactions that take place during hydrotreating.

1.3.1 Hydrogenation:

Hydrogen molecules saturate the unsaturated hydrocarbons in the feedstock, transforming them into more stable, saturated compounds. As a result, the final products operate better and are more stable.

1.3.2 Desulfurization:

In order to eliminate sulfur compounds—which can contaminate the air and harm engine parts—they must first react with hydrogen to create hydrogen sulfide, which is subsequently processed to create elemental sulfur or sulfuric acid.

1.3.3 Denitrification:

Ammonia or nitrogen gas, which are easily separated and removed, are produced from nitrogen compounds, which have the ability to deactivate catalysts and lower the efficiency of downstream processes.

1.4 Obtaining Naphtha

Crude oil is mostly distilled to produce naphtha, a crucial intermediate product in the refining process. Naphtha is separated from other fractions during the distillation process of crude oil, and it usually leaves the distillation tower in between the kerosene and gasoline fractions. It is a light hydrocarbon mixture made up of different aromatic and aliphatic molecules with five to twelve carbons.

A flexible feedstock, naphtha is utilized in a number of downstream operations, such as the manufacture of petrochemicals, reforming, and gasoline blending. It is used as an input into the manufacturing of many different items, including as polymers, solvents, and parts for gasoline.

1.5 Naphtha

Naphtha is a liquid combination of hydrocarbons that is extremely volatile and combustible. It is an essential feedstock for the petrochemical sector, especially for operations like the generation of gasoline, olefins through steam cracking, and aromatics through catalytic reforming. Depending on where it comes from—a crude oil distillation unit (straight-run naphtha) or cracking procedures (cracked naphtha)—the composition of naphtha can differ greatly.

1.5.1 Importance of Naphtha

Naphtha's versatility and usefulness make it a valuable ingredient in the petrochemical and refining industries. Naphtha is used as a major feedstock and goes through a number of processing stages to become high-octane gasoline, which is a necessary fuel for internal combustion engines. Furthermore, it is a crucial raw material in the synthesis of propylene and ethylene, which serve as the fundamental building blocks for a variety of chemical products, such as synthetic rubber, plastics, and industrial chemicals. Naphtha's quality and purity have a direct bearing on the productivity and efficiency of these downstream operations, hence purification and treatment are essential.

1.6 Importance of Pre-reforming/Hydrotreating

In order to purge hydrocarbon streams of contaminants like sulfur, nitrogen, oxygen, and organometallic compounds, hydrotreating is a crucial catalytic refining procedure. Hydrotreating is crucial when it comes to naphtha since impurities can poison catalysts used in later processing stages, notably in catalytic reforming units. For example, sulfur and nitrogen compounds can drastically shorten the life and efficacy of reforming catalysts, resulting in increased expenses and decreased efficiency.

A hydrogen-rich gas stream and a selective catalyst—typically an alumina base coated with cobalt, nickel, and molybdenum compounds—are used in the hydrotreating process. This mixture makes it easier for organic contaminants to break down and transform into more controllable substances like ammonia (NH_3) and hydrogen sulfide (H_2S), which may subsequently be eliminated from the process stream.

1.7 Reactions in Naphtha Hydrotreating

The hydrotreating process involves several key reactions aimed at removing specific impurities from naphtha:

1.7.1 Sulfur Removal

Hydrogen sulfide (H_2S) is produced from organic sulfur compounds. For example, H_2S is produced by hydrogenizing mercaptans and thiophenes, and it is subsequently extracted from the hydrotreated naphtha.

– ***Main Reactions for Sulfur Removal in Naphtha Hydrotreating:***

Sulfur removal in naphtha hydrotreating is the process of hydrogenating sulfur-containing substances, like mercaptans and sulfides, to create hydrogen sulfide (H_2S). The primary reactions for these chemicals' conversion are listed below:

1) **Conversion of Mercaptans to Hydrogen Sulfide**

Mercaptans, which are organic molecules containing sulfur and have the general formula R-SH, where R is an alkyl group, are also referred to as thiols. Mercaptans typically react as follows to produce hydrogen sulfide:

$$\boldsymbol{R - SH + H_2 \rightarrow R - H + H_2S}$$

For example, for methyl mercaptan (CH_3SH), the reaction would be:

$$\boldsymbol{CH_3SH + H_2 \rightarrow CH_4 + H_2S}$$

2) **Conversion of Sulfides to Hydrogen Sulfide**

The typical formula for sulfides is R-S-R', where R and R' might be the same or separate alkyl groups. Sulfides typically react as follows to become hydrogen sulfide:

$$\boldsymbol{R - S - R' + 2H_2 \rightarrow R - H + R' - H + H_2S}$$

For example, for dimethyl sulfide (CH_3-S-CH_3), the reaction would be:

$$\boldsymbol{CH_3S - CH_3 + 2H_2 \rightarrow 2CH_4 + H_2S}$$

3) Catalyst and Conditions

Usually, a hydrotreating catalyst—cobalt-molybdenum (Co-Mo) or nickel-molybdenum (Ni-Mo) supported on alumina—is present when these reactions take place. High pressures (20–80 bar) and temperatures (300–400°C) are typical process parameters that help efficiently convert sulfur compounds to H_2S.

4) Summary

Hydrogenation of mercaptans and sulfides to produce hydrocarbons and hydrogen sulfide (H_2S) is the main step in the hydrotreating method used to remove sulfur from naphtha. In order to ensure that the naphtha feedstock is pure and safe for use in subsequent refining procedures like catalytic reforming without running the danger of catalyst poisoning, these reactions are essential.

1.7.2 Nitrogen Removal

- Ammonia is produced via the conversion of organic nitrogen molecules (NH_3). Because of the slower reaction rates, denitrification is a more difficult process than desulfurization. Common nitrogenous contaminants like pyrrole and pyridine are hydrogenated to create NH_3, which is then eliminated in the reactor effluent wash water.

Main Reactions for Nitrogen Removal in Naphtha Hydrotreating:

Nitrogen removal in naphtha hydrotreating is the process of hydrogenating organic molecules containing nitrogen to create ammonia (NH_3). The primary reactions for these chemicals' conversion are listed below:

1) Conversion of Pyrrole to Ammonia

Pyrrole (C_4H_4NH) is a nitrogen-containing aromatic compound. The typical reaction for the conversion of pyrrole to ammonia is:

$$C_4H_4NH + 4H_2 \rightarrow C_4H_{10} + NH_3$$

In this reaction:

Pyrrole (C_4H_4NH) reacts with hydrogen (H_2).

Ammonia (NH_3) is produced along with an alkane, typically butane (C_4H_{10}).

2) Conversion of Pyridine to Ammonia

Pyridine (C_5H_5N) is another nitrogen-containing aromatic compound. The typical reaction for the conversion of pyridine to ammonia is:

$$C_5H_5N + 5H_2 \rightarrow C_5H_{12} + NH_3$$

In this reaction:

Pyridine (C_5H_5N) reacts with hydrogen (H_2).

Ammonia (NH_3) is produced along with an alkane, typically pentane (C_5H_{12}).

3) Catalyst and Conditions

Like sulfur removal, nitrogen removal reactions take place in the presence of a hydrotreating catalyst, typically alumina-supported cobalt- or nickel-molybdenum (Ni-Mo). High pressures (20–80 bar) and temperatures (300–400°C) are typical process parameters that help effectively convert nitrogen molecules to NH_3.

4) Summary

Nitrogen-containing substances like pyrrole and pyridine are hydrogenated to form hydrocarbons and ammonia (NH_3) as part of the hydrotreating process used to remove nitrogen from naphtha. The purification of the naphtha feedstock is dependent on these reactions in order to ensure that there is no chance of catalyst poisoning during subsequent refining procedures like catalytic reforming.

1.7.3 Organo-metallic Compounds Removal

- Hydrotreating also aims to target trace levels of metallic contaminants. The naphtha feed is cleaned of elements like arsenic, iron, calcium, magnesium, phosphorus, lead, silicon, copper, and sodium. Usually deposited on the catalyst, these metals can be controlled by periodically regenerating the catalyst or changing the catalyst bed. Hydrogenation, which tries to saturate unsaturated hydrocarbons, including organometallic compounds, by adding hydrogen atoms across their double or triple bonds, is a common reaction engaged in this process. By assisting in the deactivation or removal of the organometallic compounds, this reaction serves to improve the quality of refined products and avoid catalyst poisoning.

1) Hydrogenation

The general hydrogenation reaction for an unsaturated organometallic compound can be represented as:

$$R - M + H_2 \rightarrow RH + MH_2$$

Where:

R represents the organic group attached to the metal (MM)

M represents the metal atom.

RH represents the saturated hydrocarbon product.

In this reaction, hydrogen gas (H_2) and unsaturated organometallic compound (R-M) combine over an appropriate catalyst. Usually, this catalyst is a metal, such as nickel, platinum, or palladium, supported on a solid substance like silica or alumina. The hydrogenation reaction is aided by the metal's catalytic effect. Saturated hydrocarbons (RH) and hydrogenated organometallic compounds (MH_2) are produced when the unsaturated bonds in the organometallic compound are broken, and hydrogen atoms are introduced across these bonds during the hydrogenation process. In later stages of the hydrotreating procedure, the hydrogenated organometallic compounds—which are normally less reactive—may be subjected to additional processing or elimination.

2) Summary

Overall, hydrogenation is essential to the hydrotreating process because it efficiently eliminates or deactivates the organometallic compounds found in the feedstock, enhancing the quality of the refined products and guaranteeing the smooth running of the processes that follow.

CHAPTER 2

PROCESS DESCRIPTION

2.1 Hydrotreating Processes for Naphtha

In a refinery, hydrotreating naphtha is an essential process to improve its stability and octane rating while removing contaminants including sulfur, nitrogen, and metals. There are various hydrotreating process types to choose from, each with special benefits and applicability based on the refinery's needs. This is a synopsis:

2.1.1 Fixed Bed Hydrotreating:

In refineries, one of the oldest and most popular techniques is fixed bed hydrotreating. The hydrogenation processes take place in a fixed catalyst bed as the naphtha feedstock flows across it in this process. Usually, metal sulfides supported on alumina or other materials make up the catalyst bed.

2.1.2 Slurry Phase Hydrotreating:

In order to create a slurry during the slurry phase hydrotreating process, the catalyst particles are suspended in the naphtha feedstock. When compared to fixed-bed reactors, this technique enables better mass transfer and improved catalyst utilization. It has a greater resistance to fouling and can process a broader variety of feedstocks.

2.1.3 Ebullated Bed Hydrotreating:

An ebullated or fluidized bed is produced by boiling the mixture with the catalyst suspended in the feedstock in ebullated bed hydrotreating. Excellent mixing and heat transfer are made possible by this technique, which results in high conversion rates and effective impurity elimination.

2.2 Selection of Suitable Process:

2.3 Fixed Bed Hydrotreating

A fixed bed technique would be appropriate for naphtha hydrotreating, as the feedstock usually contains sulfur, nitrogen, and trace metals. The desulfurization and denitrification reactions, which are essential for raising the naphtha's quality, can be effectively selected for using these methods. Slurry phase reactors offer greater flexibility and tolerance to feedstock variability, however fixed bed reactors are recommended for smaller-scale operations with stable feedstock quality.

The following justifies the application of the fixed-bed hydro-treating method:

Continuous Operation: A constant flow of reactants through the catalyst is made possible by fixed beds. This steady-state operation guarantees reliable performance and effective catalyst use.

Uniform Contact: The catalyst particles in a fixed bed stay in place when the feedstock passes over them. Better conversion rates result from this design, which guarantees consistent contact between the reactants and the catalyst.

Easy Catalyst Replacement: It is possible to change the catalyst without having to stop the entire operation when it starts to lose its activity over time. Maintaining factory productivity and uptime requires doing this.

2.4 Choice of Catalyst

Hydrotreating frequently makes use of Ni-Mo (nickel-molybdenum) and Co-Mo (cobalt-molybdenum) catalysts. However, we favoured Co-Mo for the reasons listed below:

- **Preferred for Hydrodesulfurization (HDS):**

  Sulfur molecules are effectively eliminated from hydrocarbon feeds by Co-Mo catalysts. They effectively reduce sulfur dioxide emissions during combustion by converting sulfur-containing molecules into stable hydrocarbons.

- **Selective Hydro decarboxylation (HDC):**

  Moreover, carboxylic acid groups are removed from fatty acids through hydro decarboxylation, which is facilitated by Co-Mo catalysts and raises the product's overall quality.

- **Suitable for Lighter Feeds:**

  Co-Mo catalysts are effective even when dealing with lighter feedstocks.

2.5 Process Flow Diagram

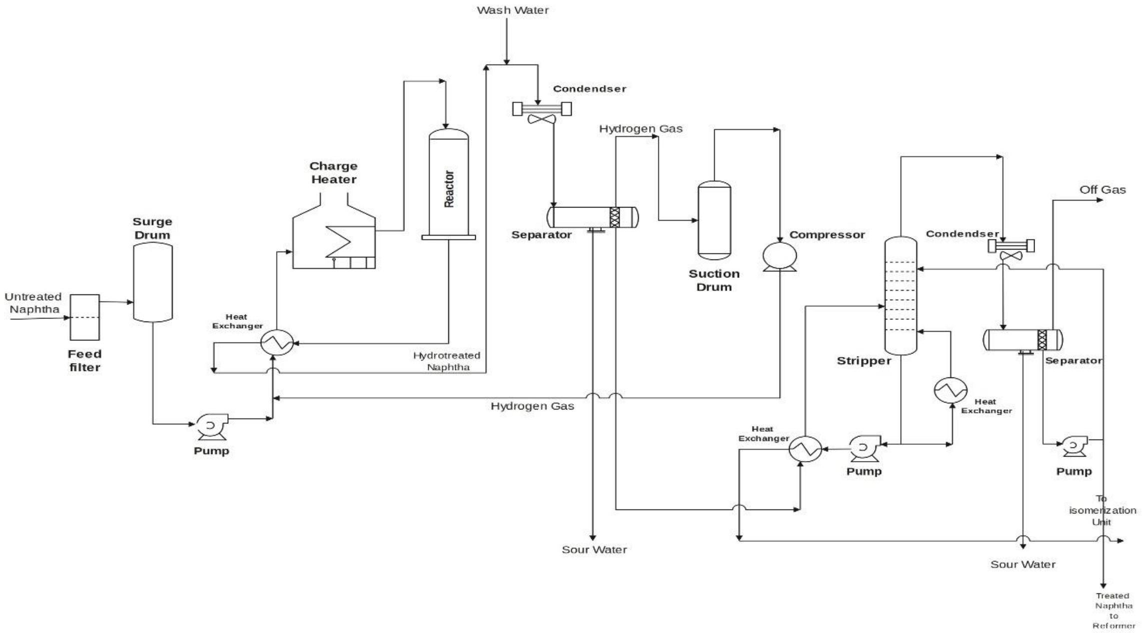


Figure 0-1: PFD

2.5.1 Prefractionation Section

For feeding into a Platforming Unit, it can be beneficial to produce a naphtha fraction with a limited boiling range in some specific situations. This is especially true for aromatic production processes, where the feed endpoint of the Platforming Unit is limited to around 160°C (325°F) in order to increase the concentration of aromatic precursors. Under such circumstances, a prefractionation section can be used to refine the crude unit's entire boiling range naphtha.

The pre-fractionator and the rerun column are typically the first two distillation columns in the prefractionation section. The pre-fractionator heats the feed at first, sometimes with the use of a steam heater in addition to heat exchange with the bottoms of the rerun columns. The hydrotreating reactor uses the overflow product from the rerun column as the heart cut for additional processing. The amount of light naphtha taken overhead in the pre-fractionation and rerun columns is controlled to maintain the heartcut's boiling range. More specifically, the pre-fractionator sets the heartcut's initial boiling point (IBP), while the rerun column determines the heartcut's final boiling point.

By adjusting the net overhead liquid control valve, the overhead temperature controller in the pre-fractionator modifies the amount of light naphtha collected as an overhead liquid product. Raising the overhead temperature affects the heartcut's initial boiling point by increasing the amount of overhead product and raising its endpoint. For instance, the procedure will extract light naphtha if a naphtha fraction with a boiling range of 38-204°C (100-400°F) is added to the prefractionation portion.

The overhead reflux rate is controlled by the overhead receiver level controller of the pre-fractionator. The reflux rate grows in tandem with the receiver level. For example, the net overhead liquid valve closes and raises the overhead receiver level if the pre-fractionator overhead temperature rises above its fixed point. By increasing the reflux rate, this raised level brings the overhead temperature back down to its predetermined level.

Without undergoing further heating, the pre-fractionator column bottoms are pumped straight to the rerun column. The flow rate is set by the lowest level controller of the pre-fractionator. The rerun column, which is likewise controlled by an overhead temperature controller, collects the desired product as overhead. Raising the overhead temperature raises the endpoint and increases the amount of material taken as overhead. For instance, modify the overhead temperature of the rerun column to obtain a heartcut with a boiling range of 82-160°C (180-320°F).

The overhead receiver level controller of the rerun column controls its overhead reflux rate. The reflux rate rises in tandem with the receiver level. Reboilers are installed in both columns to supply the heat required to evaporate naphtha and maintain sufficient reflux. Both the pre-fractionator's overhead and its bottom products are transferred to storage so that they can be combined or processed further in downstream units.

2.5.2 Reactor System

- **The Feed**

The unit can accept naphtha feed from storage, however in order to keep oxygen from dissolving in the naphtha, the tank needs to be gas blanketed. If the feed is not pre-stripped, even minute levels of oxygen and/or olefins can lead to polymerization of olefins in the combined feed/reactor effluent exchangers or in the storage tank when kept for extended periods of time. Fouling and a reduction in heat transmission efficiency follow from this.

Monitoring the feed's chloride content is also necessary for effective corrosion management, as the wash water section explains.

The feed(s) are usually gathered in the feed surge drum, which has a surge capacity that levels off the rates. A water boot is also included with the feed surge drum to aid in removing any extra water that may have come with the feed. Based on an interface level indication, the sour water is manually removed, usually to a sour water header.

A split range controller regulates the feed surge drum pressure to keep the pressure marginally above the naphtha's bubble point. That control valve will be opened to supply hydrogen or fuel gas to the drum in response to a low-pressure signal. The vent valve to the fuel gas header or relief header will open and the hydrogen or fuel gas valve will close in response to a high-pressure signal. Both valves ought to be closed while the system is steady.

The feed surge drum bottom is where naphtha is sent out to the charge pumps. With a level indication on this vessel, the feed surge level is normally not regulated and is permitted to vary. During initial startup, the catalyst can be sulfided using the sulfur injection connection located at the suction of the charge pumps. There might be a standard sulfide injection pump for systems with extremely low feed sulfur contents. The goal of the sulfide injection rate is to sustain the catalyst metals at their ideal weight percentage, which is at least 15-20 weight ppm.

To prevent damage to this pump, there is a minimum flow spillback line that runs from the charge pump discharge back to the feed surge drum. A flow-indicating controller sets the flow rate to the reactor. To keep the unit from depressurizing, low flow will close the combined-feed exchanger control valve and feed intake.

- **Reactor**

The hydrogen-rich gas stream and naphtha feed from the charge pump are mixed together and sent into the combined feed exchangers, usually on the shell side, where they are heated. After emerging from the exchanger as a vapor, the combined feed is heated to the required reaction temperature in the charge heater. The temperature of the mixed feed that exits the charge heater and enters the reactor controls how much fuel burns in the heater. The fuel gas pressure controller of the charge heater is adjusted by a temperature controller. In order to enhance heater firing control by gently cooling the total combined feed entering the charge heater, a portion of the combined feed may occasionally bypass the combined feed exchanger.

After entering the reactor, the combined feed passes via the catalyst bed. In straight-run naphtha processing, the temperature across the catalyst bed usually stays constant. Usually on the tube side, the reactor effluent enters the combined feed/reactor effluent exchangers where it is cooled. The product condenser cools it even further in order to get ready for the gas-liquid separation process. In order to prevent salt buildup in the condenser and to dilute any hydrogen chloride, a wash water injection point is positioned in the reactor effluent line that leads to the product condenser.

2.5.3 Wash Water System

Three wash water injection points are positioned in the reactor effluent line: one immediately upstream of the product condenser, and two at the combined feed exchanger. In addition to preventing salt buildup in the condenser, process lines, and combined feed exchangers, the wash water dilutes any hydrogen chloride that may be present. The wash water injection point is usually placed immediately following the final combined feed exchanger bundle; however, this should be verified by determining the temperature at which ammonium chloride desublimates and the dew point.

With a flow indicating controller, the wash water injection pump should continuously inject fresh water into the system, usually at a rate of 3% of the charge rate by liquid volume. When processing feeds containing some organic chloride, usually less than 20 weight ppm, the volume of wash water injected is sufficient to prevent salt buildup and dilute hydrochloric acid. Alternative techniques are required for controlling chloride corrosion if feed chloride levels are high. A level control system regulates the amount of cold condensate that is supplied to the wash water tank.

To guarantee effective corrosion control, the sour water from the separator should be routinely observed, as specified in the analytical schedule in Section IX. The separator sour water is intended to have a pH of 5.5 to 6.5. When the process stream cools, failing to maintain this pH range can cause equipment and pipes for reactor effluent to corrode and possibly burst. When feed chloride levels are less than 20 weight ppm, maintaining the proper pH is usually simple. Hydrogen chloride can be further diluted by adjusting the wash water injection rate, however the rate shouldn't fall below three liquid volume percent of the input rate. To guarantee that at least 25% of the injected water stays in the liquid phase, the rate needs to be raised if the injection site is shifted to a hotter area. Get in touch with UOP for further details on controlling chloride corrosion.

Water that has been injected and reactor effluent flow to a product condenser before entering the separator. The injected water is collected by a water boot in the product separator and is usually directed to a sour water stripper for interface level control disposal. At this point, wastewater quality needs to be observed.

2.5.4 Separator System

After entering the separator, the reactor effluent and injected water leave the product condenser at a temperature low enough to guarantee full naphtha recovery. A mesh blanket coalescer used in the separator guarantees total separation of water, hydrocarbon liquid, and gas.

2.5.5 Pressure Control

A pressure-indicating controller at the separator controls the pressure in the reactor circuit. Two possible scenarios are presented about the addition of make-up gas to the Naphtha Hydrotreating Unit. The composition gas's pressure determines the variations. A pressure controller directly controls the rate of make-up hydrogen entering the unit to replenish the hydrogen used up by the processes when the make-up gas pressure is higher than the separator pressure, ensuring constant pressure (Figure IV-1).

Usually, the reactor effluent line just upstream of the product condenser receives the make-up gas, which is obtained from the Platforming Unit's Net Gas Chloride Treaters. Additionally, the separator has a manually operated gas effluent line valve that, though it is normally closed, can be used in an emergency to depressurize the unit to the relief header.

A make-up compressor is required to compress the gas in units where the separator pressure is higher than the make-up hydrogen pressure. The make-up hydrogen needs to be compressed because the Platforming Unit runs at a significantly lower pressure than the Naphtha Hydrotreating Unit. This hydrogen, which originates from the Platforming Unit's Net Gas Chloride Treaters, is added to the reactor effluent line just upstream of the product condenser via the Make-up Gas Compressor Drum and Make-up Gas Compressor.

A monel mesh blanket is included in the Make-up Gas Compressor Suction Drum to catch any liquid droplets before they go into the reciprocating compressors. The low signal selector regulates the Make-up Gas Compressor's spillback valves by receiving signals from the separator pressure and the drum pressure of the compressor. More make-up hydrogen can enter the unit when the spillback control valve closes in response to a drop in the pressure signal. The controller opens the spillback control valves to lower the make-up hydrogen flow rate if the separator pressure becomes too high. The Make-up Gas Compressor is kept from overheating by a water-cooled exchanger in the spillback line.

2.5.6 Recycle Gas

There are other ways to feed the reactor with the required amount of hydrogen-rich gas. Utilizing a Recycle Gas Compressor, the most popular technique joins the naphtha feed upstream of the combined feed/reactor effluent exchanger. The compressor draws suction from the top of the product separator. Figure IV-1 provides an illustration of this flow system. After leaving the separator, the gas enters a recycle gas compressor suction drum that is lined with a monel mesh blanket to catch any liquid droplets before entering the recycle gas compressor itself.

A monel mesh blanket is inserted into the Recycle Gas Compressor Suction Drum to filter out any liquid droplets that become entrained before they reach the reciprocating compressors. In addition, this drum has two trays and water addition connections, which are utilized for catalyst regeneration. Any condensed liquid is normally manually directed in batches to the Stripper Column. Usually, there are two reciprocating compressors for recycling gas that are single-stage and can operate at 50–100% of the intended flow rate.

2.5.7 Once-Through Gas

In certain units, an equivalent volume of a hydrogen-rich gas stream is injected into the unit with flow control in place of a recycle gas compressor. This gas passes through the reactor portion once and then into the product separator, where pressure control determines how much of it is evacuated. In Figure IV-2, this flow arrangement is displayed.

Each unit's design phase determines which of these flow schemes to use based on factors including the expense of compressing each stream and the availability of a high-pressure hydrogen-rich gas stream.

2.5.8 Stripping Section

After being pumped from the separator into the stripper, the liquid hydrocarbon is heated by a reboiler, which is often a fired heater. The liquid vaporizes due to the heat, removing contaminants such as water, light hydrocarbons, and hydrogen sulfide. Following its ascent to the overhead condenser, the vapor gathers in the overhead receiver. All of the liquid in the receiver is pushed back into the stripper as reflux because no net overhead liquid product is created. This reflux aids in further purifying the drink. The level of liquid in the receiver regulates the quantity of reflux, and more heat input from the reboiler is needed to increase the reflux.

A pressure differential controller regulates the reboiler's heat output, ensuring that the proper quantity of vaporization happens. The stripper's overhead gas is transferred for additional processing, such as fuel gas and amine scrubbing. The overhead vapor pipe is treated with a corrosion inhibitor to stop corrosion. After passing via a heat exchanger, the material from the stripper bottoms is transferred to the Platforming Unit for additional processing. You can save a tiny amount of the stripper bottoms to be used as delicious naphtha later on.

For the stripped naphtha product to be used as feed for the Platforming Unit, it needs to adhere to a set of requirements. Among these requirements are:

- Total Sulfur: less than 0.5 ppm
- Total Nitrogen: less than 0.5 ppm
- Chlorides: less than 0.5 ppm
- EP (electropositive) properties: 400 max
- Lead: less than 20 ppb
- Arsenic: less than 1 ppb
- Iron Chloride: less than 1 ppm
- Copper and Heavy Metals: less than 25 ppb
- Water and total oxygen: low enough, without water injection, to generate less than 5 mole ppm of water in the Platforming Unit recycle gas.

These requirements guarantee that the product is appropriate for additional processing in the Platforming Unit."

2.5.9 Splitter Section

Before being sent to the Platforming Unit, the material from the stripper bottoms may occasionally contain C5 and other compounds that need to be separated. The hydrotreated naphtha is fractionated into light and heavy naphtha in the Naphtha Splitter. Typically, the heavy naphtha is transferred to the Platforming Unit while the light naphtha is sent to the gasoline blending station. With the majority of the light naphtha being C5, and the heavy naphtha being C6+, the splitter is intended to separate C5 and C6 components. Refiners can restrict the amount of specific chemicals like benzene, methyl-cyclopentane, and cyclohexane in the heavy naphtha product by designing the splitter to separate the C6 and C7 components.

The material from the stripper bottoms heats the Naphtha Splitter feed, which is then forced into the column. Heat input from steam is produced, and the flow of steam condensate regulates this process. A total net overhead flow controller regulates the splitter receiver, which is where the overhead vapor condenses and gathers. The amount of reflux back into the column is controlled by this controller. A temperature controller located at the top tray of the column regulates the flow of the light naphtha product.

A pressure controller on the overhead line regulates the splitter pressure, and any non-condensable or off-gas can be vented to a relief header. After cooling, the heavy naphtha product is piped to the Platforming Unit and may alternatively be delivered to tankage. Before being transferred to tankage, the heavy naphtha is typically cooled using an air cooler and a trim cooler.

CHAPTER 3

MATERIAL BALANCE

Based on the idea of mass conservation, material balance is a technique for keeping track of and accounting for all materials coming into and going out of a system. This technique allows us to accurately account for all inputs and outputs by determining the flow rates and compositions of unknown material streams within the system.

- Basis = 1hr
- All equipment operates at steady state. This implies that in the equation,
  $Input + Generation - Output - Consumption = Accumulation$
- The Accumulation term amounts to zero.
- There are no material losses.

3.1 Reactor Section

The mixture of hydrocarbons (Naphtha/Pure Hydrocarbons) and other chemicals, mostly hydrogen, together with trace amounts of other components, makes up the reactor inlet stream. The reactor has entirely transformed or eliminated the incoming components pyrrole, pyridine, thiophene, and ethyl Mercaptan. The principal byproducts of the reactor are water, NH_3, H_2S, hydrogen, and naphtha/pure hydrocarbons. In the reactor, water is created.

Table 1: Material balance on reactor

REACTOR SECTION						
	INLET STREAM			**OUTLET STREAM**		
Components	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**
		kgmole/hr	*kg/hr*		*kgmole/hr*	*kg/hr*
Ethyl Mercaptan	0.0147	10.567	656.6849	0	0	0
Thiophene	0.0114	8.195	689.5660	0	0	0
Pyrrole	0.005	3.594	241.1550	0	0	0
Pyridine	0.0064	4.60096	363.935	0	0	0
Hydrogen	0.5987	430.405	0	0.52638	349.674	699.348
Naphtha	0.3638	261.537	26430.65	0.43303	287.668	26125.1
NH_3	0	0	0	0.01233	8.19496	278.49
H_2S	0	0	0	0.02826	18.762	1279.01
Total	1	718.9	28382	1	664.3	28382

3.2 Mixer Section

The reactor's output stream, which contains hydrogen, NH_3, H_2S, pure hydrocarbons, and water, enters the mixer through its first entrance. The mixer is filled with a stream of clean water.
The reactor exit stream and extra water are combined in the mixer. Together with the additional water, the mixer's output consists of the same parts as the reactor's output.

Table 2: Material balance on mixer

MIXER SECTION									
INLET 1 (REACTOR OUTLET STREAM)				INLET 2 (WASH WATER STREAM)			MIXER OUTLET		
Components	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**
		kgmole/hr	*kg/hr*		*kgmole/hr*	*kg/hr*		*kgmole/hr*	*kg/hr*
Ethyl Mercaptan	0	0	0	0	0	0	0	0	0
Thiophene	0	0	0	0	0	0	0	0	0
Pyrrole	0	0	0	0	0	0	0	0	0
Pyridine	0	0	0	0	0	0	0	0	0
Hydrogen	0.52638	349.674	699.348	0	0	0	0.4842	349.67	699.35
Naphtha	0.4330	287.668	26125.153	0	0	0	0.3983	287.67	26125.15
NH_3	0.0123	8.19496	278.48784	0	0	0	0.0113	8.20	278.49
H_2S	0.0282	18.762	1279.0125	0	0	0	0.026	18.76	1279.0125
Water	0	0	0	1	152.8	1180	0.0802	152.8	1180
Total	1	664.3	28382	1	152.8	1180	1	817.1	29562

3.3 Separator Section

By dividing the components into three separate streams according to their respective characteristics, the separator maximizes the recovery of hydrogen and naphtha/pure hydrocarbons while isolating the waste-water that contains NH_3, H_2S.

A stream of hydrogen that is highly pure and has few contaminants.

Naphtha Stream: Hydrocarbons with very little contaminants and high purity.

Sour Water Stream: This waste stream is concentrated in NH_3, H_2S, and water.
In order to ensure the intended separation of hydrogen, hydrocarbons, and waste

components, the table illustrates how well the separation process directs the components to their proper streams.

Table 3: Material balance on separator

SEPARATOR SECTION												
INLET STREAM				**HYDROGEN STREAM**			**NAPHTHA STREAM**			**SOUR WATER STREAM**		
Components	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**
		kgmole/hr	*kg/hr*		*kgmole/hr*	*kg/hr*		*kgmole/hr*	*kg/hr*		*kgmole/hr*	*kg/hr*
Ethyl Mercaptan	0	0	0	0	0	0	0	0	0	0	0	0
Thiophene	0	0	0	0	0	0	0	0	0	0	0	0
Pyrrole	0	0	0	0	0	0	0	0	0	0	0	0
Pyridine	0	0	0	0	0	0	0	0	0	0	0	0
Hydrogen	0.4842	349.67	699.3	0.8412	427.24	854.49	0.0015	0.38	0.75	0	0	0
Naphtha	0.3983	287.66	26125.15	0.1051	53.38	6121.90	0.9909	248.42	20667.30	0	0	0
NH_3	0.0113	8.19	278.49	0.014	7.11	121.73	0.0027	0.68	11.59	0.0249	1.46	24.94
H_2S	0.026	18.76	1279.01	0.0397	20.16	686.87	0.0042	1.05	37.02	0.0002	0.01	0.398
Water	0.0802	152.8	1180	0	0	0	0.0007	0.17	3.33	0.975	57.05	1031.66
Total	1	817.1	29562	1	507.9	7785	1	250.7	20720.	1	58.52	1057.

3.4 Stripper Section

Stripping Procedure: The stripper efficiently directs the heavier components (NH_3, H_2S, and Water) into the Bottom Stream and then to the reformer or for recycling component, while separating the lighter components (Hydrogen and Naphtha/Pure Hydrocarbons) into the Top Stream.

Distribution:

- ***Top Stream:***

  Rich in Hydrogen and Naphtha/Pure Hydrocarbons.
- ***Bottom Stream:***

  Contains heavier hydrocarbons, NH_3, H_2S, and Water.
- ***Reflux Streams:***

  Portions of the streams are recycled to optimize the separation process.

- ***To Reformer:***

  The majority of the stream that is sent toward the reformer is made up of heavier substances and hydrocarbons (NH_3, H_2S, Water). The stripper section's efficient component separation and recycling are shown in the table, guaranteeing proper distribution, and setting up the area for later processing steps.

Table 3.4: Material balance on stripper

STRIPPER SECTION																		
INLET STREAM				**TOP STREAM**			**BOTTOM STREAM**			**BOTTOM REFLUX**			**TOP REFLUX**			**TO REFORMER**		
Components	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**
		kgmole/hr	*kg/hr*		*kgmole/hr*	*kg/hr*		*kgmole/hr*	*kg/hr*		*kgmole/hr*	*kg/hr*		*kgmole/hr*	*kg/hr*		*kgmole/hr*	*kg/hr*
Hydrogen	0.0015	0.37605	0.7521	0.0252	0.37605	0.7521	0	0	0	0	0	0	0.0028	0.0146552	0.0293104	0	0	0
Naphtha	0.9909	248.41863	20668.71355	0.8672	20.07906208	198.53421	0.9998	248.193	20468.49129	1	14.569	1237.648205	0.9631	5.0408654	26.21250008	1	233.624	19232.53277
NH_3	0.0027	0.67689	11.50713	0.0043	0.676890	11.50713	0	0	0	0	0	0	0.003	0.015702	0.26881824	0	0	0
H_2S	0.0042	1.05294	35.8684011	0.0354	1.00293	35.8684011	0.0002	0.05001	1.688052468	0	0	0	0.0278	0.1455052	5.115962832	0	0	0
Water	0.0007	0.17549	3.15882	0.0679	0.17549	3.15882	0	0	0	0	0	0	0.0033	0.0172722	0.3281718	0	0	0
Total	1	250.7	20720.00	1	22.310	249.8206611	1	248.24301	20470.17934	1	14.569	1237.648205	1	5.234	31.95	1	233.624	19232.53277

3.5 Separator Section

Based on the phase and chemical characteristics of each component, the separator efficiently divides the mixture into distinct streams. Significant volumes of ammonia and the majority of hydrogen and H_2S are captured by the off-gas stream. Naphtha hydrocarbons make up the majority of the stream's composition, with trace amounts of other substances. Water makes up the majority of the sour water stream, with trace amounts of hydrogen, naphtha, ammonia, and H_2S. This distribution makes sure that every part is handled or processed appropriately in the operation's later phases.

Table 4: Material balance on separator section

SEPARATOR SECTION												
INLET STREAM				**OFF GASES STREAM**			**NAPHTHA STREAM**			**SOUR WATER STREAM**		
Components	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**MolarFlows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**	**Mole Fractions**	**Molar Flows**	**Mass Flows**
		kgmole/hr	*kg/hr*		*kgmole/hr*	*kg/hr*		*kgmole/hr*	*kg/hr*		*kgmole/hr*	*kg/hr*
Hydrogen	0.0252	0.376	0.7521	0.362	3.3822	6.764	0.003	0.0146	0.02931	0	0	0
Naphtha	0.8672	20.079	198.53	0.071	0.6654	46.7144	0.963	5.0408	26.212	0	0	0
NH3	0.0043	0.676	11.50	0.296	2.7615	30.708	0.003	0.0157	0.26881	0.014	0.1431	2.47991
H2S	0.035	1.002	35.868	0.267	2.4912	77.72	0.027	0.1455	5.1159	0.001	0.0098	0.33461
Water	0.067	0.175	3.158	0.002	0.0195	0.35	0.003	0.017	0.32817	0.984	9.6540	52.79
Total	1	22.31	249.82	1	9.3200	162.27	1	5.234	31.95	1	9.807	55.61

CHAPTER 4

ENERGY BALANCE

Because energy can be created or spent during a chemical reaction, energy conservation is not the same as mass conservation. If energy is produced or consumed during the operation, the total enthalpy of the outflow streams will not match that of the entrance streams. Energy balances are created during process design in order to ascertain the process's energy needs, including electricity, heating, and cooling. An energy balance on the plant will indicate areas for savings and conservation while also displaying the energy usage trend.

4.1 CONSERVATION OF ENERGY

A general equation can be written for the conservation of energy:

Energy out = Energy in + Generation – Consumption – Accumulation

The accumulation term for processes in a steady state will be zero. The fact that energy can take on various forms complicates the concept of energy balance. The energy balance equation can be simplified to the following form, which is enough for determining the heating and cooling needs of the various unit activities involved in chemical processes:

$H_2 – H_1 = Q – W$

This equation can be used to predict the energy requirements for batch processes because it works for both static (non-flow) and flow systems. The work term in a lot of processes will be 0 or extremely minor, thus the equation simplifies to the following basic heat balance equation:

$Q = H_2 – H_1$

Where:

– H_1 = enthalpy of the inlet stream.

– H_2 = enthalpy of the outlet stream.

– (positive) for heat entering the system.

– (negative) for heat leaving the system.

Assumption:

- Heat losses to the surrounding are neglected.

- Specific heats of solids are assumed to be constant.

- Specific heat of water and water vapor calculated at (T + To) /2

4.2 Energy Balance Calculation:

We calculate the specific heat C_P (kJ/kg. K) for feed and products in various temperatures (T) according to the following equation:

$C_P = a + bT + CT^2$

4.2.1 PUMP

The duty of pump was calculated using the Bernoulli's equation given as under:

Bernoulli's Equation:

$$\eta Q_{pump} = \dot{m}(\Delta Z + \Delta P\rho g + \Delta\vartheta 2g) + Q_{losses}$$

$$\eta Q_{pump} = \dot{m}(\Delta P\rho)$$

Table 5: Energy balance on pump

Pump1			
Parameters	**Units**	**Inlet**	**Outlet**
Temperature	*(C)*	58	59.32
Pressure	*(N/m2)*	275.5669	2710.5581
Density	*(kg/m^3)*	636.3	636.3
Mass Flow Rate	*(kg/h)*	26654.00	26654.00
C_P	*(kJ/kg*C)*	2.271	2.271
Adiabatic Efficiency	*(η)*	75%	75%
Pump Duty	*(kJ/hr)*	22.289	-

Table 6: Energy balance on pump 2.

Pump - 2			
Parameters	**Units**	**Inlet**	**Outlet**
Temperature	*(°C)*	141	141
Pressure	*(N/m2)*	806.1066	1361.163
Density	*(kg/m^3)*	14.32	14.32
Mass Flow Rate	*(kg/h)*	26654.00	26654.00
Cp	*(kJ/kg*C)*	2.271	1.33965
Adiabatic Efficiency	*(η)*	75%	75%
Pump Duty	*(kJ/s)*	-800.1254388	-

4.2.2 Heat Exchanger

The heat loss in the tube side and the heat gain in the tube side were calculated using the equation:

$Q = mCp\Delta T$

Table 7: Energy balance on heat exchanger

Heat Exchanger					
Parameters	**Units**	**Inlet -Shell**	**Outlet-Shell**	**Inlet-Tube**	**Outlet-Tube**
Temperature	*(°C)*	64	288	343	108
Pressure	*(N/m2)*	2710.5581	2433.0299	2079.99	1775.98
Mass Flow Rate	*(kg/h)*	28382.00	28382.00	28383.00	28382.00
C_p	*(kJ/kg*C)*	3.328566855	3.67387834	3.553248385	0.923134911
Q	*KJ/hr*	3684383.994	27423539.96	32070887.95	8331731.987
Q	*(kJ/s) - KW*	1023.439998	7617.649988	8908.579986	2314.369996

Del Q Shell Side	6594.21

Del Q Tube Side	6594.21

4.2.3 Charge Heater

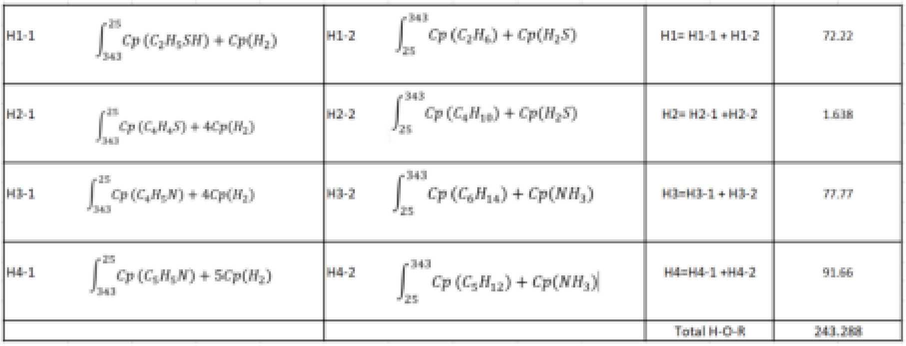
The heats of the inlet stream and the fuel given were calculated and the sum of both was equal to the heat of the outlet stream.

Table 8: Energy balance on charge heater

Charge Heater				
Parameters	**Units**	**Fuel In**	**Stream In**	**Stream Out**
Temperature	*(˚C)*	55	288	343
Pressure	*(N/m2)*	2710.5581	2433.0299	2079.99
Mass Flow Rate	*(kg/h)*	28382.00	28382.00	28383.00
Cp	*(kJ/kg*C)*	2.627665484	3.67387834	3.567164501
Q	*KJ/hr*	4772952.022	27423539.96	32196491.95
Q	*(kJ/s) - KW*	1325.820006	7617.649988	8943.469986

4.2.4 Reactor

The input and outflow streams' heat contents were computed. Additionally, the following formulae were used to compute the Heat of Reactions. The outlet stream's heat and the total heat of reactions added up to the entrance stream's heat.



H1-1 $\int_{343}^{25} Cp\,(C_2H_5SH) + Cp(H_2)$	H1-2 $\int_{25}^{343} Cp\,(C_2H_6) + Cp(H_2S)$	H1= H1-1 + H1-2	72.22
H2-1 $\int_{343}^{25} Cp\,(C_4H_4S) + 4Cp(H_2)$	H2-2 $\int_{25}^{343} Cp\,(C_4H_{10}) + Cp(H_2S)$	H2= H2-1 +H2-2	1.638
H3-1 $\int_{343}^{25} Cp\,(C_4H_5N) + 4Cp(H_2)$	H3-2 $\int_{25}^{343} Cp\,(C_6H_{14}) + Cp(NH_3)$	H3=H3-1 + H3-2	77.77
H4-1 $\int_{343}^{25} Cp\,(C_5H_5N) + 5Cp(H_2)$	H4-2 $\int_{25}^{343} Cp\,(C_5H_{12}) + Cp(NH_3)$	H4=H4-1 +H4-2	91.66
		Total H-O-R	243.288

Figure 4-1: Reactions and heats of reaction

Table 4.5: Energy balance on reactor

Reactor			
Parameters	**Units**	**Inlet Stream**	**Outlet Stream**
Temperature	*(˚C)*	343	343
Pressure	*(N/m2)*	2079.99	2079.99
Mass Flow Rate	*(kg/h)*	28383.00	28383.00
Cp	*(kJ/kg*C)*	3.567164501	3.470121293
Q	*KJ/hr*	32196491.95	31320599.95
Q	*(kJ/s) - KW*	8943.469986	8700.166653

Heat of Reaction

Table 9: Heat of Reaction

Heat Of Reactions *(kJ/hr)*	
Reaction-thiophene	72.22
Reaction-E-Merc	1.638
Reaction-Pyrrole	77.77
Reaction-Pyridine	91.66
Total	243.288

Reactor Balance = Q(inlet) + Heats of reactions

Table 10: Reactor balance

Q (inlet)	8943.47
Q (outlet)	8700.167
Reactor Balance	8943.455

Reactor Balance:

Q (outlet) = Q (inlet) + Heats of Reactions

Where:

− Q (inlet) = 8943.47

− Q (outlet) = 8700.167

− Heats Of Reactions = 243.167

4.2.5 Compressor

The compressor work was calculated using the equations:

$$W = Z \mathbf{R} T_1 \frac{n}{n-1}\left[\left(\frac{P_2}{P_1}\right)^{(n-1)/n} - 1\right]$$

$$\gamma = \frac{C_p}{C_V}$$

$$m = \frac{(\gamma - 1)E_p}{\gamma}$$

$$n = \frac{1}{1-m}$$

Figure 4-2: Compressor work calculation formula

Table 11: Energy balance on compressor

Compressor			
Parameters	**Units**	**Stage-1**	**Stage-2**

Compressibility factor	*Z*	0.9995	0.9985
General Gas Constant (R)	*kJ/kmol. K*	8.314	8.314
Heat Capacity Ratio	γ	1.401	1.404
Polytropic Efficiency	*Ep*	0.76	76%
Inlet Temperature (T_1)	*˚C*	58	93
Inlet Pressure (P_1)	N/m^2	1724.0091	2223.16
Outlet Pressure (P_2)	N/m^2	2223.16	2722.326
Polytropic Temperature Exponent	*M*	0.21753033	0.218689459
Polytropic Exponent	*N*	1.278004816	1.279900817
Polytropic Work (W)	*kJ/kmol*	126.006974	159.8979553
Polytropic Work (W)	*kJ/hr*	904100.044	1147267.829
Polytropic Work (W)	*kJ/s*	251.1389013	318.6855081
Pressure Ratio	P_2/P_1	1.28952915	1.224529948
Pressure Ratio Power	*(n-1)/n*	0.21753033	0.218689459
n/(n-1)	*n/(n-1)*	4.59705998	4.572694112
Z*R*T1	*kJ/kmole*	481.970894	772.042197
Molar Flow	*kmole/hr*	7175	7175
Actual Work Required	*KW*	-330.445922	-419.3230369

4.2.6 Condenser

Condenser balance was carried out by equating inlet and outlet heat flows using the equations:

$Q = mCp\Delta T$ and $Q = m\lambda$

Table 12: Energy balance on condenser.

Condenser			
Parameters	**Units**	**Inlet Stream**	**Outlet Stream**
Temperature	*(˚C)*	77	58
Pressure	*(N/m2)*	2079.99	2079.99
Mass Flow Rate	*(kg/h)*	28383.00	28383.00
C_P	*(kJ/kg*C)*	3.567164501	3.470121293
Q	*KJ/hr*	32196491.95	31320599.95
Latent heat	*kJ/kg*	0	243.291
Q	*(kJ/s) - KW*	8943.469986	8943.457653

4.2.7 Stripper

Energy balance on stripper is given in table.

Table 13: Energy balance on stripper

Stripper		
Units	kJ/hr	kJ/s - Kw
Inlet Heat Flow	6.26E+07	1.74E+04
Condenser Heat Flow	2.92E+07	8.11E+03
Reboiler Heat Flow	3.34E+07	9.27E+03

CHAPTER 5

EQUIPMENT DESIGN

5.1 Heat Exchanger

The following calculations are done to design the heat exchanger E-005 which is used to cool the bottom of distillation column by exchanging heat with the feed of column.

Mass flow of hot fluid = 25000 kg/h.
Inlet and outlet temperatures of hot fluid are 130.8 °C and 89 °C respectively.

Mass flow of cold fluid = 26460 kg/h.
Inlet and outlet temperatures of cold fluid are 57.34 °C and 90 °C respectively.

Tube outer diameter = ¾ in.	Pitch = 1 in. square
Type of tubes = 13 BWG	Length of a tube = 16 ft.
No. of tubes = 160	No. of passes = 2
Shell inner diameter = 17 ¼ in.	Baffle spacing = 5 in.

5.1.1 Heat Balance:

$$Q = m\,Cp\,\Delta T = 28382 * 2.6455 * (345\text{-}195) = 9457133\ kJ/h$$

$$Q = 8963631.42 \text{ BTU/hr}$$

5.1.2 ΔT Calculation:

Table 14: ΔT calculation

Hot Fluid		Cold Fluid	Difference
345	**Higher Temp**	160	180
195	**Lower Temp**	57	138
150	**Difference**	103	42

5.1.3 LMTD Calculations:

$$LMTD = \frac{\Delta T_1 - \Delta T_2}{\ln\left(\frac{\Delta T_1}{\Delta T_2}\right)} = \frac{180 - 138}{\ln\left(\frac{180}{138}\right)}$$

$$LMTD = 158.07\ ℃$$

$$R = \frac{T_1 - T_2}{t_2 - t_1} = \frac{150}{103} = 1.45$$

$$S = \frac{t_2 - t_1}{T_1 - t_1} = \frac{103}{345 - 57} = 0.357$$

$$F_t\ = R - S$$
$$= 1.45 - 0.357$$
$$F_t\ = 1.093$$

$$\Delta t = F_t\ x\ LMTD$$
$$\Delta t = 172.77\ ℃$$

5.1.4 Calorific Temperatures:

$$\frac{\Delta t_C}{\Delta T_h} = \frac{138}{180} = 0.76$$

$$K_c = 1\ , F_c = 0.44$$

$$T_c = T_2 + F_c\ \times (T_2 - t_1) = 195\ \times 0.44\ \times 150 = 261℃ = 501.8\ °F$$

$$t_c = t_2 + F_C\ \times\ (\mathrm{t}_2 - t_1) = 57 + 0.44\ \times\ 103 =\ 102.3\ ℃\ = 216.14\ °F$$

Shell side (Cold Fluid)

5.1.5 Flow Area:

$$a_s = I.D\ \times\ \mathrm{C}'' \times \frac{\mathrm{B}}{144\ P_T}$$

$$= 17.25\ \times\ 0.25\ \times \frac{5}{144\ \times\ 1}$$

$$a_S = 0.1497\ ft^2$$

5.1.6 Mass Velocity:

Tube Side (Hot Fluid)

Flow Area:

$$a_t' = 0.247\ in^2$$

$$a_t = \frac{N\ \times\ \mathrm{a}_\mathrm{t}'}{144\ \times\ \mathrm{N}}$$

$$a_t = 0.137\ ft^2$$

Mass Velocity:

$$G = \frac{W}{a_s}$$

$$= \frac{62571.69}{0.1497} = 417980 \frac{lb}{hr.ft^2}$$

5.1.7 Reynold's Number:

At $T_c = 501.8\ °F$

$$\mu = 0.3773 \frac{lb}{ft.hr}$$

$$D_e = 0.079\ ft$$

$$R_{e_s} = \frac{D_e G_S}{\mu}$$

$$= \frac{0.079 \times 417980}{0.3773}$$

$$R_{e_s} = 8751.76$$

5.1.8 JH

$J_H = 170$

At $T_c = 226$°F

$c = 0.624\ {}^{Btu}/_{lb.}$°F

$k = 0.0502\ {}^{Btu}/_{hr.\ ft.}$°F

$$\left(\frac{c\ .\mu}{k}\right)^{\frac{1}{3}} = 1.654$$

5.1.9 Heat Transfer Coefficient:

$$\frac{h_0}{Ø_s} = \frac{J_h\ K}{D_e} \left(\frac{c\ .\mu}{k}\right)^{\frac{1}{3}}$$

$$\frac{h_0}{Ø_s} = 178.2$$

$$G = \frac{W}{a_s}$$

$$= \frac{62571.69}{0.137} = 456729.19 \frac{lb}{hr.ft^2}$$

Reynold's Number:

At $t_c = 216.14\ °F$

$$\mu = 0.2042 \frac{lb}{ft\ .hr}$$

$$D = 0.0467\ ft$$

$$R_{e_t} = \frac{DG_t}{\mu}$$

$$R_{e_t} = \frac{0.0467 \times 417980}{0.2042}$$

$$R_{e_t} = 97033.08$$

JH:

$\frac{L}{D} = 342.6$; $J_H = 120$

At $t_c = 161$°F

$c = 0.573\ {}^{Btu}/_{lb.}$°F

$k = 0.056\ {}^{Btu}/_{hr.\ ft.}$°F

$$\left(\frac{c\ .\mu}{k}\right)^{\frac{1}{3}} = 1.716$$

Heat Transfer Coefficient:

$$\frac{h_0}{Ø_s} = \frac{J_h\ K}{D_e} \left(\frac{c\ .\mu}{k}\right)^{\frac{1}{3}}$$

$$\frac{h_i}{Ø_i} = 247$$

5.1.10 Tube- wall Temperature:

$$t_w = t_c + \frac{\frac{h_0}{\emptyset_s}}{\frac{h_0}{\emptyset_s} + \frac{h_i}{\emptyset_i}} \times (T_c - t_c)$$

$$= 161 + \frac{178.2}{178.2 + 247} \times 399.5$$

$$t_w = 328.21\ ^{\circ}F$$

At t_w = 328.21 ° F

$$\mu_w = 0.451 \frac{lb}{ft.hr}$$

$$\mu_w = 0.451 \frac{lb}{ft.hr}$$

$$\emptyset_s = \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

$$\emptyset_s = \left(\frac{0.377}{0.451}\right)^{0.14}$$

$$\emptyset_s = 0.975$$

5.1.11 Corrected Coefficient:

$$h_0 = \frac{h_0}{\emptyset_s} \times \emptyset_s$$

$$h_0 = 178.2 \times 0.975$$

$$h_0 = 173.75 \frac{BTU}{hr.ft^2.^{\circ}F}$$

$$\frac{h_{i0}}{\emptyset_t} = \frac{h_i}{\emptyset_i} \times \frac{I.D}{O.D} = 247 \times \frac{0.560}{0.75}$$

$$= 247$$

Tube- wall Temperature:

$$t_w = t_c + \frac{\frac{h_0}{\emptyset_s}}{\frac{h_0}{\emptyset_s} + \frac{h_i}{\emptyset_i}} \times (T_c - t_c)$$

$$= 161 + \frac{178.2}{178.2 + 247} \times 399.5$$

$$t_w = 328.21\ ^{\circ}F$$

At t_w = 328.21 ° F

$$\mu_w = 0.432 \frac{lb}{ft.hr}$$

$$\emptyset_t = \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

$$\emptyset_t = \left(\frac{0.494}{0.432}\right)^{0.14}$$

$$\emptyset_t = 1.02$$

Corrected Coefficient:

$$h_{i0} = \frac{h_{i0}}{\emptyset_t} \times \emptyset_t = 184.4 \times 1.02$$

$$h_{i0} = 1881.1 \frac{BTU}{hr.ft^2.^{\circ}F}$$

|

5.1.12 Clean Overall Coefficient:

$$U_c = \frac{h_{i_0} \times h_0}{h_{i_0} + h_0}$$

$$U_c = \frac{173.75 \times 188.1}{173.75 + 188.1}$$

$$U_c = 90.3 \frac{BTU}{hr.ft^2.\ °F}$$

5.1.13 Design Overall Coefficient:

$$a'' = 0.1963 \frac{ft^2}{lin.ft^2}$$

$$\boldsymbol{Total\ Surface};$$

$$A = N \times \mathrm{L} \times \mathrm{a}''$$

$$A = 160 \times 16 \times 0.1963$$

$$A = 502.53\ ft^2$$

$$U_D = \frac{Q}{A\ .\Delta T} = \frac{8963631.42}{502.5 \times 342.98}$$

$$U_D = 52.01 \frac{BTU}{hr.ft^2.°F}$$

5.1.14 Dirt Factor ' Rd ' :

$$R_d = \frac{U_c - U_D}{U_c\ .U_D}$$

$$R_d = \frac{90.3 - 59.3}{90.3 \times 59.3}$$

$$R_d = 0.0058 \frac{hr.ft^2.°F}{BTU}$$

After this, pressure drop on both shell and tube side is found out to verify it that the design is feasible and in the allowable pressure drop limitation of less than 10 psi.

5.1.15 Pressure Drop

The following calculations are carried out to find out the pressure drop across both sides of the heat exchanger:

1)

$For\ R_{e_s} = 8751.76$

$f = 0.00145$

$D_s = 1.4375\,ft$

$s = 0.5874$

2) No. of crosses:

$$N + 1 = 12 \times \frac{L}{D}$$

$$N + 1 = 12 \times \frac{16}{5}$$

$$\mathrm{N} + 1 = 39$$

3) $\Delta \boldsymbol{P_s}$:

$$\Delta P_s = \frac{f\ .G_s^2\ .D_s\ .(N+1)}{5.22\ \times\ 10^{10}\ \times\ \mathrm{D_e}s\text{Ø}_{\mathrm{s}}}$$

$$\Delta P_s = 4.6\ psi$$

1)

$For\ R_{e_t} = 97033.08$

$f = 0.0002$

$s = 0.6208$

2) $\Delta \boldsymbol{P_t}$:

$$\Delta P_t = \frac{f\ .G_t^2\ .L\ .n}{5.22\ \times\ 10^{10}\ \times\ \mathrm{D_e}s\text{Ø}_t}$$

$$\Delta P_s = 0.75\ psi$$

3) $\Delta \boldsymbol{P_T}$:

$$G_t = 456729.19\frac{lb}{hr.ft^2}$$

$$\frac{V^2}{2g'} = 0.024$$

$$\Delta P_r = \frac{4n}{s} \times \frac{V^2}{2g'}$$

$$\Delta P_r = \frac{4 \times 2}{0.6208} \times 0.024$$

$$\Delta P_r = 0.031\ psi$$

$$\Delta P_T = \Delta P_r + \Delta P_T$$

$$\Delta P_T = 0.75 + 0.31$$

$$\Delta P_T = 1.1\ psi$$

5.2 Three Phase Separator

The calculations shown below are for the design of a three phase separator. First, we have to select the Retention time of both fluids:

$$Retention \text{ time of oil } = (\mathrm{t_r})_\mathrm{o} = 0.16 \text{ min } (10 \text{ sec})$$

$$\text{Retention time of water } = (\mathrm{t_r})_\mathrm{w} = 0.16 \text{ min } (10 \text{ sec})$$

5.2.1 Calculate (h_o)$_{max}$:

$$(h_{o_{max}}) = \frac{320\ t_{r_o}\ (\Delta S\,.G)}{\mu o}$$

$$T_{r_o} = 0.16\ min$$

$$\Delta S.G = (S.G)_w - (S.G)_o = 1 - 0.647$$

$$\Delta S.G = 0.353$$

$$Viscosity\ of\ oil = 0.2331\ C_p$$

$$h_{o_{max}} = \frac{320 \times 0.16 \times 0.353}{0.2331} = 77.5\ inch$$

Calculate $\frac{A_w}{A}$ (fraction of vessel cross-section occupied by the water phase).

$$\frac{A_w}{A} = \frac{0.5 \times \mathrm{Q_w} \times (\mathrm{t_r})_w}{(t_r)_o \times \mathrm{Q_o} + (t_r)_w \times \mathrm{Q_w}}$$

$$Q_w = 17.13\frac{kg}{hr} = 0.000168\frac{ft^3}{s}$$

$$\frac{A_w}{A} = \frac{0.5 \times 0.000168 \times 0.16}{0.16 \times 0.71 + 0.16 \times 0.000168}$$

$$\frac{A_w}{A} = 1.2 \times 10^{-6}$$

5.2.2 Calculating d_{max}:

$$d_{max} = \frac{(h_o)_{max}}{\frac{h_o}{d}}$$

$\frac{h_o}{d}$ is taken from the graph

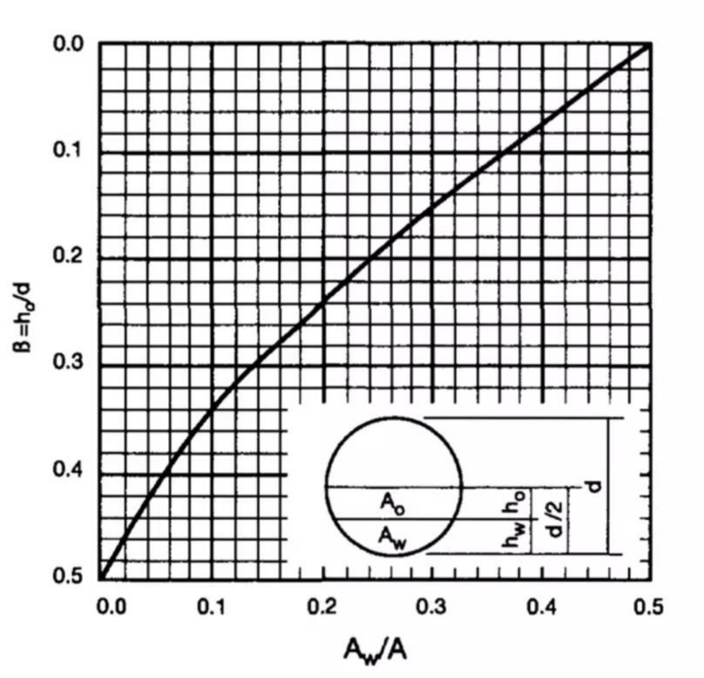


Figure 5-1: Graph for calculating dmax

Fig. Graph of $\frac{A_w}{A}$

From the graph the value of $\frac{h_o}{d}$ comes out to be 0.499 against the $\frac{A_w}{A}$ value 1.2×10^6

$$d_{max} = \frac{77.5}{0.499}$$

For gas capacity,

$$d\,.\,l_{eff} = 420 \times \left(\frac{T \times Z \times Q_g}{P}\right)\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right) \times \frac{C_D}{d_m}\right]^{\frac{1}{2}}$$

Calculating l_{eff} for $d<d_{max}$:

$$T = 57.34\ °C\ \boldsymbol{or}\ 594\ °R$$

$$P = 9.82 \frac{kg}{cm^2} = 139.7\ psia$$

$$Density\ of\ Gas = 5.535 \frac{kg}{m^3}\ \boldsymbol{or}\ 0.3455 \frac{lb}{ft^3}$$

$$Density\ of\ Liquid = 650\frac{kg}{m^3} = 41.2\frac{lb}{ft^3}$$

$$d_m = 100\ micron$$

$$Q_g = 380\frac{kg}{hr} = 0.01\ MMScf$$

$$Z = 0.997$$

$$C_D = 1.5$$

$$d.l_{leff} = 420\ \times\ \left(\frac{594\ \times 0.997\ \times\ 0.01}{139.7}\right)\left[\left(\frac{0.3455}{41.2\ \times 0.3455}\right)\times\frac{1.5}{100}\right]^{\frac{1}{2}}$$

$$d.l_{eff} = 201.2$$

For fluid Capacity;

$$d^2.l_{eff} = 1.42\ \times\ (t_r)_o \times Q_o + (t_r)_w \times Q_w$$

$$Q_o\ = 6279\frac{B}{D}$$

$$Q_w\ = 2.631\frac{B}{D}$$

$$d^2.l_{eff} = 1.42\ \times\ 0.16\ \times 6279 + 0.16\ \times 2.631 = 1427$$

5.2.3 Calculating L_{ss}:

$$L_{ss} = L_{eff} + \frac{d}{12}\ \text{(for gas capacity)}$$

$$L_{ss} = \frac{4}{3}\times\ L_{eff}\ \ \text{(for liquid capacity)}$$

Calculating slenderness ratio ($12.\frac{L_{ss}}{d}$) which would be between 3 and 5 for both gas and liquid capacities.

Table 15: Vessel length based on liquid capacity.

D	**L_{eff}**	**L_{ss}**	**$12.L_{ss}/d$**
(inch)	**(ft)**	**(ft)**	**-**
24	8.33	10.38	5.19
20	10.06	11.72	7.032
16	12.57	13.9	10.42
30	6.7	9.2	3.68

Table 16: Vessel length based on liquid capacity

D	**L_{eff}**	**L_{ss}**	**$12.L_{ss}/d$**
(inch)	**(ft)**	**(ft)**	**-**
24	2.47	3.2	1.6
30	1.58	2.06	0.82
20	3.56	4.63	2.78
16	5.57	7.2	5.4
18	4.4	5.7	3.8

A comparison of tables shows that liquid capacity is dominant parameter, so

$$18\ in\ \times\ 5.7\ ft\ \text{ vessel is sufficient}$$

$$18\ in\ \times\ 6\ ft\ \text{(rounded figure)}$$

5.2.4 Weir's Height:

$$\Delta h = h_o\ \left[1 - \frac{\rho_o}{\rho_w}\right]$$

We know that,

$$\frac{h_o}{d} = 0.499$$

$$h_o\ = 0.499\ .d$$

$$d = 18$$

$$h_o = 0.499\ .18 = 9\ inches$$

$$\Delta h = 9\,9 \times 0.535 = 3\ inches$$

So, oil weir should be placed at 9 inches while water weir should be places at 6 inches radial distances.

5.2.5 Demister Sizing:

$$V_m\ = K_d\sqrt{\frac{\rho_t\ -\ \rho_g}{\rho_g}}$$

K_d = Demister capacity factor, ft/s

V_m =maximum velocity, ft/s

K_d= 0.1 m/s = 0.35 ft/s

$$V_m = 0.35\sqrt{\frac{41.2 -\ 0.3455}{0.3455}}\ = 3.8\frac{ft}{s}$$

Now calculating the area of demister,

$$A_d = \frac{0.327 \times \frac{T \times Z \times Q_g}{P}}{V_m} = 3.65\ ft^2$$

A demister having this much area is required.

5.2.6 Nozzle Sizing:

Nozzle are generally sized depending on the momentum of the incoming feed. It normally lies in the range of 600 – 700 lb/ft . s^2

$$Momentum = \rho_m \times V_m^2$$

$$= 42 \times 3.8^2$$

$$= 606 \frac{lb}{ft.s^2}$$

5.3 Fixed Bed Reactor

The reactor we have chosen is isothermal fixed bed reactor.

The catalyst is Co-Mo with alumina base with name tag S-12 and the shape is extrudate form with size of 1/6 in.

The average bulk density of catalyst is 45 lb/ft^3.

We kmow that LHSV (Liquid Hourly Space Velocity) is given by,

$$LHSV = \frac{Volume\ of\ (liquid) feed\ flowing\ per\ hour}{Volume\ of\ catalyst}$$

For naphtha, LHSV is between 4-12 hr^{-1}. Taking a value of 8 , and the volume of feed flowing at STD is 43.91 m^3/h.

Hence the volume of catalyst is;

$$Volume\ of\ catalyst = \frac{43.91\frac{m^3}{h}}{\frac{4}{hr}}$$

$$Volume\ of\ catalyst = 10.98\ m^3$$

The weight of the catalyst is given by;

$$Weight\ of\ catalyst = Density \times\ Volume$$

$$Weight\ of\ catalyst = 45\frac{lb}{ft^3} \times\ 10.98\, m^3$$

Using conversion factors,

$$1\ lb = 0.453\ kg$$

$$1\ m^3\ = 35.31\ ft^3$$

So, we get,

$$Weight\ of\ the\ catalyst = 7903\ kg$$

On industrial scale plant a guard bed is usually required to remove impurities,

$$V_{guard-bed}\ = 0.2\ \times V_{catalyst}$$

$$V_{guard-bed}\ = 0.2\ \times 10.98\ m^3$$

$$V_{bed} = V_{catalyst} + V_{guard-bed}$$

$$V_{bed}\ = 10.98\ m^3\ + 2.2\ m^3$$

$$V_{bed}\ = 13.18\ m^3$$

Allowing for top and bottom clearance, considering 1-% space on entrance and 10% space on the exit of the reactor, we get volume of reactor as,

$$V_{reactor}\ = 20\%\ more\ of\ V_{bed} = 13.18\ m^3 + (0.2 \times\ 13.18\, m^3)$$

$$V_{reactor}\ \ = 15.8\ m^3$$

Dimensions of Reactor:

$$V\ = \left(\frac{\pi\ \times\ d^2}{4}\right) \times h$$

Where,

d = diameter of the reactor

h = height of the reactor

According to the literature, the is height is given as,

$$\frac{h}{d} = 2$$

So,

$$V = \left(\frac{\pi \times d^2}{4}\right) \times 2d$$

$$15.8\ m^3 = \left(\frac{\pi \times d^2}{4}\right) \times 2d$$

$$d = 1.71\ m$$

$$h = 6.84\ m$$

5.4 Fired Heater

In the design of the fired heater following main features should be considered:

- Radiant Section
- Convective Section
- Bridge wall Section
- Tubes, pipes and their specifications
- Burner
- Insulation and heat loss

5.4.1 Total hourly heat transfer to cold surface:

The total hourly heat transfer to cold surface is estimated by the relation,

$$q_T = 0.173\ f\left[\left(\frac{T_G}{100}\right)^4 - \left(\frac{T_s}{100}\right)^4\right] \alpha_{\mathrm{CP}} A_{CP} + hA(T_G - T_S)$$

Where,

$f =$ $overall\ exchange\ factor$

$T_G = Temperature\ of\ the\ flue\ gases\ leaving\ the\ radiant\ section\ in\ degree\ Rankine$

$T_S = Surface\ Temperature\ of\ the\ tubes\ in\ degree\ Rankine$

A_{cp} $= Equivalent\ cold\ plane\ surface\ in\ ft^2$

$\alpha_{\mathrm{CP}} =$ $factor\ by\ which\ A_{cp} must\ be\ reduced\ to\ obtain\ effective\ cold\ surface$

For natural convection,

$$h \approx 2\ , \qquad A \approx 2\alpha_{\mathrm{CP}} A_{CP} \qquad and \qquad f \approx 0.57$$

Therefore, the above equation can be re-written as,

$$\frac{\sum Q}{\alpha_{\mathrm{CP}} A_{CP} f} = 0.173\ \left[\left(\frac{T_G}{100}\right)^4 - \left(\frac{T_S}{100}\right)^4\right] + 7(T_G - T_S)$$

5.4.2 T_G and T_S Calculations:

T_S is assumed and T_G is calculated from the graph:

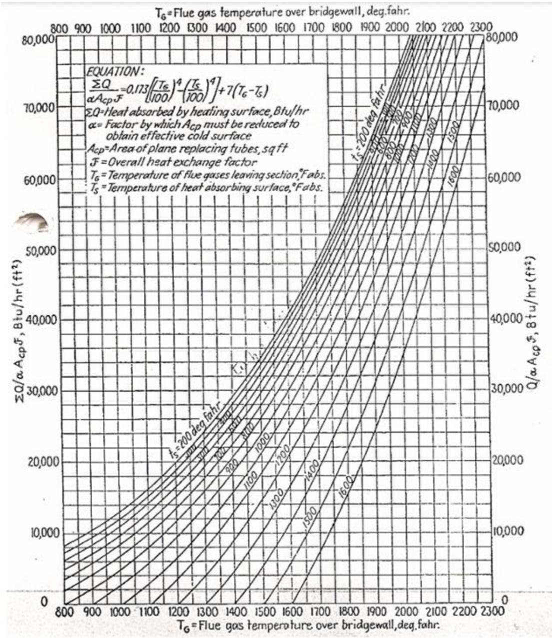


Figure 5-2: Tg and Ts calculation graph

To design the furnace, the following are taken into account:

- $Tube\ surface\ temperature = T_s = \ 331.6\ °C\ or\ 628.8\ °F$
- $Flue\ Gas\ temperature = T_G = \ 362.1\ °C\ or\ 683.78\ °F$
- $Total\ required\ heater\ duty\ \left(\frac{BTU}{hr}\right) = \ 331.6\ °C\ or\ 628.8\ °F$
- $Efficiency = \eta = \ 331.6\ °C\ or\ 628.8\ °F$
- $Fuel\ Value\ \left(\frac{BTU}{hr}\right) = \ 8.31 \times 10^6 \frac{BTU}{hr}$
- $Air\ to\ fuel\ ratio = 8.2$
- $Temperature\ of\ inlet\ air\ (\text{°F}) = 77\ \text{°F}$
- $Tubes\ diameter\ (in.) = d_o = 4\ in.$
- $Center - to - center\ distance\ (in.) = ctc = 4\ in.$
- $Exposed\ Tube\ length\ (ft) \ = \ L \ = \ 12.47\ ft.$
- $Average\ flux\ of\ the\ radiant\ section\ \left(\frac{BTU}{hr.ft^2}\right) = q = 5.87\ \times 10^7 \frac{BTU}{hr.ft^2}$
- $Average\ flux\ of\ the\ convective\ section\ \left(\frac{BTU}{hr.ft^2}\right) = q = 5.87\ \times 10^7 \frac{BTU}{hr.ft^2}$

Assuming,

$$\frac{\Sigma Q}{\alpha_{\text{CP}} A_{CP} f} = 2 \times \ (Average\ Flux)$$

$$Total\ hourly\ heat\ transfer\ to\ cold\ surface = 11.74 \times \frac{10^7 BTU}{hr.m^2}$$

5.4.3 Heat liberated by fuel:

$$Q_F = \frac{Total\ required\ heater\ duty}{\eta}$$

$$Q_F = 19543.14 \frac{BTU}{hr}$$

5.4.4 Amount of required fuel:

$$m_{fuel} = \frac{Q_F}{Fuel\ Value}$$

$$m_{fuel} = 0.12\ kg$$

5.4.5 Amount of air required:

$$m_{air} = \frac{air}{fuel} . m_{fuel}$$

$$m_{air} = 0.95\ kg$$

5.4.6 Excess Air:

Assuming 25% access air then:

$$Air\ required = 1.25\ . m_{air}$$

$$Air\ required = 1.09\ kg$$

5.4.7 Inlet heat by air:

$$Q_{air} = m_{air} . C_{p_{air}} (T_{air_{in}} - T_{ref})$$

$$Assuming\ T_{ref} = 60\ °\mathrm{F}$$

$$Q_{air} = 16.31 \frac{BTU}{hr}$$

5.4.8 Total steam required:

The amount of atomizing steam is 0.3 lb/lb fuel.

Therefore,

$$Steam\ required = 0.3\ m_{fuel}$$

$$Steam\ required = 0.036\ kg$$

5.4.9 Heat associated with steam:

$$Q_{steam} = m_{steam} . C_{p_{steam}} . (T_{steam_{in}} - T_{ref})$$

$$Q_{steam} = 1.81 \frac{BTU}{hr}$$

5.4.10 Heat absorbed by furnace walls:

$$Q_{wall} = 2\%\ Q_F$$

$$Q_{wall} = \ 3903\frac{BTU}{hr}$$

5.4.11 Heat Of Exhaust Gases:

$$Q_{exhaust} = m_{steam} \,.\, C_{p_{steam}} \,.\, (T_{steam_{in}} - T_{ref})$$

$$Q_{exhaust} = 661.48\frac{BTU}{hr}$$

5.4.12 Net heat liberated:

$$Q \ = \ Q_{fuel} + Q_{air} + Q_{steam} + Q_R - Q_{wall} - Q_{exhaust}$$

$Neglecting\ Q_R\ in\ our\ case,$

$$Q \ = \ 14996.78\frac{BTU}{hr}$$

5.4.13 Number of tubes required:

$$N_{tubes} = \frac{Q}{2\pi r L q} \ ; where\ r = \frac{d_o}{2}$$

$$N_{tubes} = 25$$

5.4.14 Cold Plane Area:

$$A_{CP} = ctc \ \times \ L \times N$$

$$A_{CP} = 2205.2\ m^2$$

5.4.15 $'\boldsymbol{\alpha}'$ calculations:

Total $\boldsymbol{\alpha}$ for single row, refractory backed surface is obtained through the graph,

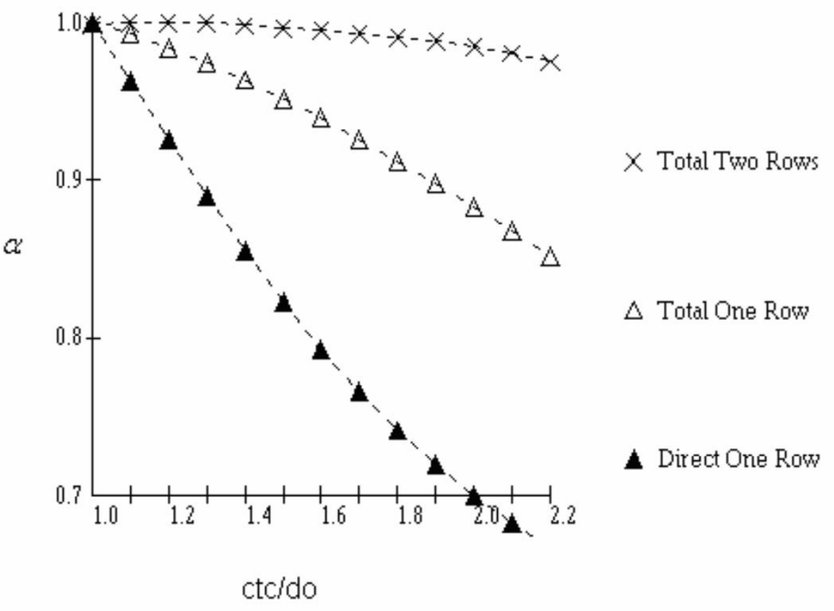


Figure 5-3: Calculating α from graph

Total $\boldsymbol{\alpha}$ for a single row is found out to be 0.97

$\alpha = 0.97$

5.4.16 Calculations for surfaces:

Assuming r = 3 and h = 5,

i. Top Surface Area:

$A_{top} = \pi r^2$

$A_{top} = 9\ \pi m^2$

ii. **Bottom Surface Area:**

$A_{bottom} = \pi r^2$

$A_{bottom} = 9\ \pi m^2$

iii. **Side Surface Area:**

$A_{side} = 2\pi r h$

$A_{side} = 30\ \pi m^2$

iv. **Total Surface Area:**

$A_{total} = 2\pi(3)^2 + 2\pi(3)(5)$

$A_{total} = 48\ \pi m^2\ \ \boldsymbol{or}\ \ 150.8\ m^2$

5.4.17 Mean length of radiant beams:

The dimensions of length , width and height should be 3:2:1

For cylindrical vertical heaters if the length over diameter ratio is greater than or equals to 2, then the mean beam length is equals to diameter.

Then,

$Beam\ length = 12.47\ ft$

5.4.18 Effective refracting surface:

$A_R = A_T - \alpha A_{CP}$

$A_R = 1988.24\ m^2$

CHAPTER 6

SIMULATION

Aspen HYSYS simulation software was used as the simulation software to simulate the whole process. As we know that the world is shifting towards Artificial Intelligence, so the need for simulation software has become significantly important.

6.1 Advantages of Aspen HYSYS:

Aspen HYSYS simulation software is one of the best software for the modeling and simulation of process industries because of the following reasons:

Prediction and Analysis:

It allows for the creation of virtual models that mimic real- world systems or processes. By running simulations, we can predict and analyze the behavior of these systems before implementing them in the physical world.

Cost and Time Savings:

Simulating complex systems or processes can save significant costs and time compared to physical prototyping and testing.

Design and Optimization:

It enables engineers and designers to create virtual prototypes and test different design iterations. This iterative process helps in optimizing designs and improving product quality.

Performance Evaluation:

It allows us to evaluate and compare the performance of different systems or processes under various conditions. It can help optimize manufacturing processes.

Visualization and Communication:

It often provides visual representations of complex data and models, making it easier to understand and communicate findings.

6.2 Simulation Flowsheet:

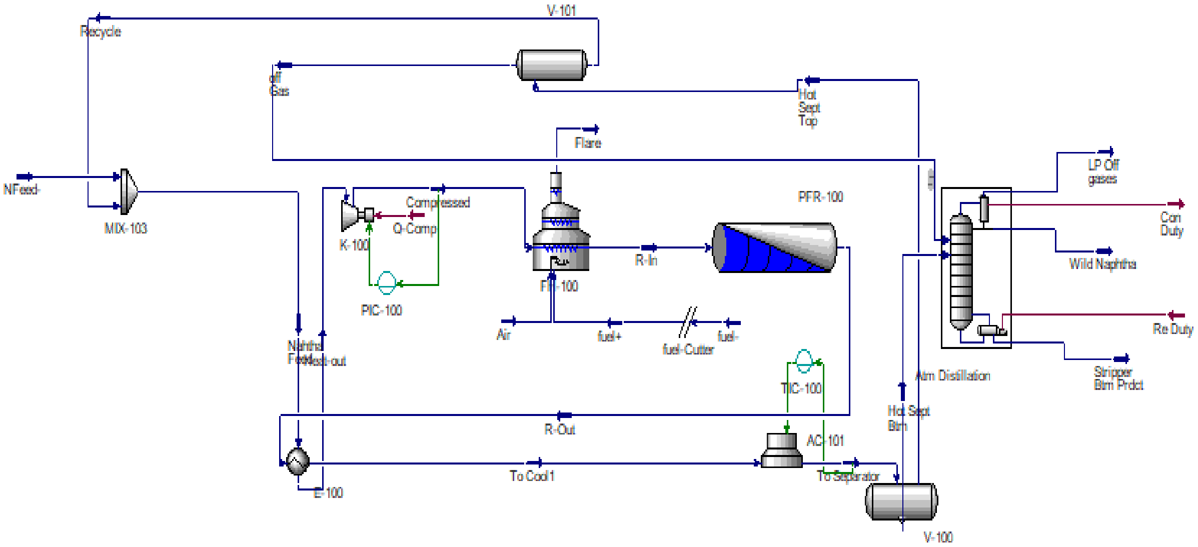


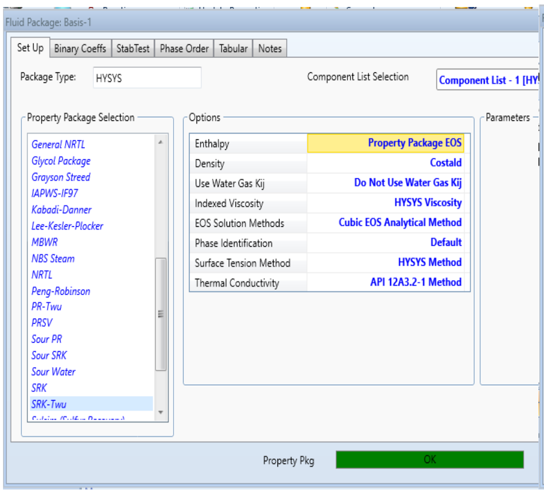
Figure 6-1: Simulation flow sheet

6.3 Property Package:

In Aspen PLUS simulation software, property package is termed as Methods which can be found in the Properties tab under the Components. Property packages are added to Aspen Plus to provide accurate thermodynamic and physical property data necessary for modeling, simulating, and optimizing chemical processes. They enable users to obtain realistic simulation results, optimize process performance, and ensure compatibility with experimental data and industry standards.

There are two property packages used:

1. Peng-Robinson, as it provides relatively accurate representation for the hydrocarbons in naphtha feed. It is the most widely used equation of state in refining processes.
2. Soave-Redich-Kwong (SRK), used for the sour gases, polar compounds and compounds containing sulphur, nitrogen and oxygen. The TWU modification further enhances the accuracy of the SRK equation of state.



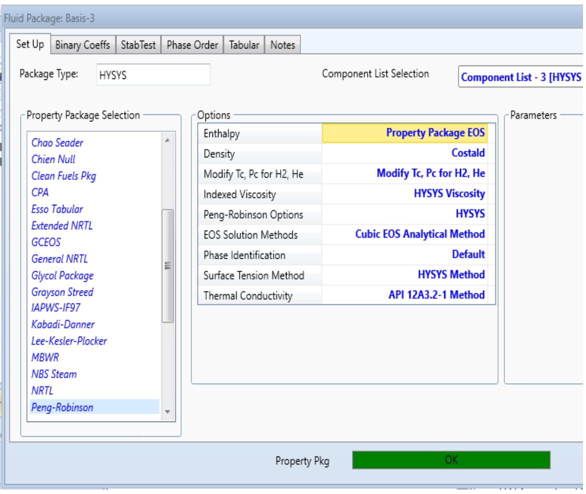
Package Type: HYSYS

Component List Selection: Component List - 1 [HY

Options	
Enthalpy	Property Package EOS
Density	Costald
Use Water Gas Kij	Do Not Use Water Gas Kij
Indexed Viscosity	HYSYS Viscosity
EOS Solution Methods	Cubic EOS Analytical Method
Phase Identification	Default
Surface Tension Method	HYSYS Method
Thermal Conductivity	API 12A3.2-1 Method

Property Pkg: OK

Figure 6-2: Property package on HYSYS



Package Type: HYSYS

Component List Selection: Component List - 3 [HYSYS

Options	
Enthalpy	Property Package EOS
Density	Costald
Modify Tc, Pc for H2, He	Modify Tc, Pc for H2, He
Indexed Viscosity	HYSYS Viscosity
Peng-Robinson Options	HYSYS
EOS Solution Methods	Cubic EOS Analytical Method
Phase Identification	Default
Surface Tension Method	HYSYS Method
Thermal Conductivity	API 12A3.2-1 Method

Property Pkg: OK

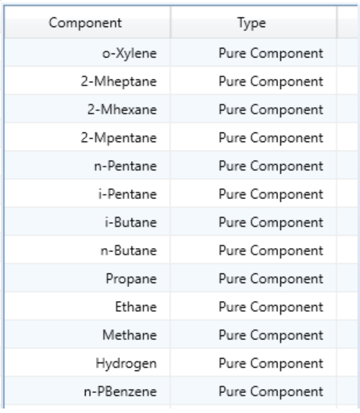
Figure 6-3: Property package on HYSYS

6.4 Feed Components:

Following are the components of naphtha feed used in Aspen HYSYS.

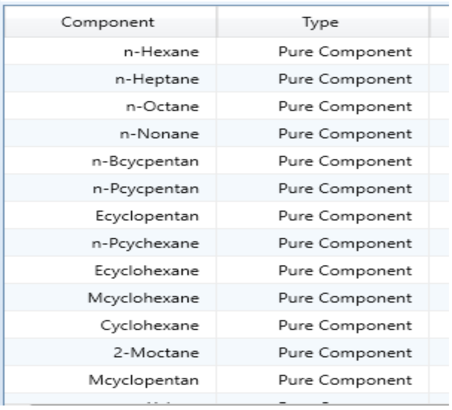
Component	Type
p-Xylene	Pure Component
m-Xylene	Pure Component
E-Benzene	Pure Component
Toluene	Pure Component
Benzene	Pure Component
Cyclopentane	Pure Component
E-Mercaptan	Pure Component
H2S	Pure Component
Thiophene	Pure Component
Ammonia	Pure Component
Pyrrole	Pure Component
Pyridine	Pure Component
Nitrogen	Pure Component

Figure 6-4: Feed Component Component



Component	Type
o-Xylene	Pure Component
2-Mheptane	Pure Component
2-Mhexane	Pure Component
2-Mpentane	Pure Component
n-Pentane	Pure Component
i-Pentane	Pure Component
i-Butane	Pure Component
n-Butane	Pure Component
Propane	Pure Component
Ethane	Pure Component
Methane	Pure Component
Hydrogen	Pure Component
n-PBenzene	Pure Component

Figure 6-5 : Feed component



Component	Type
n-Hexane	Pure Component
n-Heptane	Pure Component
n-Octane	Pure Component
n-Nonane	Pure Component
n-Bcycpentan	Pure Component
n-Pcycpentan	Pure Component
Ecyclopentan	Pure Component
n-Pcychexane	Pure Component
Ecyclohexane	Pure Component
Mcyclohexane	Pure Component
Cyclohexane	Pure Component
2-Moctane	Pure Component
Mcyclopentan	Pure Component

Figure 6-6: Feed

To add components in Aspen HYSYS, following procedure must be adopted:

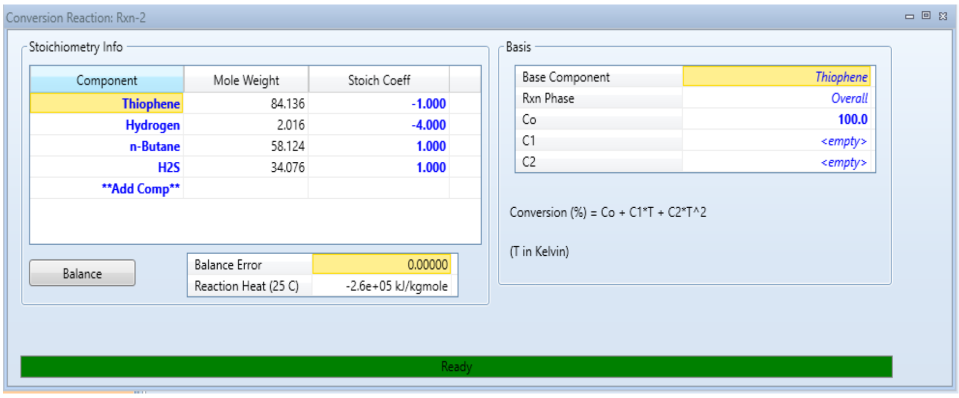
1. Open Aspen HYSYS Simulation Software.
2. Click on the Find tab below component list.
3. Type the name of the component and then click on the Find Now button.
4. Select the component you want to add.
5. After selecting the component, click on the Add Selected Compounds button.

6. Component is added to the Component List.

6.5 Reaction sets:

As the hydrotreater is responsible for the removal of Sulphur and Nitrogen, the major reactions that take place in the reactor for which the corresponding reaction sets are added by adding the stoichiometric coefficients.

6.5.1 Sulphur removal:



Component	Mole Weight	Stoich Coeff
Thiophene	84.136	-1.000
Hydrogen	2.016	-4.000
n-Butane	58.124	1.000
H2S	34.076	1.000
Add Comp		

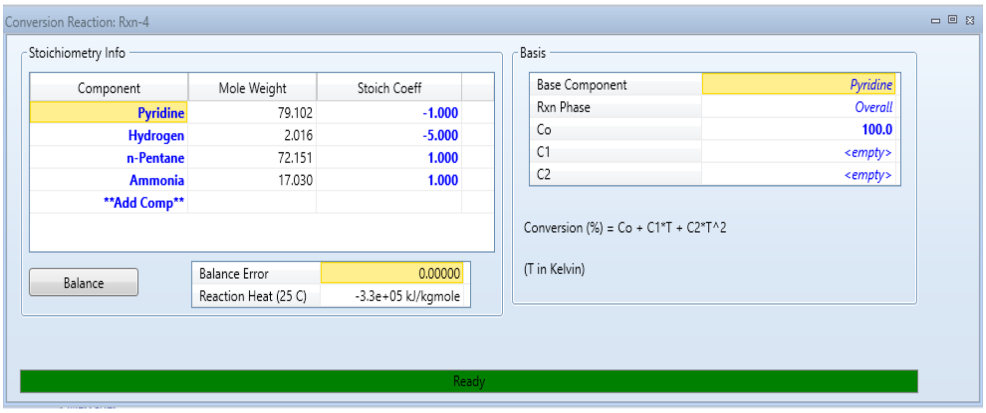
Balance Error	0.00000
Reaction Heat (25 C)	-2.6e+05 kJ/kgmole

Base Component	Thiophene
Rxn Phase	Overall
Co	100.0
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

(T in Kelvin)

Figure 6-7: Sulfur removal on HYSYS



Component	Mole Weight	Stoich Coeff
Pyridine	79.102	-1.000
Hydrogen	2.016	-5.000
n-Pentane	72.151	1.000
Ammonia	17.030	1.000
Add Comp		

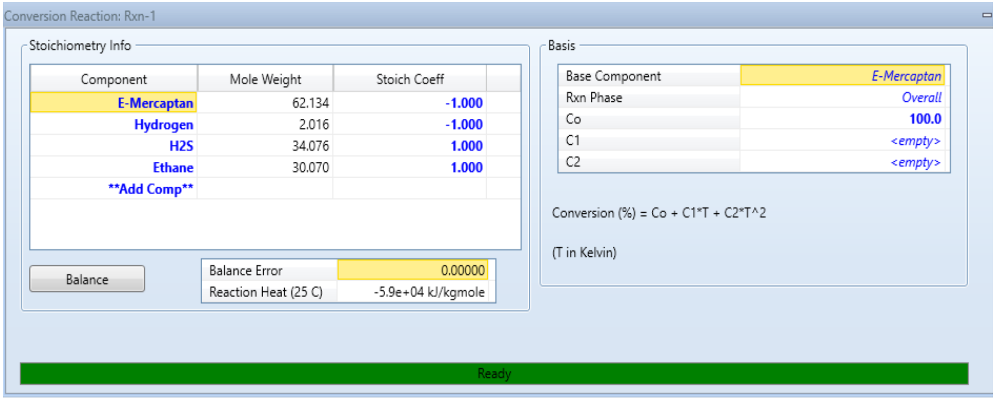
Balance Error	0.00000
Reaction Heat (25 C)	-3.3e+05 kJ/kgmole

Base Component	Pyridine
Rxn Phase	Overall
Co	100.0
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

(T in Kelvin)

Figure 6-8: Sulfur removal on HYSYS



Component	Mole Weight	Stoich Coeff
E-Mercaptan	62.134	-1.000
Hydrogen	2.016	-1.000
H2S	34.076	1.000
Ethane	30.070	1.000
Add Comp		

Balance Error	0.00000
Reaction Heat (25 C)	-5.9e+04 kJ/kgmole

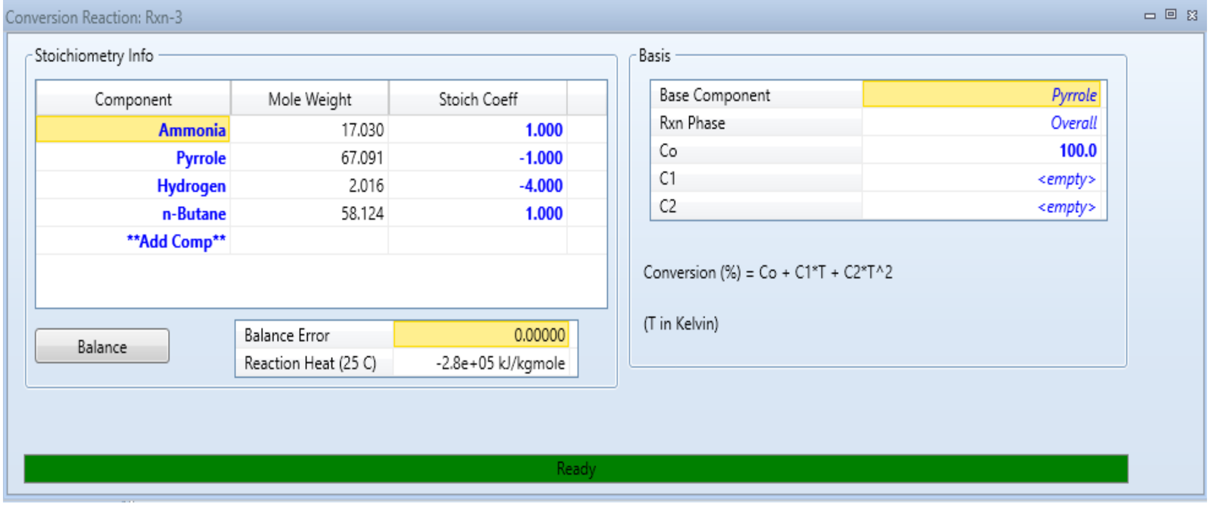
Base Component	E-Mercaptan
Rxn Phase	Overall
Co	100.0
C1	<empty>
C2	<empty>

Conversion (%) = Co + C1*T + C2*T^2

(T in Kelvin)

Figure 6-9: Stoichiometry on HYSYS

6.5.2 Nitrogen removal:



Component	Mole Weight	Stoich Coeff
Ammonia	17.030	1.000
Pyrrole	67.091	-1.000
Hydrogen	2.016	-4.000
n-Butane	58.124	1.000
Add Comp		

Balance Error	0.00000
Reaction Heat (25 C)	-2.8e+05 kJ/kgmole

Base Component	Pyrrole
Rxn Phase	Overall
Co	100.0
C1	<empty>
C2	<empty>

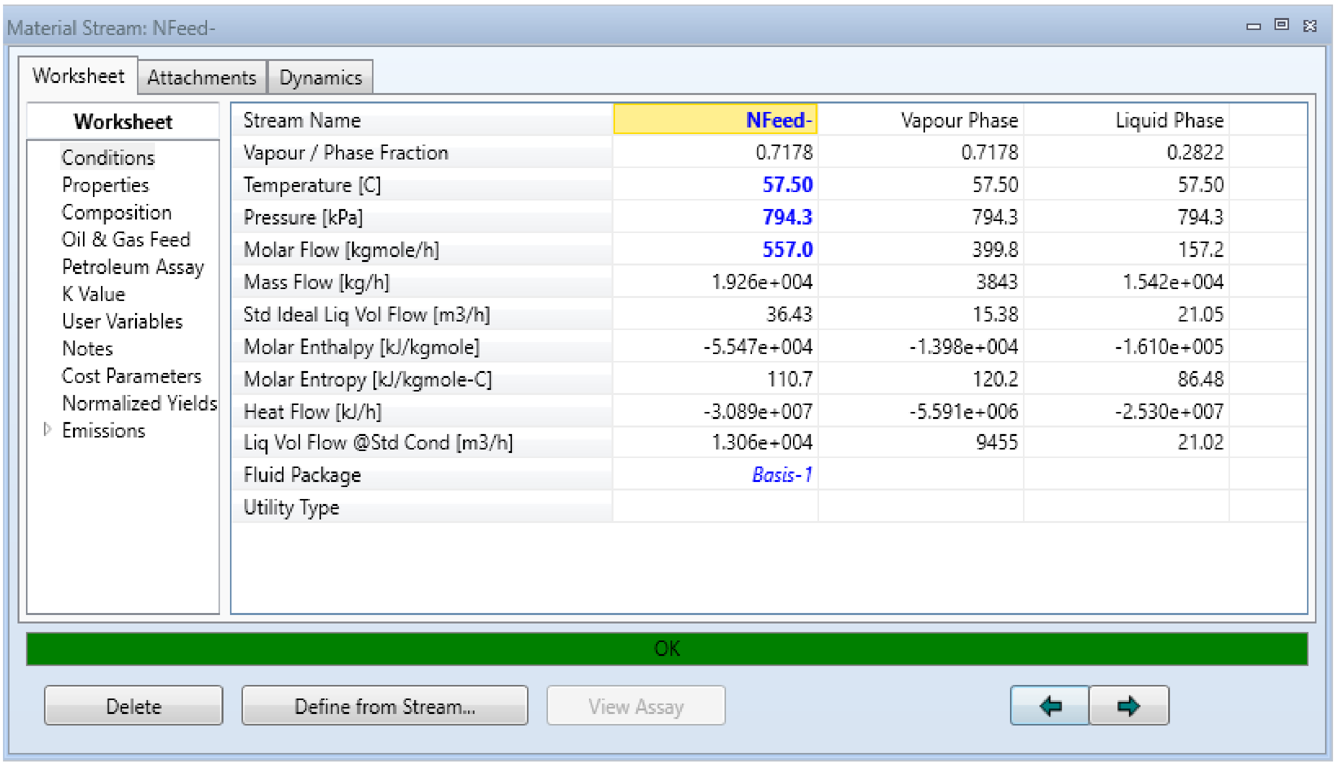
Conversion (%) = Co + C1*T + C2*T^2

(T in Kelvin)

Figure 6-10: Nitrogen removal on HYSYS

6.5.3 Inlet stream (Naphtha feed):

The naphtha feed coming from the crude distillation unit, after it's temperature is reduced, has the following conditions.

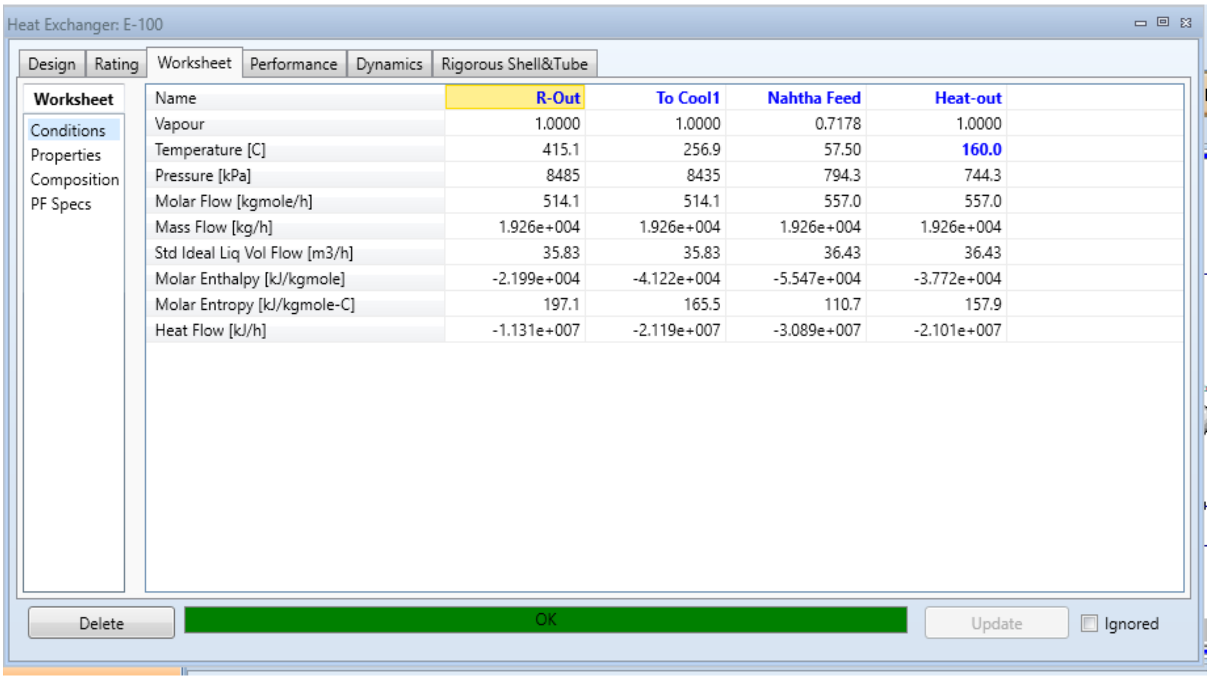


Stream Name	NFeed-	Vapour Phase	Liquid Phase
Vapour / Phase Fraction	0.7178	0.7178	0.2822
Temperature [C]	57.50	57.50	57.50
Pressure [kPa]	794.3	794.3	794.3
Molar Flow [kgmole/h]	557.0	399.8	157.2
Mass Flow [kg/h]	1.926e+004	3843	1.542e+004
Std Ideal Liq Vol Flow [m3/h]	36.43	15.38	21.05
Molar Enthalpy [kJ/kgmole]	-5.547e+004	-1.398e+004	-1.610e+005
Molar Entropy [kJ/kgmole-C]	110.7	120.2	86.48
Heat Flow [kJ/h]	-3.089e+007	-5.591e+006	-2.530e+007
Liq Vol Flow @Std Cond [m3/h]	1.306e+004	9455	21.02
Fluid Package	Basis-1		
Utility Type			

Figure 6-11: Inlet stream condition

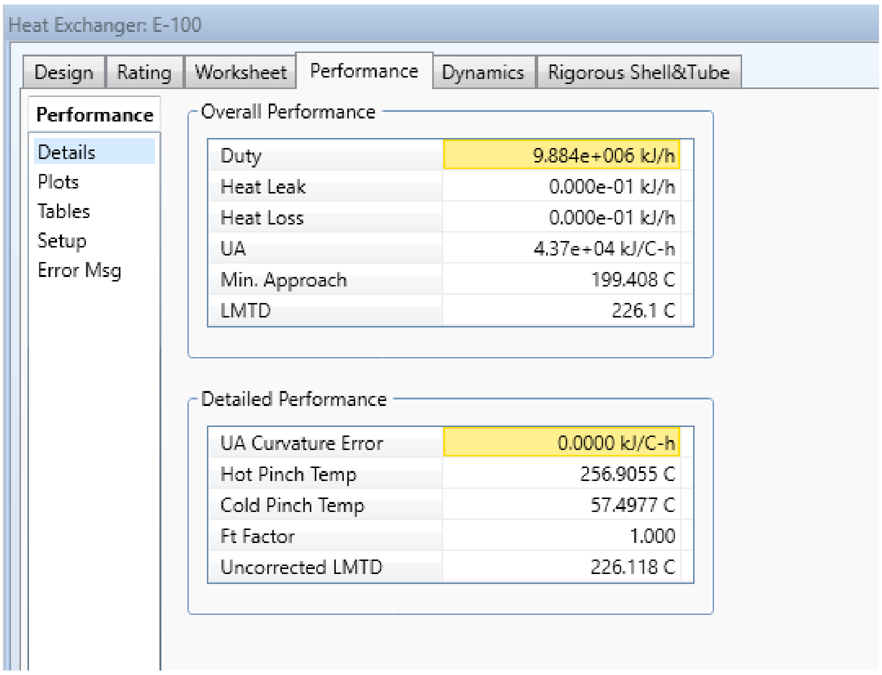
6.5.4 Heat exchanger:

Shell and tube type of heat exchanger is used to preheat the untreated naphtha feed, using hot treated naphtha feed coming from the reactor, before it enters into the hydrotreater.



Name	R-Out	To Cool1	Nahtha Feed	Heat-out
Vapour	1.0000	1.0000	0.7178	1.0000
Temperature [C]	415.1	256.9	57.50	160.0
Pressure [kPa]	8485	8435	794.3	744.3
Molar Flow [kgmole/h]	514.1	514.1	557.0	557.0
Mass Flow [kg/h]	1.926e+004	1.926e+004	1.926e+004	1.926e+004
Std Ideal Liq Vol Flow [m3/h]	35.83	35.83	36.43	36.43
Molar Enthalpy [kJ/kgmole]	-2.199e+004	-4.122e+004	-5.547e+004	-3.772e+004
Molar Entropy [kJ/kgmole-C]	197.1	165.5	110.7	157.9
Heat Flow [kJ/h]	-1.131e+007	-2.119e+007	-3.089e+007	-2.101e+007

Figure 6-12: Heat exchanger



Overall Performance

Duty	9.884e+006 kJ/h
Heat Leak	0.000e-01 kJ/h
Heat Loss	0.000e-01 kJ/h
UA	4.37e+04 kJ/C-h
Min. Approach	199.408 C
LMTD	226.1 C

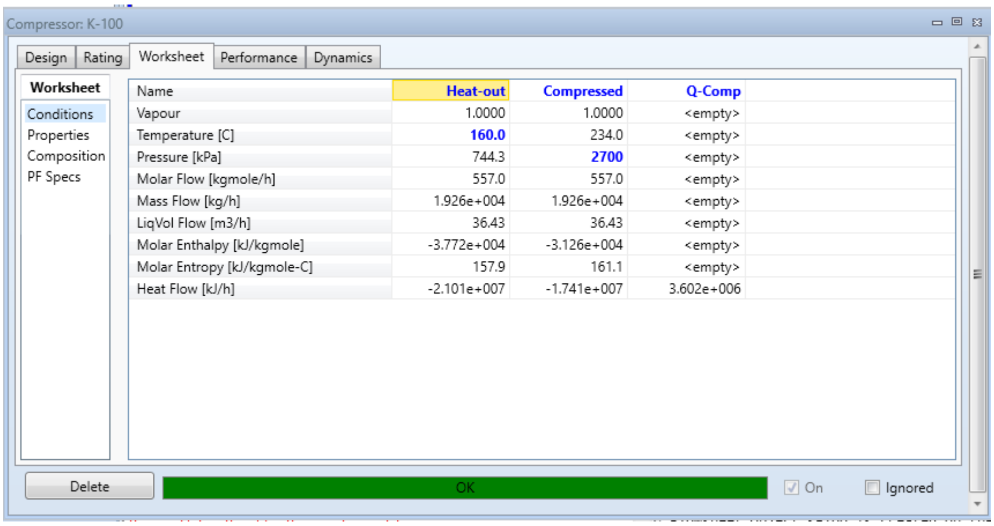
Detailed Performance

UA Curvature Error	0.0000 kJ/C-h
Hot Pinch Temp	256.9055 C
Cold Pinch Temp	57.4977 C
Ft Factor	1.000
Uncorrected LMTD	226.118 C

Figure 6-13: Heat exchanger

6.5.5 Compressor:

It is used to increase the pressure of the naphtha feed before it enters into hydrotreater as the reaction is favored at pressures between 20-35 bars.

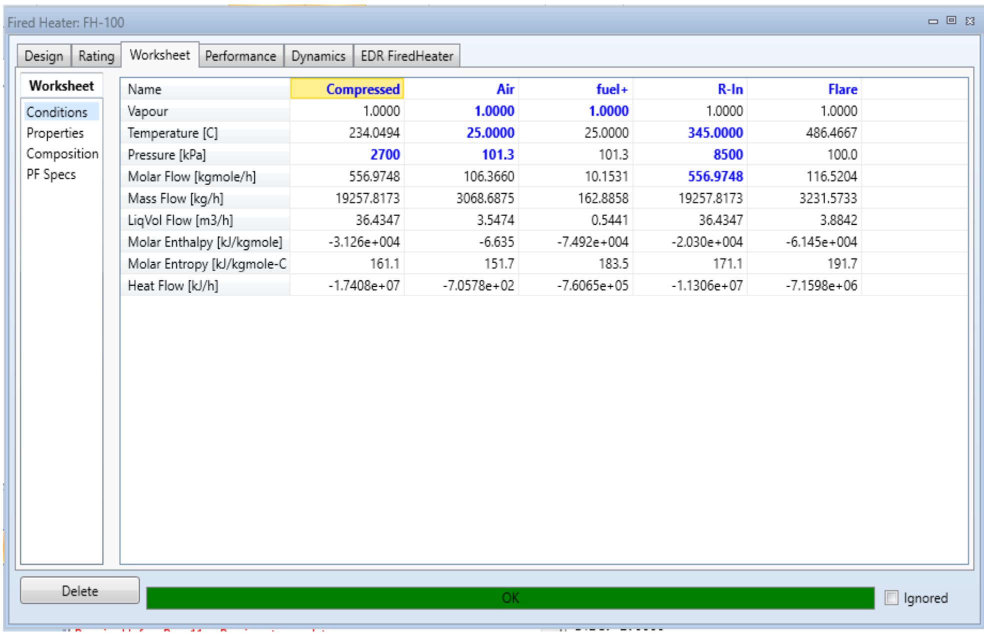


Name	Heat-out	Compressed	Q-Comp
Vapour	1.0000	1.0000	<empty>
Temperature [C]	160.0	234.0	<empty>
Pressure [kPa]	744.3	2700	<empty>
Molar Flow [kgmole/h]	557.0	557.0	<empty>
Mass Flow [kg/h]	1.926e+004	1.926e+004	<empty>
LiqVol Flow [m3/h]	36.43	36.43	<empty>
Molar Enthalpy [kJ/kgmole]	-3.772e+004	-3.126e+004	<empty>
Molar Entropy [kJ/kgmole-C]	157.9	161.1	<empty>
Heat Flow [kJ/h]	-2.101e+007	-1.741e+007	3.602e+006

Figure 6-14: Compressor simulation

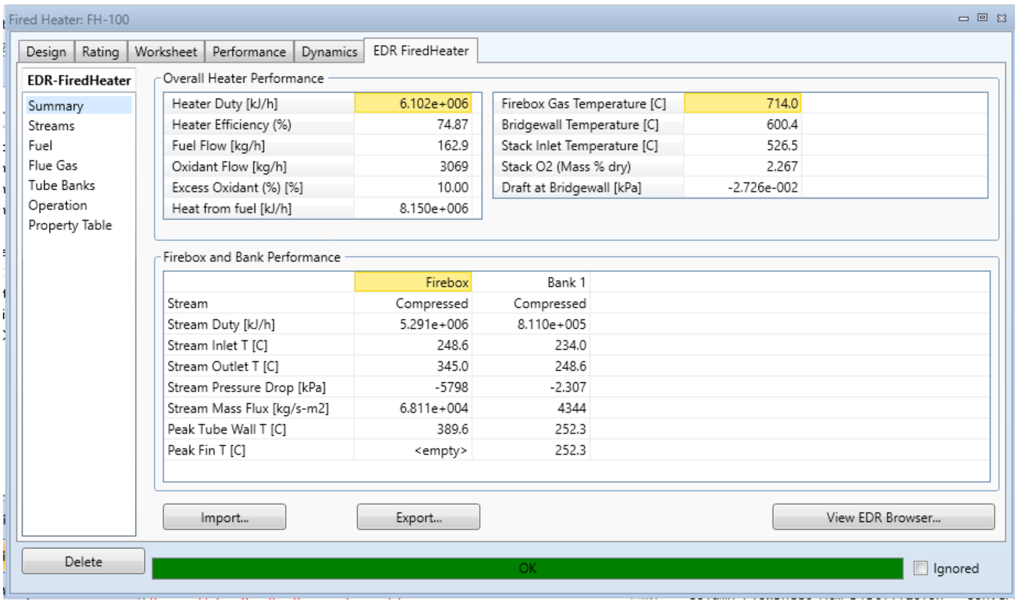
6.5.6 Fired heater:

A Direct fired type heater is used to heat the naphtha feed, methane gas is used as fuel and air from atmosphere is used for combustion.



Name	Compressed	Air	fuel+	R-In	Flare
Vapour	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [C]	234.0494	25.0000	25.0000	345.0000	486.4667
Pressure [kPa]	2700	101.3	101.3	8500	100.0
Molar Flow [kgmole/h]	556.9748	106.3660	10.1531	556.9748	116.5204
Mass Flow [kg/h]	19257.8173	3068.6875	162.8858	19257.8173	3231.5733
LiqVol Flow [m3/h]	36.4347	3.5474	0.5441	36.4347	3.8842
Molar Enthalpy [kJ/kgmole]	-3.126e+004	-6.635	-7.492e+004	-2.030e+004	-6.145e+004
Molar Entropy [kJ/kgmole-C	161.1	151.7	183.5	171.1	191.7
Heat Flow [kJ/h]	-1.7408e+07	-7.0578e+02	-7.6065e+05	-1.1306e+07	-7.1598e+06

Figure 6-15: Fired heater simulation



Overall Heater Performance

Heater Duty [kJ/h]	6.102e+006
Heater Efficiency (%)	74.87
Fuel Flow [kg/h]	162.9
Oxidant Flow [kg/h]	3069
Excess Oxidant (%) [%]	10.00
Heat from fuel [kJ/h]	8.150e+006

Firebox Gas Temperature [C]	714.0
Bridgewall Temperature [C]	600.4
Stack Inlet Temperature [C]	526.5
Stack O2 (Mass % dry)	2.267
Draft at Bridgewall [kPa]	-2.726e-002

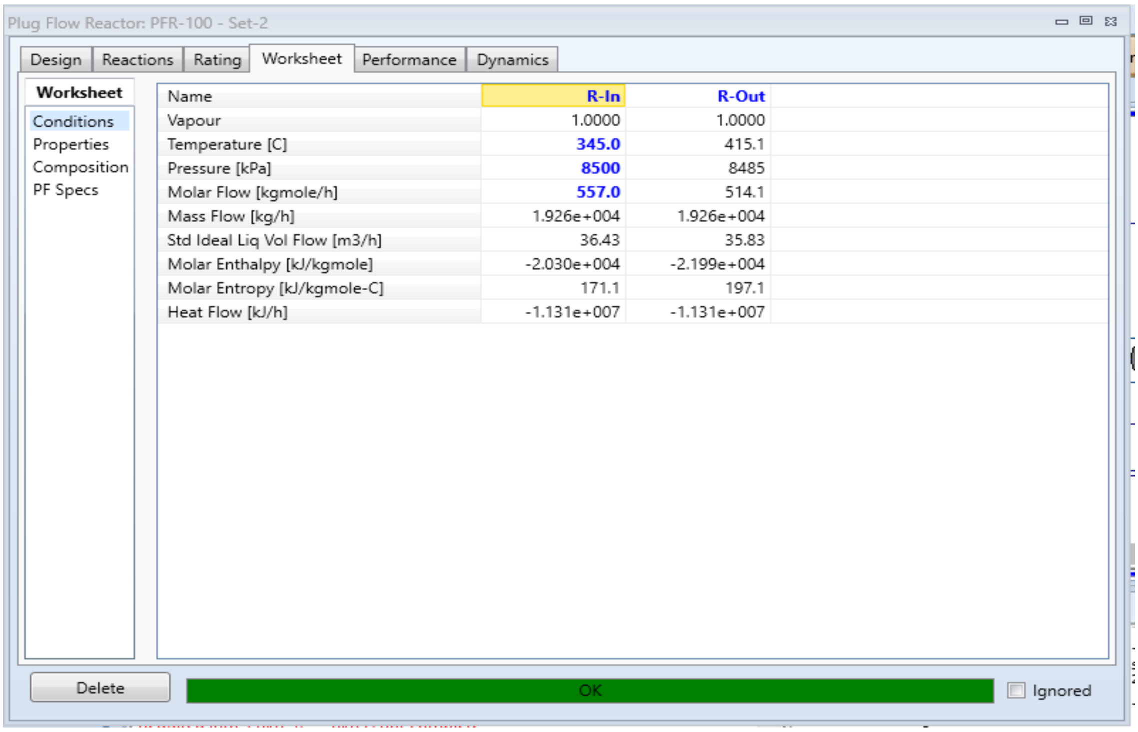
Firebox and Bank Performance

	Firebox	Bank 1
Stream	Compressed	Compressed
Stream Duty [kJ/h]	5.291e+006	8.110e+005
Stream Inlet T [C]	248.6	234.0
Stream Outlet T [C]	345.0	248.6
Stream Pressure Drop [kPa]	-5798	-2.307
Stream Mass Flux [kg/s-m2]	6.811e+004	4344
Peak Tube Wall T [C]	389.6	252.3
Peak Fin T [C]	<empty>	252.3

Figure 6-16: Fired heater simulation

6.5.7 Reactor (Hydrotreater):

The reaction sets are added in the reactor specification.

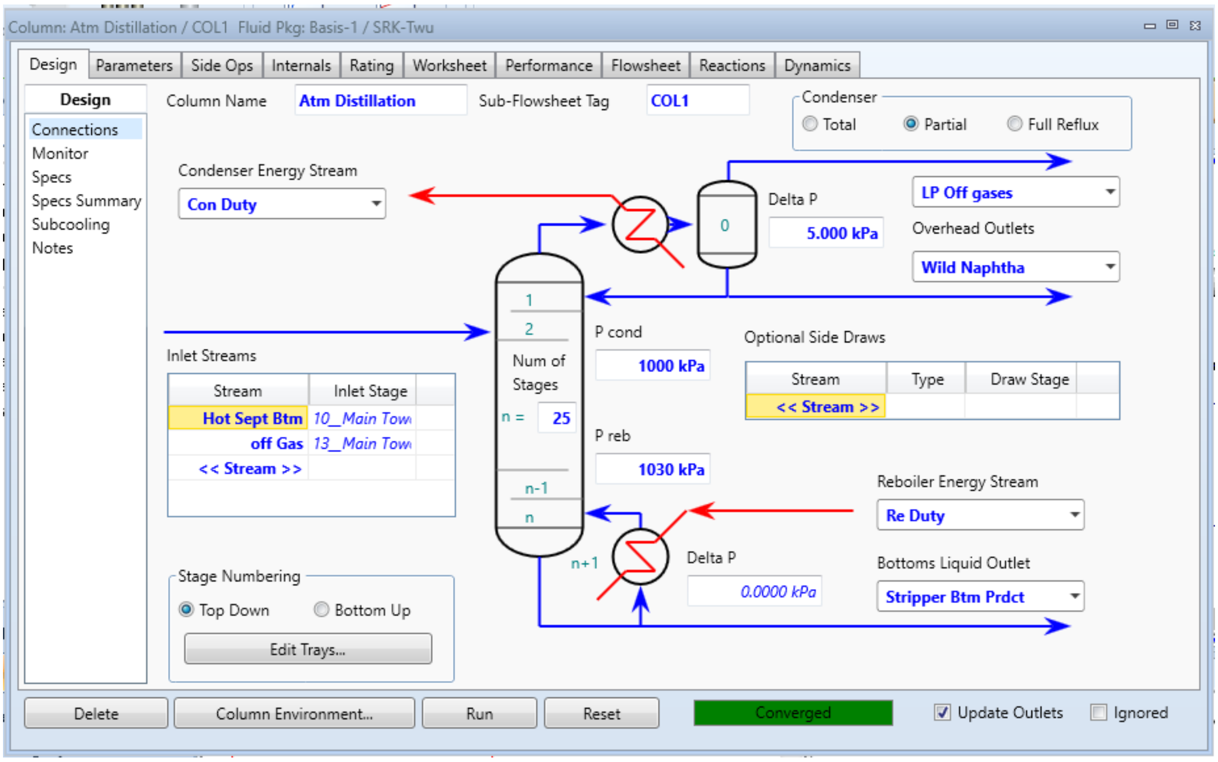


Name	R-In	R-Out
Vapour	1.0000	1.0000
Temperature [C]	345.0	415.1
Pressure [kPa]	8500	8485
Molar Flow [kgmole/h]	557.0	514.1
Mass Flow [kg/h]	1.926e+004	1.926e+004
Std Ideal Liq Vol Flow [m3/h]	36.43	35.83
Molar Enthalpy [kJ/kgmole]	-2.030e+004	-2.199e+004
Molar Entropy [kJ/kgmole-C]	171.1	197.1
Heat Flow [kJ/h]	-1.131e+007	-1.131e+007

Figure 6-17: Reactor simulation

6.5.8 Distillation Column (Stripper):

The stripper section is used to strip of the sour gases along with the separation of the light naphtha from heavy naphtha. The light naphtha is sent to the isomerization unit and heavy naphtha is sent to the reformer.


Column: Atm Distillation / COL1 Fluid Pkg: Basis-1 / SRK-Twu

Column Name: Atm Distillation
Sub-Flowsheet Tag: COL1
Condenser: ○ Total ● Partial ○ Full Reflux
Condenser Energy Stream: Con Duty
Delta P: 5.000 kPa
Overhead Outlets: LP Off gases; Wild Naphtha
P cond: 1000 kPa
Num of Stages n = 25
P reb: 1030 kPa
Reboiler Energy Stream: Re Duty
Delta P: 0.0000 kPa
Bottoms Liquid Outlet: Stripper Btm Prdct
Stage Numbering: ● Top Down ○ Bottom Up
☑ Update Outlets ☐ Ignored

Inlet Streams

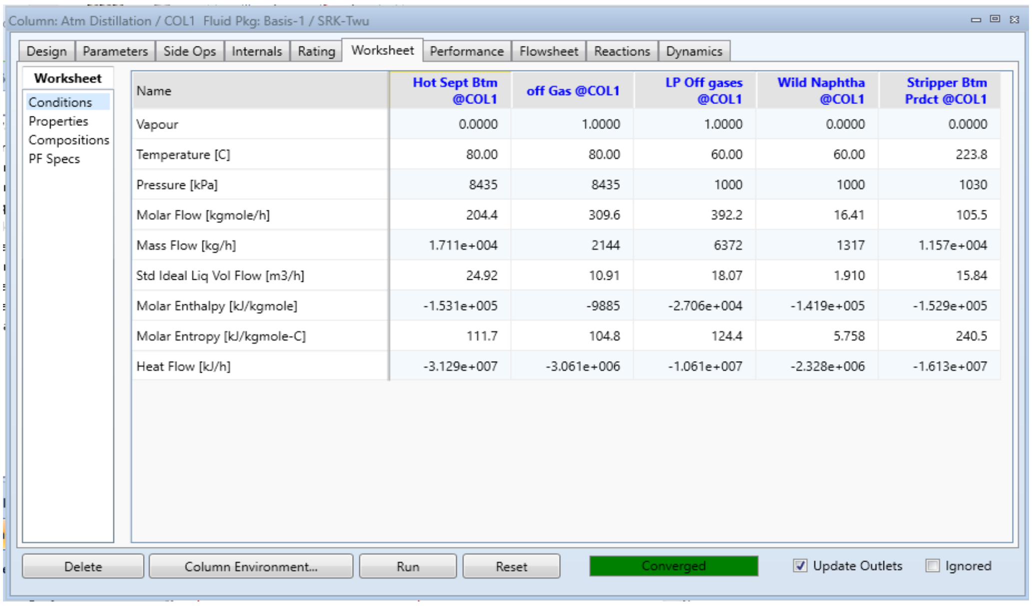
Stream	Inlet Stage
Hot Sept Btm	10__Main Tow
off Gas	13__Main Tow
<< Stream >>	

Optional Side Draws

Stream	Type	Draw Stage
<< Stream >>		

Figure 6-18: Distillation column simulation

The following results were obtained:


Column: Atm Distillation / COL1 Fluid Pkg: Basis-1 / SRK-Twu

Name	Hot Sept Btm @COL1	off Gas @COL1	LP Off gases @COL1	Wild Naphtha @COL1	Stripper Btm Prdct @COL1
Vapour	0.0000	1.0000	1.0000	0.0000	0.0000
Temperature [C]	80.00	80.00	60.00	60.00	223.8
Pressure [kPa]	8435	8435	1000	1000	1030
Molar Flow [kgmole/h]	204.4	309.6	392.2	16.41	105.5
Mass Flow [kg/h]	1.711e+004	2144	6372	1317	1.157e+004
Std Ideal Liq Vol Flow [m3/h]	24.92	10.91	18.07	1.910	15.84
Molar Enthalpy [kJ/kgmole]	-1.531e+005	-9885	-2.706e+004	-1.419e+005	-1.529e+005
Molar Entropy [kJ/kgmole-C]	111.7	104.8	124.4	5.758	240.5
Heat Flow [kJ/h]	-3.129e+007	-3.061e+006	-1.061e+007	-2.328e+006	-1.613e+007

☑ Update Outlets ☐ Ignored

Figure 6-19: Distillation column results.

CHAPTER 7

INSTRUMENTATION

Proportional, integral and derivative control loops are common types of feedback control systems which are used to regulate and control various parameters in a process

Two control loops are used in this simulation

1. Pressure control loop on the compressor.
2. Temperature control loop on air cooler.

7.1 Pressure control loop:

The pressure of the incoming naphtha feed from the crude distillation units may vary, the pressure control loop ensures that the inlet feed to the reactor is maintained at 27 bars as this pressure is optimum for the naphtha hydrotreating.

A set point of 27 bars is provided, pressure sensors sense any change in pressure and give feedback to the compressor which then changes the rpm accordingly to achieve the given set point.

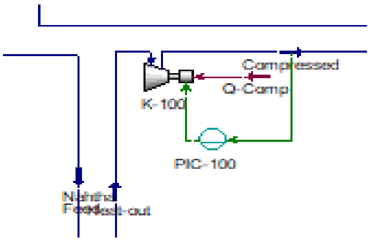


Figure 7-1 Pressure controller

Following results were obtained.

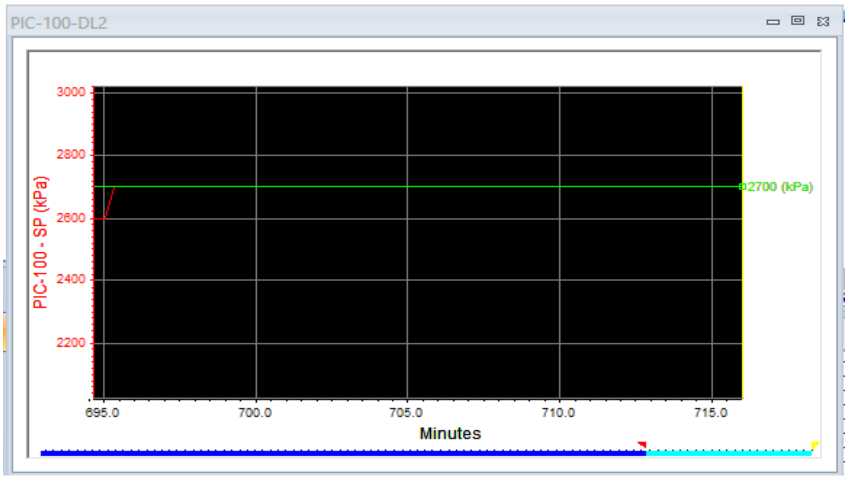


Figure 7-2: Result of pressure controller.

7.2 Temperature control loop:

The temperature of the naphtha feed is to be reduced as it enters the air cooler, this is done by installing a temperature control loop. Temperature sensors are present at the outlet of the air cooler and a set point of 25 degree celsius is given. Any deviations in the temperature are recorded and the temperature is maintained at the set point by adjusting the rpm of the fans.

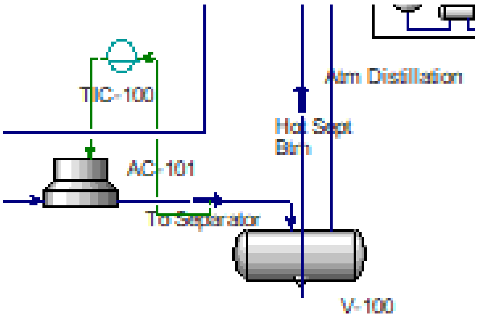


Figure 7-3: Temperature controller on air cooler

The following result was obtained:

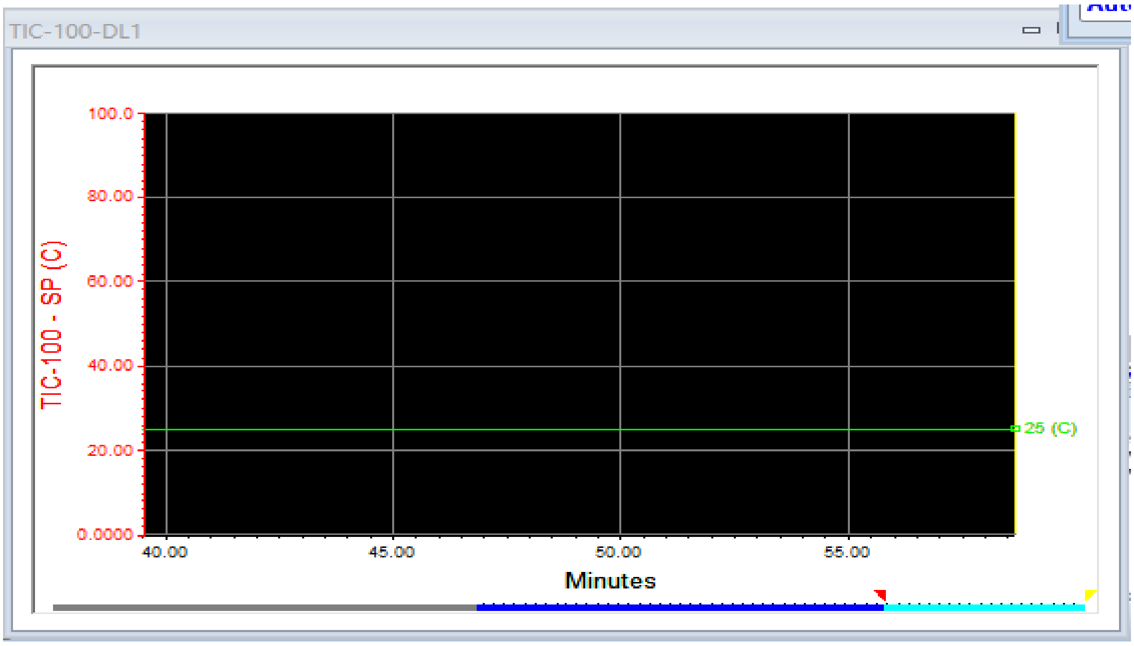


Figure 7-4: Result of temperature controller

CHAPTER 8

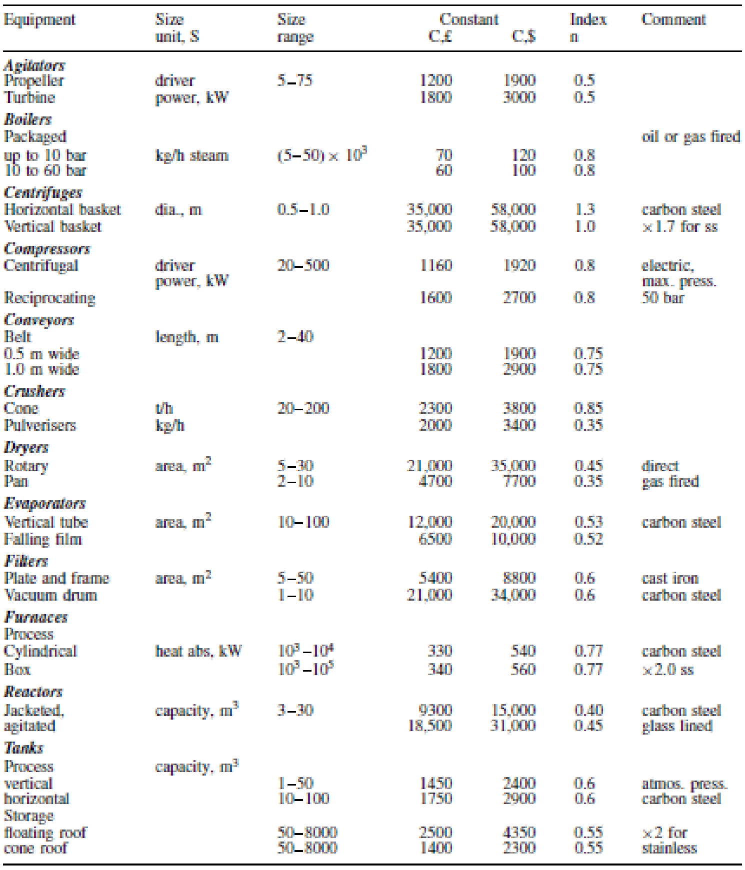
ECONOMIC ANALYSIS

Cost analysis is, at its core, the evaluation and cataloging of every project-related expense. The main objective of cost analysis is to assess the financial viability and profitability of a project. Cost analysis provides the foundation for financial planning, decision-making, and risk management. The main objective of any plant is to generate revenue; thus a comprehensive economic analysis must be completed before the installation of any new ones.

The economic analysis for our project was conducted using the Coulson and Richardson method.

The project's total cost is determined by adding fixed capital expenses to the total cost of production.

8.1 Purchased Cost of Equipment



Equipment	Size unit, S	Size range	Constant		Index n	Comment
			C,£	C,$		
Agitators						
Propeller	driver power, kW	5–75	1200	1900	0.5	
Turbine			1800	3000	0.5	
Boilers						
Packaged						oil or gas fired
up to 10 bar	kg/h steam	$(5-50) \times 10^3$	70	120	0.8	
10 to 60 bar			60	100	0.8	
Centrifuges						
Horizontal basket	dia., m	0.5–1.0	35,000	58,000	1.3	carbon steel
Vertical basket			35,000	58,000	1.0	×1.7 for ss
Compressors						
Centrifugal	driver power, kW	20–500	1160	1920	0.8	electric, max. press.
Reciprocating			1600	2700	0.8	50 bar
Conveyors						
Belt	length, m	2–40				
0.5 m wide			1200	1900	0.75	
1.0 m wide			1800	2900	0.75	
Crushers						
Cone	t/h	20–200	2300	3800	0.85	
Pulverisers	kg/h		2000	3400	0.35	
Dryers						
Rotary	area, m^2	5–30	21,000	35,000	0.45	direct
Pan		2–10	4700	7700	0.35	gas fired
Evaporators						
Vertical tube	area, m^2	10–100	12,000	20,000	0.53	carbon steel
Falling film			6500	10,000	0.52	
Filters						
Plate and frame	area, m^2	5–50	5400	8800	0.6	cast iron
Vacuum drum		1–10	21,000	34,000	0.6	carbon steel
Furnaces						
Process						
Cylindrical	heat abs, kW	10^3-10^4	330	540	0.77	carbon steel
Box		10^3-10^5	340	560	0.77	×2.0 ss
Reactors						
Jacketed, agitated	capacity, m^3	3–30	9300	15,000	0.40	carbon steel
			18,500	31,000	0.45	glass lined
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 8-1: PCEs of different equipments

8.1.1 Surge Drum

Volume of the Drum = 50 m^3

Purchase Cost of the Drum in £ = $Cost\ Constant\ x\ Characteristic\ Size\ Parameter^{Index}$

Purchase Cost of the Drum in £ = $1450\ x\ 50^{0.6}$ = 15,161.8 £

8.1.2 Suction Drum

Volume of the Drum = 40 m^3

Purchase Cost of the Drum in £ = $Cost\ Constant\ x\ Characteristic\ Size\ Parameter^{Index}$

Purchase Cost of the Drum in £ = $1450\ x\ 40^{0.6}$ = 13,261.8 £

8.1.3 Charged Heater

Absolute Heat of Heater in KW = 1,805

Purchase Cost of the Furnace/ Heater in £:

$$Purchase\ cost\ of\ the\ furnace\ in\ £ =\ Cost\ Constant\ x\ Characteristic\ Size\ Parameter^{Index}$$

Purchase Cost of the 2 - Furnace/ Heater in £ = 2* ($340\ x\ 1805^{0.77}$)= 2,18,774.3 £

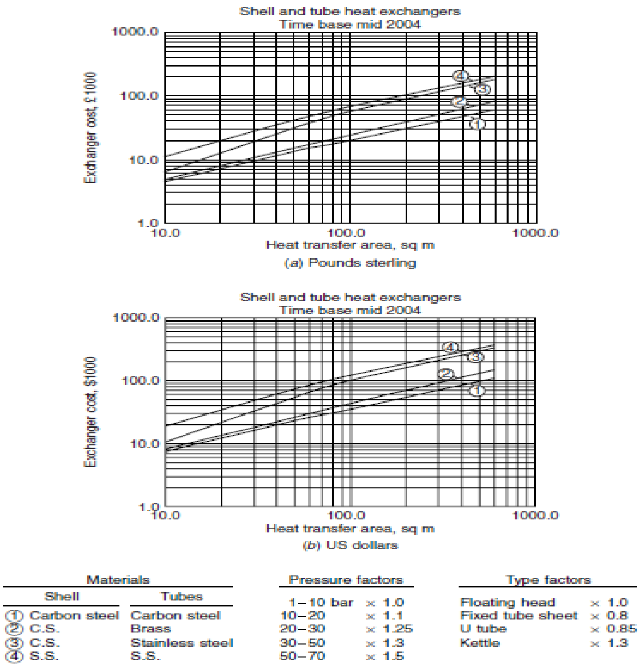


Figure 6.3*a*, *b*. Shell and tube heat exchangers. Time base mid-2004
Purchased cost = (bare cost from figure) × Type factor × Pressure factor

Figure 8-2: Cost of heat exchanger

8.1.4 Heat Exchanger

Heat Transfer Area -1 = 60 m^2

Heat Transfer Area -2 = 60 m^2

Type of Heat Exchanger = Shell and Tube (Fixed Tube Sheet)

Material of Shell and Tubes = Carbon Steel (Shell), Stainless Steel (Tubes)

Pressure Factor = 1.0

Using the same graph, Heat Exchanger cost is calculated,

$$Purchased\ Cost\ = bare\ cost\ from\ graph\ \times Pressure\ Factor\ \times Type\ Factor$$

$$\text{Purchased Cost of Heat Exchanger} - 1 = 39{,}000 \times 1.0 \times 0.8 = 31{,}200\ £$$

$$\text{Purchased Cost of Heat Exchanger} - 2 = 39{,}000 \times 1.0 \times 0.8 = 31{,}200\ £$$

$$\text{Total Purchased Cost of Condensers} = 62{,}400\ £$$

8.1.5 Reactor

Volume of the Reactor = 158.23 m^3

Purchase Cost of the Reactor in £ :

$$Purchase\ Cost\ of\ the\ reactor\ in\ £\ = Cost\ Constant\ x\ Characteristic\ Size\ Parameter^{Index}$$

Purchase Cost of the Reactor in £ = 13,950 x $158.23^{0.8}$ = 8,01,687.8 £

Cost of Catalyst in £/ m^3 = 200

Volume of Catalyst in m^3 = 64.64

Cost of Catalyst in £ = 18,988

Total cost of Reactor in £ = 8,20,675.8

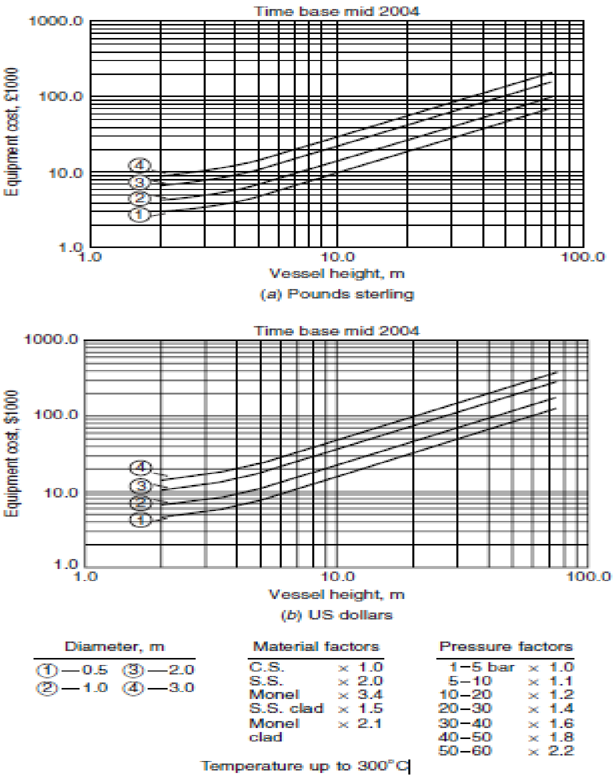


Figure 6.5*a*, *b*. Vertical pressure vessels. Time base mid-2004.
Purchased cost = (bare cost from figure) × Material factor × Pressure factor

Figure 8-3: Cost of reactor

8.1.6 Separators

Height of the Column = 4 m

Diameter of the column = 1 m

Material Used = Stainless Steel

Purchased Cost = bare cost from graph × Pressure Factor × Material Factor

Purchased Cost = 8,000 × 1.4 × 2 = 22,400 £

Purchased Cost of 2 Separators = 44,800 £

8.1.7 Condensers (Fin Fan)

Purchase Cost of the Condenser from Literature at respective conditions in £ = 70,000 £

Total Purchased Cost of 2 Condensers = 1,40,000 £

8.1.8 Compressor

Power in KW = 1415

Purchase Cost in £ = $Cost\ Constant\ x\ Characteristic\ Size\ Parameter^{Index}$

Purchase Cost of the Compressor in £ = $2320\ x\ 1415^{0.77}$ = 6,188,28.7 £

8.1.9 Distillation Column

Diameter of the columns = 1.2 m

Type of the column = Plates Used

Material Factor = 1.7 m

$$\boldsymbol{Purchased\ Cost\ = bare\ cost\ from\ graph \times Pressure\ Factor \times Material\ Factor}$$

Column Cost = $20{,}000 \times\ 1.4\ \times\ 2 = 56{,}000$ £

Column Plate Cost in £ = 1,700

$$\boldsymbol{Total\ Cost\ of\ Column = Column\ Cost + Column\ Plate\ Cost}$$

$$\text{Total Column Cost} = 57{,}700\ £$$

Total purchased cost of major equipment items (PCE)

Table 17: PCEs of major equipments.

Equipment	**Purchased Cost of Equipment**
Surge Drum	15,161.8 £
Suction Drum	13,261,8 £
Charger Heater	2,18,774.3 £
Heat Exchanger	62,400 £
Reactor	8,20,675.8 £
Separators	44,800 £
Condenser	1,40,000 £
Compressor	6,18,828.7 £
Distillation Column	57,700 £
Total PCE in 2004	**1,991,602.4 £**

8.2 PCE in 2024

Cost of inflation index in 2004 = 444.2

Cost of inflation index in 2024 = 795.1

Total PEC in 2004 = 1,991,602.4 £

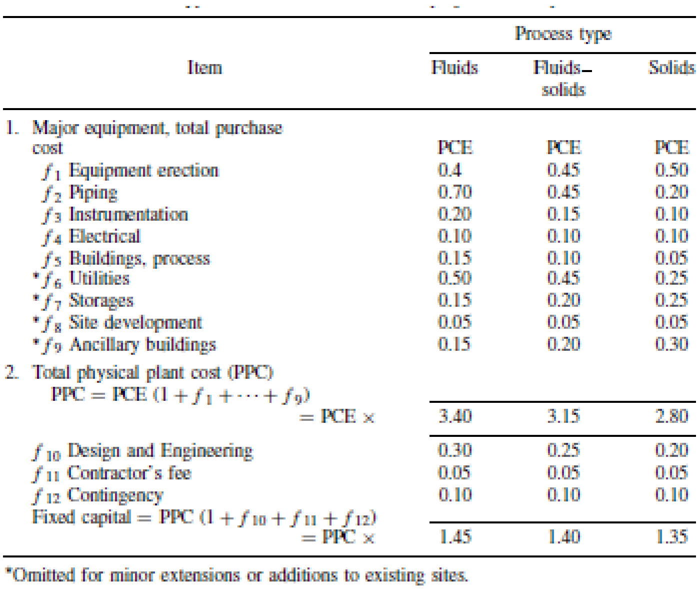
Following formula is used to find total PEC in 2024,

$$\boldsymbol{Total\ PEC\ in\ 2024 = Total\ PEC\ in\ 2004 \times \frac{Cost\ \ index\ in\ 2024}{Cost\ \ index\ in\ 2004}}$$

$$\text{Total PEC in 2024} = 1991602.4\ \times \frac{795.1}{444.2} = 3{,}564{,}888\ £$$

8.3 Estimation of Fixed Capital Cost (FCC)

Typical Factors for Estimation of FCC



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 ×	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 ×	1.45	1.40	1.35

*Omitted for minor extensions or additions to existing sites.

Figure 8-4: Factors for FCC

Since no solids are involved in our operation, fluids have been chosen as the process type.

The following formula is used to determine the physical plant cost:

$$\boldsymbol{PCC = PEC\ (1 +\ f_1\ + \cdots +\ f_9\,)}$$

$$\text{PCC} = 3{,}564{,}888 \text{ x } (1 +\ 0.4 + 0.7 + 0.2 + 0.1 + 0.5) =\ 1{,}033{,}817\ £$$

Fixed Capital is calculated as,

$$Fixed\ Capital\ Cost = PCC\ (1 + f_{10} + f_{11} + f_{12})$$

Fixed Capital Cost = 1,033,817 x (1 + 0.3 + 0.0 + 0.1) = 1,447,344 £

An estimate of the working capital is 5% of the fixed capital.

$$Working\ Capital\ Cost = 0.5 \times Fixed\ Capital\ Cost$$

$$\text{Working Capital Cost} = 0.5 \times 1,447,344 = 7,23,672\ £$$

The entire amount of investment needed is now determined as,

$$Total\ Investment\ Required = Fixed\ Capital + Working\ Capital$$

$$\text{Total Investment Required} = 1,447,344\ £ + 7,23,672\ £ = 1,519,711\ £$$

8.4 Annual Operating Cost

Plant attainment = 0.95

Operating time of Plant = $365 \times 0.95 = 347\ \frac{days}{year}$

8.5 Variable Operating Cost

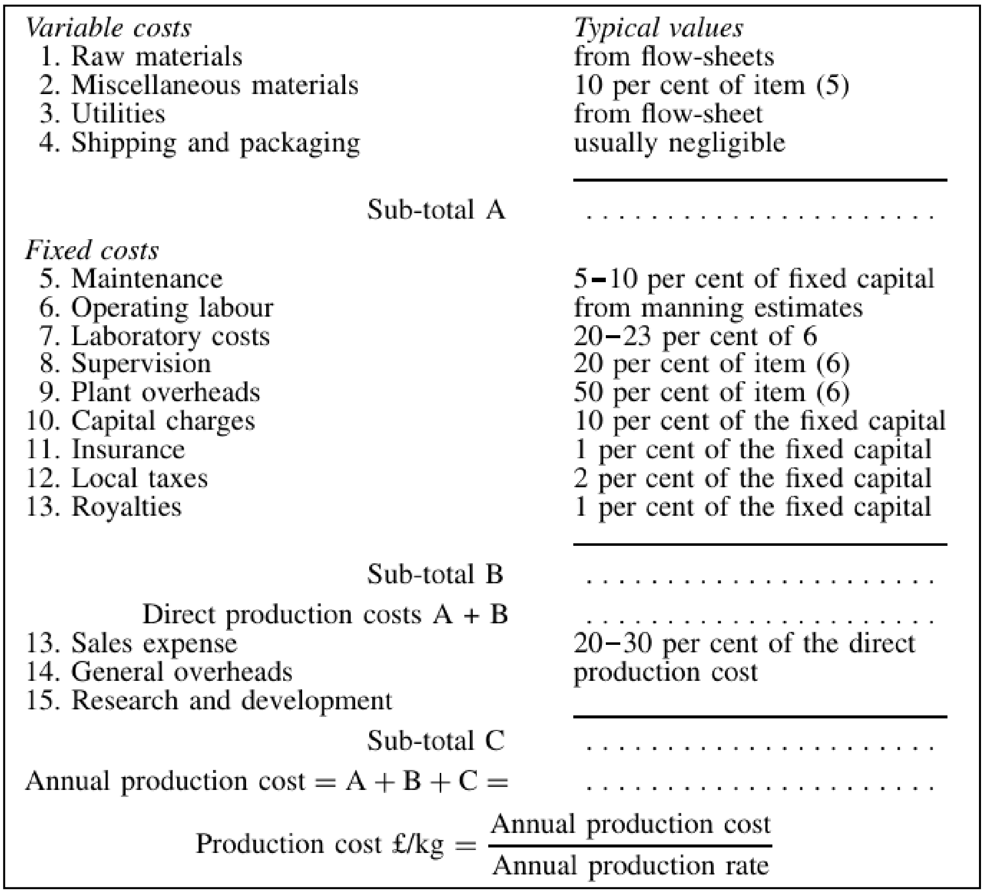
Table 18: Variable Operating Cost

Miscellaneous Operating Material (10% of Maintenance Cost)	72,367.2
Total VOC	**72,367.2** £

7.4.2 Fixed Operating Cost

Table 19: Fixed Operating Cost

Maintenance (5% of FC)	7,23,672
Total Operating Labor (OL)	30,000
Laboratory (30% of OL)	9,000
Plant Overheads (50% of OL)	15,000
Total FOC	**7,77,672** £



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 =	. . . .

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

Figure 8-5: Production cost factor

8.6 Annual Production Cost

Direct or Annual production cost is the sum of Fixed and Variable Costs that we calculated above,

$$Annual\ Production\ Cost = FOC + VOC$$

$$Annual\ Production\ Cost = 8{,}50{,}039\ £$$

For calculating the production cost per kg we use the following formula,

$$Production\ Cost\left(\frac{£}{kg}\right) = \frac{Annual\ Production\ Cost}{Annual\ Production\ Rate}$$

Plant Capacity tons/year = 1,73,448

Production Cost per ton = 4.90£

Profit per ton = 24.50

Annual Profit = 4,250,197

8.7 Payback Period

The following calculation is used to determine the payback period:

$$Payback\ Period = \frac{Total\ Investment}{Annual\ Profit}$$

Payback Period = 3.57 years

The cost research has helped us gain a better grasp of the projected plant's financial aspects and feasibility. We may say that our project is financially feasible based on current cost estimates because our payback period is only 3.5 years, and the capital cost is low. The strategy should be considered financially sustainable if the repayment term is less than five years. As a result, we can say that we are well-positioned to complete this job.

CHAPTER 9

HAZOP ANALYSIS

The hazard and operability analysis is an essential step in the design of a chemical plant. It assists in identifying any risks that may result in accidents. After evaluating the list of possible hazards, it is critical to put the recommended preventive actions into action to either eliminate or drastically decrease the possibility that any incident would occur.

9.1 HAZOP Analysis on Heat Exchanger

Table 20: HAZOP study on heat exchanger

PARAMETER	GUIDE WORDS	DEVIATION	POTENTIAL CAUSES	POTENTIAL CONSEQUENCES	ACTIONS REQUIRED
FLOW RATE	MORE	Increased flow velocity of untreated naphtha on the shell side	The naphtha inlet valve did not shut. Malfunctioning pump.	Decreased heat transfer efficiency. Reduced increase in temperature.	Set up alarms in case of very high flow rates. To control flow, install flow control valves.
	LESS	Reduced flow velocity of untreated naphtha on the shell side	leaks in pipes Obstruction of a pipe	The process fluid's outlet temperature is greater than necessary.	Install flow meters at the shell side inlet. To control flow, install flow control valves.
	NONE	Absence of naphtha flow	Pump malfunction Closing of a valve	No change in naphtha's temperature.	Put temperature sensors at the untreated naphtha's intake and output. Install flow meters at the shell side's intake.

TEMPERATURE	HIGH	The highest temperature of the raw naphtha's outflow.	increased reactor-generated treated naphtha flow rate. Reduced flow rate of untreated naphtha	Overly high output temperature is necessary.	High-temperature sensors at the shell side's outflow. The flow control valve is located at the tuning side's intake.
	LOW	Lower the untreated naphtha's output temperature.	The increased flow rate of untreated naphtha	The process fluid's ideal temperature was not attained.	Install the flow meter at the tube side's intake. At the tube intake, install a flow control valve.

9.2 HAZOP Analysis on Fired Heater

Table 21: HAZOP study on fired heater

PARAMETER	GUIDE WORD	DEVIATION	POTENTIAL CAUSES	POTENTIAL CONSEQUENCES	ACTIONS REQUIRED
FLOW	LOW	Low-pressure intake	The pipeline is leaking. The valve doesn't open all the way.	incomplete burning. Buildup of flammable gas inside the furnace. may cause an explosion.	Put in a flow metre. routine upkeep and observation of the valve.
	NO	Absence of oxygen supply	defective operation of the valve.	No reaction occurs.	Replace valve.

PRESSURE	HIGH	elevated general pressure	impediment or blockage in the flue gas stream. The pressure release valve has failed.	Pressure within the fired heater increases may result in an explosion.	Put pressure sensors in the route of the flue gas. Change the valve for pressure relief.
	LOW	The heater that is fired has low pressure.	inadequate availability of gasoline. Insufficient air available for burning. leaks throughout the structure.	reduced effectiveness of heat transmission. Potential for a flameout.	Check for any obstructions or leaks. Assure a sufficient flow of air for combustion.

9.3 HAZOP Analysis on Hydrotreater:

Table 22: HAZOP study on Hydrotreater

PARAMETER	GUIDE WORD	DEVIATION	POTENTIAL CAUSES	POTENTIAL CONSEQUENCES	ACTIONS REQUIRED
TEMPERATURE	HIGH	The reactor temperature is high.	Catalyst Degradation Increased Reactor Pressure	Reduced effectiveness of the catalyst. Hydrodesulfurization has decreased.	Regularly check the condition and functioning of the catalyst. Install pressure control valves and give them routine inspections.
	LOW	the low temperature of the reactor	Insulation is inadequate. Ineffective transmission of heat in a fired heater or heat exchanger.	reduced reaction time. low reactor conversion.	Boost the insulation. Install temperature sensors at the reactor's intake and exit.

VOLUME	LESS	The reactor volume is less.	Inlet valve malfunction for the feed. Breach within the reactor. directing the reactants.	Kinetics of the reaction would be impacted.	Check the accumulation inside the reactor regularly. Flow sensors are used to monitor the flow rate. Examine any potential leaks.

CONCLUSION

The proposed process has been efficiently designed to meet the requirements i.e. the final product which is desulphurized naphtha contains less than 1 ppm sulphur. The process has also been optimized with main focus on reactor and distillation column. Process optimization is also done in terms of production, cost and energy efficiency while taking into consideration the environmental hazards at the same time. The project offers the following benefits:

- The use of CoMo/ Al_2O_3 catalyst ensures the conversion of sulphur compounds up to 97% while being cost effective as well.

- The process is made energy efficient by use of heat exchangers for preheating reactor inlet and for preheating distillation column inlet with column bottom product thereby reducing energy requirements.

- The process is environment friendly since the hydrogen sulphide which is highly toxic gas is sent to sour gas treatment section where it is treated to form H_2S water. Hence in this way it is made sure that it does not directly release into environment.

- The results of the process simulation validated the calculations and design of the process and verified the operability of the process.

- The process is evaluated for a pay-out period of 3.57 years which indicate that the process is not only feasible but economically profitable as well.

- Different controllers are used throughout the process in order to make sure that the designed process works within safety regulations so that there is not unnecessary downtime or equipment failure.

- Hazop analysis of the process was also done in order to make sure that the possible hazards can be dealt efficiently while maintaining safe environment for the hydrodesulphurization of naphtha.

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