STUDY TO MAXIMIZE ASPHALT AND MINIMIZE FURNACE FUEL OIL (FFO) PRODUCTION AT HEAVY CRUDE UNIT OF ATTOCK REFINERY LIMITED.

By

Safia Sabir

Usama Zulfiqar

Muhammad Hamza Nazeer

School of Chemical and Materials Engineering

National University of Sciences and Technology

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By

Leader - 242481 Safia Sabir

Member 1 - 254029 Usama Zulfiqar

Member 2 - 265349 Muhammad Hamza Nazeer

A THESIS

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CERTIFICATE

This is to certify that work in this thesis has been completed by **Ms. Safia Sabir, Mr. Usama Zulfiqar**, and **Mr. Hamza Nazeer** under the supervision of Dr. Ameen Shahid at the School of Chemical and Materials Engineering (SCME), National University of Sciences and Technology, H-12, Islamabad, Pakistan.

Advisor--

Dr. Ameen Shahid

Department of Chemical Engineering

School of Chemical and Materials Engineering

National University of Sciences and Technology

HOD--

Dr. Erum Pervaiz

Department of Chemical Engineering

School of Chemical and Materials Engineering

National University of Sciences and Technology

Principal/Dean--

Dr. Amir Azam Khan

Department of Chemical Engineering

School of Chemical and Materials Engineering

National University of Sciences and Technology

DEDICATION

This Thesis is dedicated to our Parents and Teachers, who supported us throughout the educational journey. Thanks for enabling us to see this road to the end.

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In the first place as always before and after, Gratefulness and Gratitude for Allah.

This thesis might not have been possible to conclude without the support and assistance of our Parents and Teachers.

We would like to voice our appreciation to **Mr. Mohsin Ali Khan** from **Attock Refinery Limited** for providing us with all the data needed to complete this project.

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Authors

ABSTRACT

The project focuses on the Vacuum Distillation Unit (VDU) design to refine the residue of the Atmospheric Distillation Column into the Asphalt, which is our targeted product.

The simulation has been carried out through Aspen HYSYS V11 Program. The project includes mass and energy balance calculations, the design of equipment, the economic analysis, the HAZOP Analysis, and improving the production of Asphalt.

Cost analysis helps us to estimate the total cost of VDU, through the economic summary from Aspen HYSYS. This study concluded that it is feasible to increase Asphalt production by adjusting the flow rate of Atmospheric Distillation Column residue. This is the main objective we worked for.

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ΔH: Change in enthalpy

A: Area

C: Cost

^{o}C: Degree centigrade

D: Diameter

F: Flow Rate

g: Acceleration due to gravity

H: Height

J: Joule

L: *Length*

mol: Mole

P: Pressure Q:
Energy Flow rate t:
time

V: Volume

W: *Width*

Cp: Specific Heat Capacity at Constant Pressure

T: Temperature

V: Volumetric Flow Rate

K: Thermal Conductivity
h: Thermal Convectivity

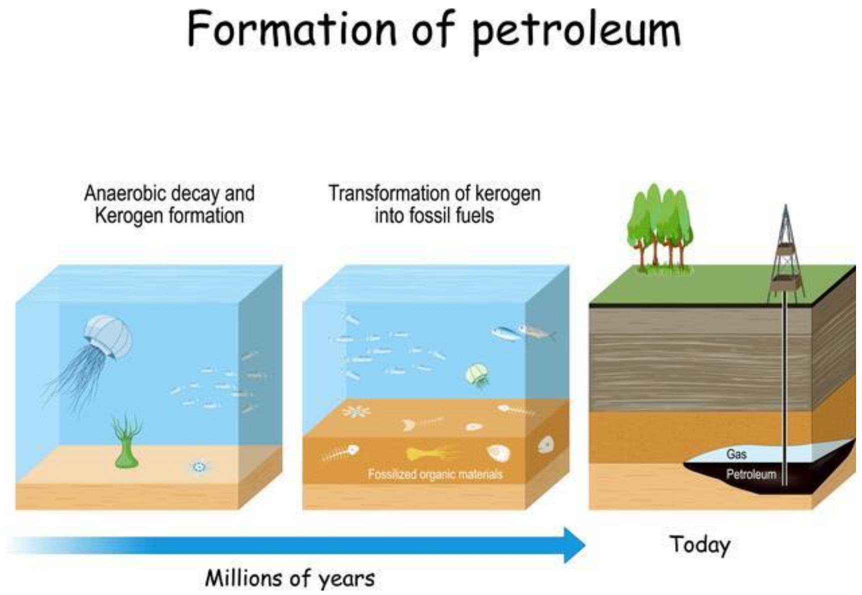
CHAPTER 1

INTRODUCTION

1.1 OVERVIEW

Petroleum with coal and natural gas are known fossil fuels. Fossil fuels are produced when animals and plants in the sea die, their remains get deposited under several layers of sand, mud, or silt. Usually, fossil fuels are formed after millions and billions of years that's why petroleum is also classified as a non-renewable energy source.

Figure 1: Formation of petroleum



Petroleum comprises hydrocarbons (a compound composed of hydrogen and carbon) and some other compounds, mainly sulfur. When petroleum is collected in its natural form, it is known as crude oil and may be black, green, or clear. It can be thick like tar or thin like gasoline, depending upon the geographical location of extraction.

1.2 POTENTIAL SITES OF PETROLEUM

There are many oil-containing reservoirs in various regions around the globe. The largest oil fields are in Saudi Arabia and Kuwait, although other Middle East countries like Iran and Iraq are also significant contributors to the total world production rates.

The North Sea oil sites are as yet almost full and can be classified as the second most important in terms of economics.

Texas oil fields are empty now, but they were known as the world's largest oil-producing region.

The first known oil was drilled by Edwin Drake in 1859, in Pennsylvania. After this time petroleum and oil production graph grew drastically. [1]

1.3 CHEMISTRY OF PETROLEUM

Petroleum is composed of different hydrocarbons. The most abundant hydrocarbons are alkanes, also known as linear or branched hydrocarbons.

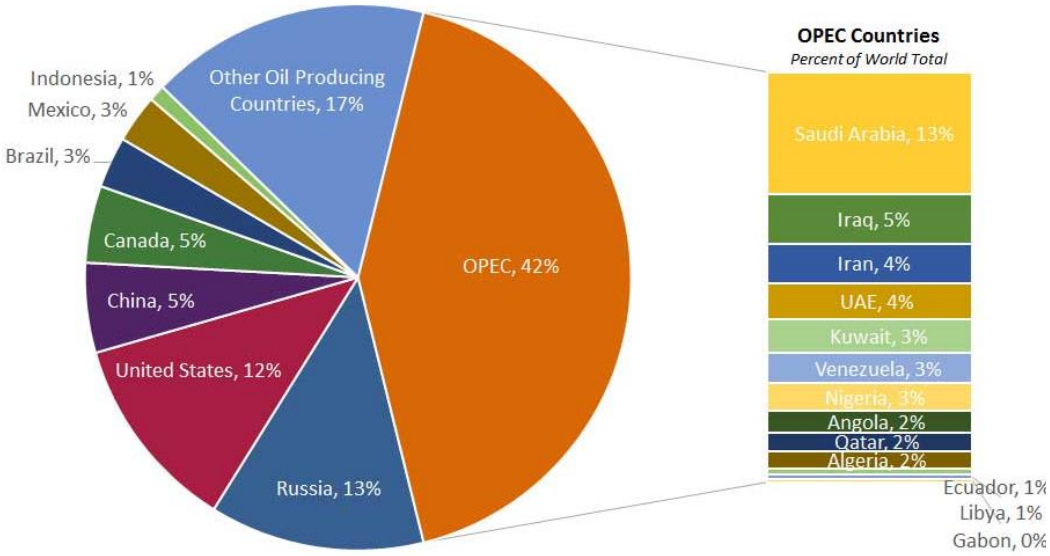


Figure 2: Petroleum abundances

The remaining part of the composition contains aromatic hydrocarbons and cycloalkanes. Moreover, petroleum also contains asphaltenes, a group of complex hydrocarbons.

Different types of crude oil obtained from different geographical locations have different properties like viscosity and color. This diversity is because different arrangements of molecules are affected by the overall percentage of different hydrocarbons it contains.

- **Alkanes** are the primary form in which most of the hydrocarbons are present, also known as paraffin. They are classified as saturated hydrocarbons that are present in straight molecule chains or branched chains. The alkanes are a type of very pure hydrocarbons that contain only carbon and hydrogen. The combustible nature of petroleum is due to the presence of this paraffin.
- The **cycloalkanes**, also known as naphthenes, are classified as saturated hydrocarbons. Here saturation means that each molecule contains one or more carbon rings with hydrogen atoms linked to them. This type of hydrocarbon has identical properties to paraffin but are combustible at very elevated temperature.
- **Aromatic hydrocarbons** are another type of unsaturated hydrocarbons, that contain benzene rings with hydrogen linked to them. They tend to produce a higher degree of emissions when combusted, most of them have a sweet odor hence known as aromatic hydrocarbons.

In industry, the percentage and quantity of different hydrocarbons can find out in a laboratory through testing. The testing process consists of extracting the specific molecule using a suitable solvent and then subsequent separation by gas chromatogram. In the end mass spectrometer can be used to explore the separated molecule in the chemical compound of the extracted sample. [2]

1.4 CLASSIFICATION OF PETROLEUM

For many decades now the petroleum industry has been classifying crude oil based on the geographical location from which it is extracted. But all crude oils extracted from the same location are not ranked equal. However, the further classification is based on the different non-hydrocarbon compounds (specifically Sulfur) and density (API Gravity).

If crude oil has a high degree of Sulfur content, then it is recognized as sour.

If it contains a relatively low degree of Sulfur content then it is classified as sweet. If crude oil is of high density, it is known as heavy and has low density then known as light. Density is a function of the length of the hydrocarbon chain. The longer the chain higher will be the density. The ratio of carbon to hydrogen(C/H) also influences the density of hydrocarbon. The greater the amount of carbon in relation to hydrogen, the denser the hydrocarbon is. More dense oil sinks to the bottom of lighter oil and is generally difficult to pump.

Classification of crude oil is done into the following four categories:

1.4.1 PARAFFIN

About 15 to 60% of crude oil is composed of paraffin and has a C/H of 1:2, which means the amount of hydrogen is almost double the amount of carbon. They have branched or the straight chain but never have cyclic compounds. These are the most required component of crude and impart combustion properties to it. The less complex the paraffin is, the lighter the crude is.

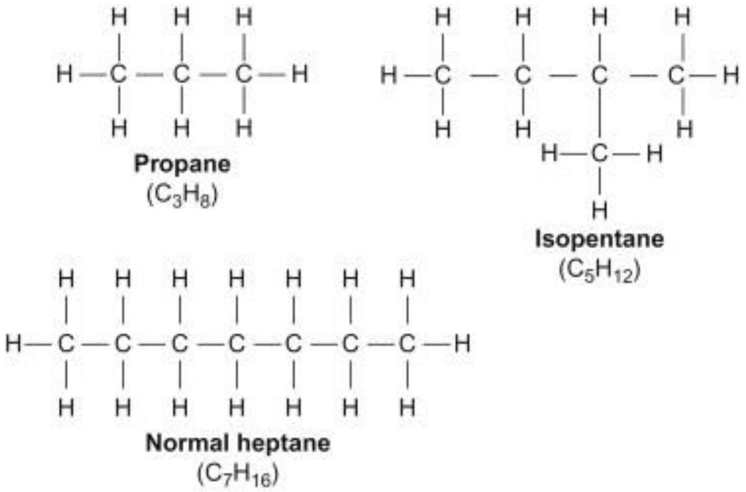
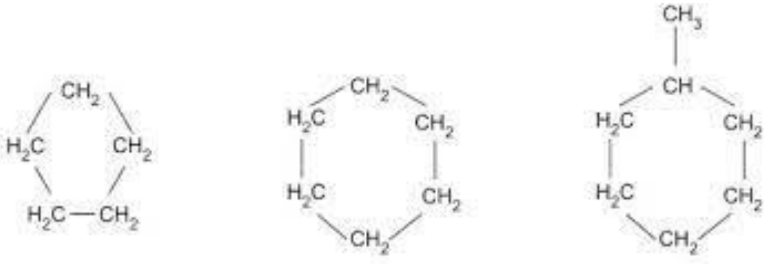


Figure 3: Structures of common hydrocarbon compounds

1.4.2 NAPHTHENES

About 30 to 60% of crude oil consists of naphthenes and has a C/H of 1:2.

Figure 4: Cyclic compound structures



Cyclopentane (C_5H_{10})

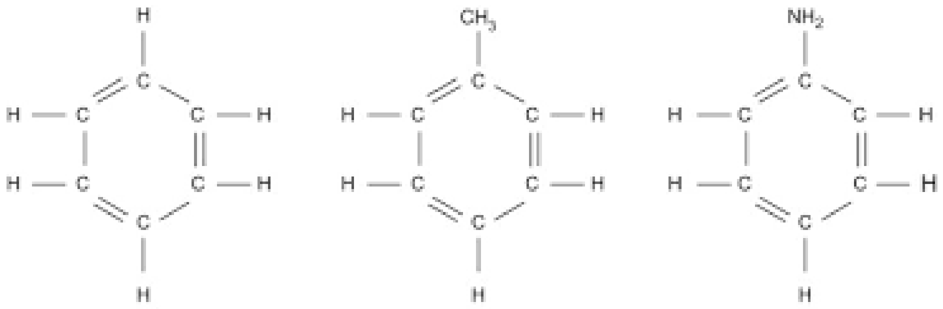
Cyclohexane (C_6H_{12})

Methylcyclohexane (C_7H_{14})

These are also known as cycloparaffins. They are more viscous and have a higher density as compared to equivalent paraffin.

1.4.3 AROMATICS

The composition of crude oil ranges from 3 to 30% of aromatics. They have much low hydrogen as compared to carbon, highly undesirable because their burning produces soot and is highly viscous. Often, they are present in solid or semi-solid form.



Benzene (C_6H_6)

Toluene (C_7H_8)

Aniline ($C_6H_5NH_2$)

Figure 5: Aromatic compound structures

1.4.4 ASPHALTICS

These can make up about 6% of crude oil. They have a C/H of 1:1. This density is very high and usually undesirable because their sticky nature makes them good for road construction.

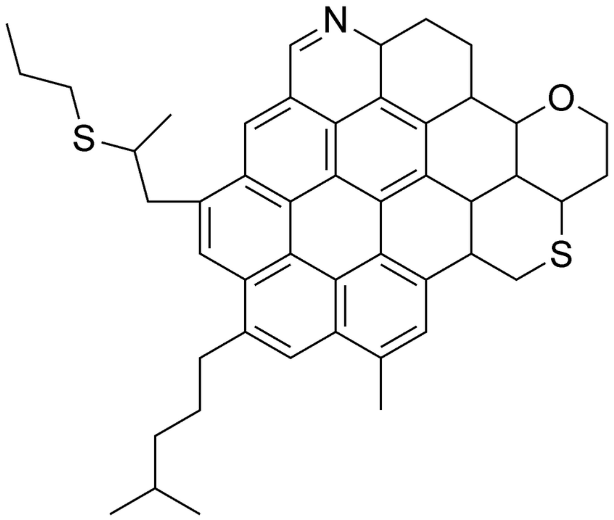


Figure 6: Asphaltic compound structure

When dealing with petroleum classification it is important to understand that overall petroleum classification affects the physical properties, transportation expenses, and refining expenses.

1.5 TYPES OF CRUDE OIL

Normally petroleum classification is based on geographical location, but this is not a suitable method of classification. Because this method provides very less information about general properties like physical state, toxicity, and changes occurring as a result of weathering and time. Oil spill response is based on these characteristics. In the response scenario, the following classification is the most suitable one.

- **Class A:** Volatile and Light Oils have a strong odor, high evaporation rate, is highly fluid, spread fastly on water and solid surfaces, are often clear, and are flammable. they can penetrate porous material like sand and dirt and may have the retarded flow in such material. They do not adhere to the surface, washing with water usually removes them. These are highly toxic and carcinogenic for humans, marine life, and other organisms. Most of the highest quality crude oil and most refined petroleum products are generally included in this category.

- **Class B:** These are non-sticky oils, and have oily and waxy structures. These can firmly adhere to the surfaces and can only be removed by vigorous washing. Their penetration tendency to porous material increases with an increase in temperature. These are less toxic as compared to class A. by evaporating them we can get the residue of class C or class D.
- **Class C:** These are sticky and heavy, highly tarry, viscous, and brown or black in appearance. The doesn't penetrate readily penetrate porous surfaces and washing water will not remove them from surfaces. They have a density nearly equal to water, so they usually sink. Evaporation of class C oils leads to the formation of class D residue. They offer less toxicity but endanger living species when consumed by them. Medium crude, heavy crude, and residual fuels make a considerable portion of it.
- **Class D**: These are black or brownish in appearance, less toxic, and do not penetrate porous matrix. When they are heated, they melted and cover the surfaces subjected to difficult cleaning. Heavy crude oils, residual oils, and high paraffin are included in this class.

These classifications are generally fluid for spilled oils. Oil behavior and composition of finished products are influenced by temperature and other weathering conditions. For example, when light components are evaporated from class B oil, we may get class C oil. If the temperature in the surroundings decreases, class C solidifies into class B oil.

1.6 REFINING OF CRUDE OIL

Refineries convert crude oil into refined petroleum products that can be used for heating, generating electricity, paving roads, and as feedstock to produce various chemicals.

Refineries convert crude oil into various products which are then purified separately. A petroleum refinery is an industrial facility that is expensive and complex. Each refinery has the following three basic products:

1. Separation
2. Conversion
3. Treatment

1.6.1 SEPARATION

In the atmospheric distillation column, the vapor-liquid mixture separates into petroleum fractions based on their boiling point. Light fractions are taken from the top and heavy from the bottom.

Liquified Petroleum Gas (LPG) and gasoline are light fractions, that vaporize and reach the top where they condensed again to liquid.

Medium fractions have distillates and kerosene present in the middle of the column.

The heaviest fractions with elevated boiling points stay at the bottom of the column.

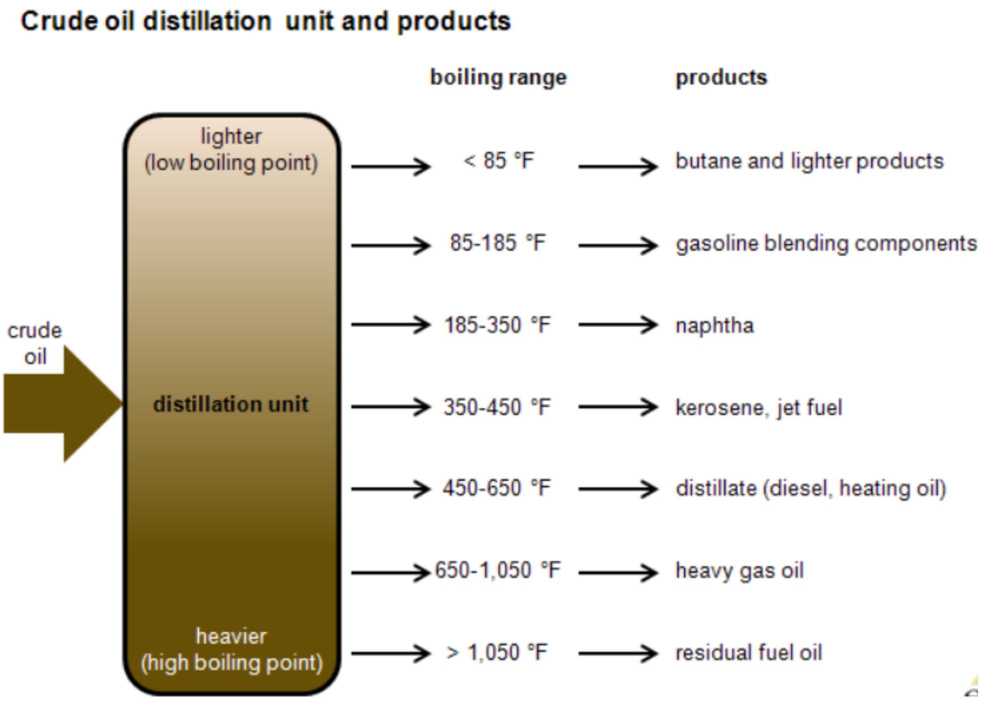


Figure 7: Division of crude oil products

1.6.2 CONVERSION

After distillation is completed, lower value and heavy fractions are further converted into lighter and more refined products such as gasoline. From here fractions of distillation are transported to other sections where they become finished products.

Cracking is one of the most widely used conversion methods because it utilizes temperature, pressure and catalyst, and occasionally hydrogen as well for a breakdown of heavier hydrocarbons into lighter ones.

A cracking unit constitutes one or more thick-walled reactors with a series of furnaces, other vessels, and heat exchangers. They are further classified as **hydrocracking units** and **fluid catalytic units** depending upon the method of operation.

Cracking is not always the only option in cracking. Sometimes the process of rearranging molecules adds more value than breaking molecules.

Alkylation, for instance, synthesis of gasoline by combining gaseous products of cracking. This process is the inverse of cracking, usually taking place in skinny towers and horizontal, large vessels.

Reforming utilizes high temperature, moderate pressure, and a suitable catalyst to convert naphtha into gasoline with high octane number.

1.6.3 TREATMENT

This is the final finishing touch to the products. To produce synthesized gasoline with proper octane level, flash point, and other important properties, refineries combine carefully different streams from different processing units.

The incoming raw crude oil and outgoing finished products are needed to be stored in tanks or vessels in a location near the refinery. From here the products are dispatched and crude oil is directed towards the refinery for processing.

1.7 FUTURE OF PETROLEUM

Experts seem to agree on the fact that petroleum production will reach its peak in the next 20 years. Peak doesn't mean that petroleum reserves would be unable to deliver oil, but that the rate of extraction has been raised to the highest level and further extraction will not accelerate the rate forward. Consequently, the rate of petroleum products output will fall. So, the following are the scenarios that might be possible in near future.

1.7.1 HEAVY AND SHALE OIL

Efforts have already been put to excavate the oil that one considered uneconomical to draw. As the demand for light, easily drawn able oil continues to increase and its supply continues to decrease, then the price of one barrel of crude oil will also sike. As a result, heavy oil that was once costly to extract now generates profit.

Countries like the US, Canada, and Venezuela are sitting on the top of the largest reserves of heavy and shale oil. It is calculated that there is more total oil in Venezuela than in the whole Middle East. Canada is so blessed with heavy oil that it can have the capacity to meet the whole world's future oil demand for nearly 200 years.

The main two considerations for the heavy oil production method are that energy invested is higher than energy returned ad these have less tendency to impact the environment. While oil demand is rising exponentially, the environmental lobbies are concerned about the long-lasting effect of the extraction of heavy oil. Environmental effects arise from the fact, that increase energy invested than energy returned and there is the possibility of the production of more greenhouse gases and other contaminants than equivalent crude oil. We can conclude that by extracting and using heavy oil we are going to increase global warming by accelerating the emission of greenhouse gases.

What is obvious is that the production of heavy oil will be mandatory in the coming era unless there is a rapid decrease in demand. Development of several techniques has been initiated to extract heavy oil in the way to have to lessen the impact of extraction on the environment, there is still little ambiguity that the use of these resources may hurt the environment. Due to this reason conservation is more crucial to do.

Conservation is having more concern about accelerating the use of oil and less about running out of oil reserves. Environmentalists believe that sufficient time and money should be spent on research and innovation in extraction processes to develop alternate strategies.

1.7.2 ELECTRIC VEHICLES

The transportation industry is utilizing about 70% of crude oil production, and many initiatives have been taken to design an electric vehicle that can have similar performance to the vehicle utilizing petroleum. The development of new technologies has been lengthening the mileage range from 100 to more than 200 miles. Major contributors that hinder the use of the electric vehicle are the price of batteries, manufacturing, and recycling of batteries and the time battery takes for charging. In conclusion, the major contributor to the less use of electric vehicles is how to store energy in the form of electricity when the vehicle is idle. A realistic, efficient, and robust way to replace the batteries would make use of electric vehicles possible.

It is important to consider that although electric vehicles reduce petroleum utilization of fuel but still a source of electricity is needed to charge vehicles. If this energy source is not clean and renewable then the problem is still unsolved, it has been subjected to change of shape just. Implementation of sources that are renewable for the generation of electricity is needed to develop successful use of electric vehicles.

1.7.3 CONCLUSIONS

We agree on a point that petroleum plays a major role in our daily life and will continue to do so in the coming years. While technologies have been developed to decrease our reliance on fossil fuels but still it will be many years before they become practical and affordable.

Even if the world wants to shift to an energy source that is petroleum independent, we still need petroleum to produce versatile products like cosmetics, pharmaceuticals, plastics, etc. so this is our reason to be conservative about fuel.

CHAPTER 2

LITERATURE REVIEW

2.1 TYPES OF CRUDE OIL

Generally, crude oil has been classified based on API Gravity into the following main categories:

I. Light Crude Oil
II. Medium Crude Oil
III. Heavy Crude Oil

Figure 8: Crude oil types

Crude Oils	Familiar Substances	Typical API Range
Tar, Bitumen and Kerogen	Window Putty Caulk Vegetable Shortening	6 – 10
Extra-Heavy Oil	Peanut Butter Tomato Ketchup	10 – 12
Heavy Oil	Molasses Honey	14 – 22
Intermediate Oil	Maple Syrup Corn Oil	25 – 30
Light Oil	Water	31 – 40
Ultra-Light Oil	Nail Polish Remover	41 – 50+

2.1.1 LIGHT CRUDE OIL

It is characterized by very low density and can move freely at ambient room temperature. It is known for its low specific gravity, high API Gravity, and low viscosity because it contains a high percentage of low molecular weight hydrocarbons. Light crude oil finds a high price in the consumer market because it is the type of oil that yields gasoline and diesel when subjected to refinery operations.

2.1.1.1 STANDARDS

- NYMEX (New York Mercantile Exchange) illustrates that US light crude oil has API gravity ranges between 37° API and 42° API and density ranges between 840 kg/m^3 and 816 kg/m^3 whereas non-USA oil has API gravity ranges between 32° API and 42° API and density ranges between 865 kg/m^3 and 816 kg/m^3.
- NEB (National Energy Board) of Canada illustrates the light crude oil with API gravity higher than 30.1° API and density lower than 875.7 kg/m^3.
- However, Alberta's government did not agree with Canada and defines light crude oil with API gravity higher than 35° API and density lower than 850 kg/m^3.
- Pemex (Mexican State Oil Company) illustrates light crude oil with API gravity ranges between 27° API and 38° API with density ranges between 893 kg/m^3 and 835 kg/m^3.

The reason for the difference in definition is due to fact that Mexico and Canada have large crude oil reserves than the USA where oil that is produced from reserves is lighter. Moreover, the difference in measuring units is also a major contributor to this variation as Canada uses SI units while the US uses American units. Thirdly difference lies in the reference temperature used for measuring density as Canada does the calculation at 15 °C while the US used to do the calculations at 15.56 °C.

2.1.2 MEDIUM CRUDE OIL

These are the most used type of crude oil. They are less viscous than heavy crude oil within API gravity ranges between 22.3° and 31.1°. These are less dense and less volatile than lighter crude oil, so they are less prone to evaporation.

These are a combination of light and heavy oils. A suitable refinery operation is performed to separate the desirable light fractions of hydrocarbons. The leftover heavy fractions after refining are known as residue and are further processed to separate useful products from it.

2.1.3 HEAVY CRUDE OIL

It cannot flow freely at room temperature and is highly viscous with specific gravity, molecular weight, and density higher than light and medium crude oil. It has API gravity lesser than 20^{o}. Asphalt and heavy oil are DNAPLs (Dense Non-Aqueous Phase Liquids). Their viscosity and density are higher than water. A large lump of DNAPLs can quickly penetrate the aquifers and settle at the bottom.

2.1.3.1 STANDARDS

In 2010, WEC (World Energy Council) defined extra-heavy crude oil with API gravity less than 10^{o} and viscosity higher than 10,000 centipoises. When there is no way to measure reservoir viscosity then according to WEC, extra heavy oils have API gravity less than 4^{o}. so heavy crude oil has a density greater than 100 kg/m^{3}, the viscosity of the reservoir greater than 10,000 centipoises, and specific gravity of more than 1.

2.1.3.2 CHEMISTRY

Heavy crude oil contains resins and asphaltenic compounds. It is called heavy oil because it contains a higher proportion of naphthenes and aromatics than linear chains and higher percentages of nitrogen, sulfur, oxygen, and metals (NSOs). Compounds with several carbon atoms greater than 60, are a major constituent of heavy crude oil thus contributing to its heavy molecular weight. For example, 1000-5000 cP viscosity has been reported for Venezuela's extra-heavy crude oil while the extra-heavy crude oil of Canada has a viscosity lies between 5000-10000 cP.

Chevron Phillips, a famous chemical company quoted:

The strength of heavy crude oil is due to the presence of complex, nonparaffinic, high molecular weight, and a low percentage of volatile and low molecular weight compounds.

2.1.3.3 CLASSIFICATION OF HEAVY CRUDE OIL

Heavy crude oil has been classified into the following two categories:

Those that contain sulfur greater than 1% in asphaltenes and aromatics. These are mostly present in the United States, North America, South America, and the Middle East.

Those that contain sulfur less than 1% in asphaltenes, resins, and aromatics. These are mostly present in Central Africa, Western Africa, and East Africa.

2.1.3.4 RELATED COMPOUNDS

Heavy crude oil is equivalent in physical properties to bitumen obtained from oil sands because the density of bitumen is less than 10°API. According to a recent geological survey carried by the US, bitumen also famous as tar sand, is classified as extra-heavy crude oil with a viscosity higher than 10,000 cP, and heavy oil is different from light oil in terms of viscosity measured at reservoir temperature, low API gravity and higher percentages of nitrogen, sulfur, oxygen, and metal compounds. Heavy crude oil is like the residue obtained after refining light oil. Heavy crude oil is present at the edge of earth basins and is assumed as the residue of already present light oil which has been subjected to bacterial activity, and weathering conditions.

2.1.3.5 HANDLING CHALLENGES

Refining, transportation, and production of heavy crude oil offer a significant level of challenges. Usually, a diluent is added to heavy crude oil to assist its flow in a pipeline at a regular distance. This diluent is usually a Dilbit. Dilbit is a mixture of condensate, heavy crude oil, and asphalt to achieve the level of pipeline density and viscosity specifications. [3,4]

Figure 9: Heavy crude oil

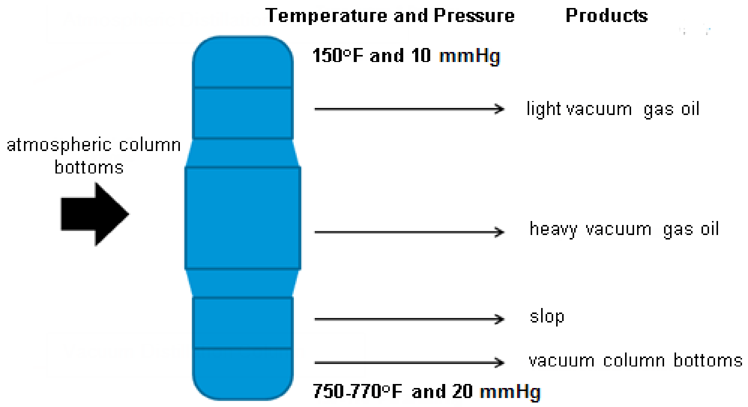
2.2 VACUUM DISTILLATION UNIT

Around 80% of refineries operating in the world possess Vacuum Distillation Column (VDU) along with Atmospheric Distillation Column. It is a secondary processing unit, used to process the residues of atmospheric distillation into useful petroleum products.

The distillation process which is carried out at below atmospheric pressure is known as vacuum distillation. This is carried out by reducing the pressure in the distillation column.

2.2.1 WORKING PRINCIPLE

Figure 10: Atmospheric distillation



In Atmospheric Distillation Unit (ADU), the separation of lighter hydrocarbons from heavy oil is carried out based on boiling point up to a temperature of 750°F. by increasing temperature above this limit, thermal cracking of oil will take place which hinders the further separation process. In an ADU, lighter hydrocarbons are boiled off at relatively low temperatures, but heavier hydrocarbons remain at the bottom of it. To increase the overall efficiency of the process and volume of the products, this residue is directed towards VDU for further refining.

In VDU, vacuum is maintained at a pressure significantly less than atmospheric pressure of 1atm. Thus, the boiling point of the heavier fraction is low enough to cause evaporation without cracking the oil.

The process of vacuum distillation produces various types of gas oils. These are slightly dense than the middle distillate of ADU such as kerosene, jet fuel, diesel, etc. in the next step these gas oils are further purified into lighter products like naphtha, light cycle oil, and gasoline by the process known as Fluid Catalytic Cracking.

2.2.2 HOW VACUUM IS CREATED?

To generate a vacuum in VDU, steam ejectors are used. A steam ejector is attached to a vacuum trap and used to store a vacuum. The receiver of the distillation column is connected by a line to the vacuum trap. Through this receiver, the vacuum is distributed in a vacuum tower. [4]

2.3 STEAM EJECTORS

Gas or steam is used in steam ejectors to compress the gas instead of moving parts. In an ejector, highly pressurized gas-like air is steam is expanded through a cross-section of the nozzle. The air or steam transforms this potential energy of pressure into kinetic energy or velocity. The very high-velocity jet of gas or steam carried the gas to be pumped or ejected out of the suction of the ejector. The mixed stream is then directed towards the diffuser where the transformation of velocity to pressure takes place at the discharge of the ejector.

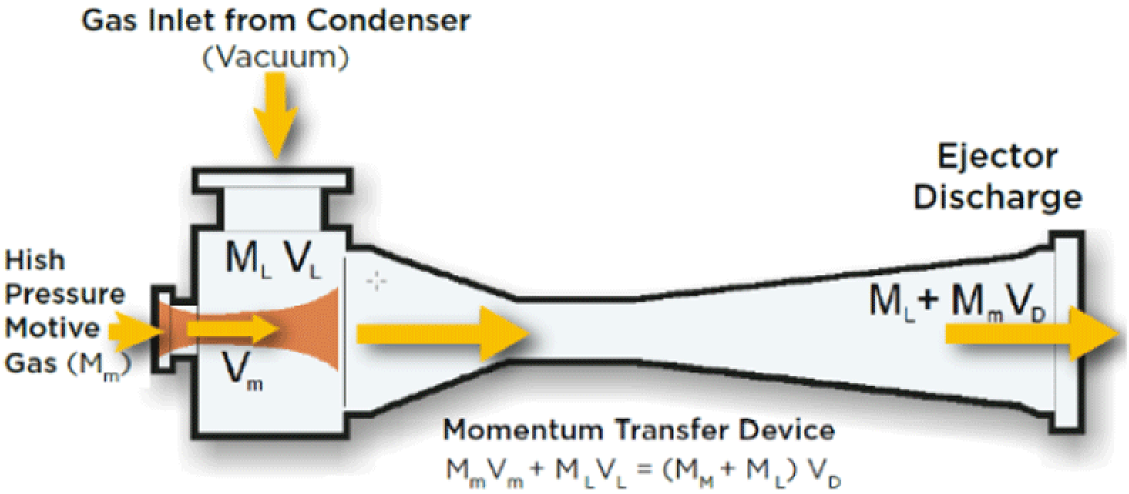


Figure 11: Steam ejector

2.3.1 AIR JETS

Ejectors in which air is used are known as air jets or air ejectors. When steam is not present in a sufficient amount to carry the process then air is used with small ejectors. Along with air jets when the NASH vacuum pump is present, this system can pull the air from the room or the exhaust air of the pump to enhance the vacuum level that can be reached by the pump. A most common application of this type of system is the deaeration of tanks where the vacuum must be able to bring down the vapor pressure of water vapors that are going to be degassed. Systems of this kind of air jet are usually less costly because no highly pressurized air or steam is used to cause it to work, and just a vacuum pump is needed.

2.3.2 HYBRID STEAM EJECTORS

Steam ejectors that are combined with a liquid ring vacuum pump make a hybrid that can make a strong vacuum by consuming a small amount of energy. NASH is mostly used for combining air jets and steam ejectors and ejector vacuum systems. These systems are used by application engineers to have maximum efficiency and performance benefits while optimizing the hybrid system to be applicable for applications, processes, and technological requirements. NASH hybrid steam reduces the emission of greenhouse gases and improves system stability while maintaining the efficiency of the system at the maximum level.

2.3.3 WORKING OF STEAM EJECTORS

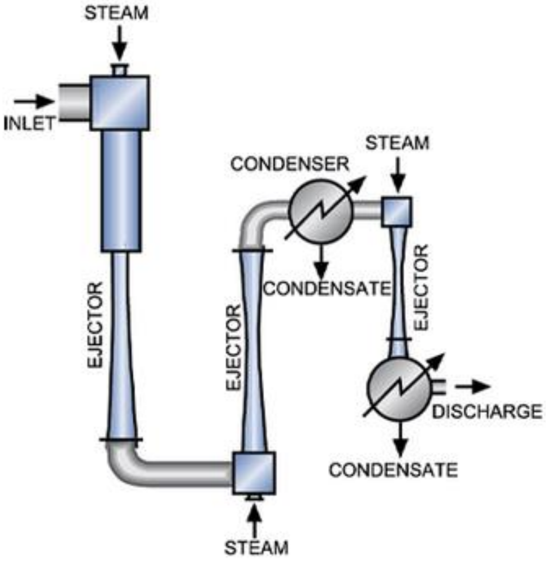


Figure 12: Functioning of steam ejector

In steam ejectors, steam passes through a diverging nozzle.
Expansion of steam and conversion of pressure into velocity is controlled by the nozzle and thus it creates a vacuum for the transportation of gases.
The operation of the steam ejector is on a mass basis not on a volumetric

base. Therefore, steam ejectors apply to low molecular weight gases and considerably low pressures. These find the best use in high vacuum applications.

A jet of fluid moving with supersonic velocity causes the inlet stream to be entrained and accelerate its velocity to the sonic speed as soon as the two fluids mix. In the throat of the diffuser, a stationary sonic wave is generated, and absolute pressure increases rapidly at this point. More pressure is increased as flowing fluid slows down along the exiting end. The most used fluid to create a vacuum is steam which ranges between 80 psig to 400 psig.

Steam ejectors are the most common type of ejectors. A single ejector can create a vacuum up to 27 in Hg. To create a vacuum deeper series of ejectors can be installed. This arrangement is usually preferred because sometimes steam can condense between two stages to reduce the load of the following stage. Steam ejectors have the capacity to create a vacuum to the level where water can freeze. These can be arranged in series to create an absolute pressure of less than 0.1 mm Hg. [5]

2.3.4 STEPS FOR EFFICIENCY IMPROVEMENT

- Use a hybrid system of steam ejectors with NASHA liquid ring vacuum pumps
- Last jet and after condenser should be eliminated and replaced with a highly efficient liquid ring pump
- Pressure of interstage condensers should be optimized
- Interstage pressure and steam flow can be optimized by the interstage ejector.

2.3.5 FITTING OF STEAM EJECTORS

- System can be fitted in any direction but make sure to choose that direction where proper drainage is possible
- Barometric or shell and tube condensers should be placed at the

maximum possible height to prevent flooding and drain the water from the condenser

- Ejector’s outlet should be directed towards a hot well
- If sufficient height is not available for the condenser, then use the
- NASHA pump

2.3.6 PROS OF STEAM EJECTORS

- Doesn’t have any moving parts
- Maintenance is easy
- Construction is easy
- Available in a wide range of materials
- Low operating cost. [6]

2.4 PUMPAROUNDS IN VDU

Pumparounds make it possible to utilize the heat available in the vacuum distillation column through the extended network of heat exchangers while at the same time maintaining the offloading vapors from the column in stable condition. The second important parameter which is needed to be considered is to control the flow of liquid in the column to achieve the desired separation. By using the average vapor-liquid molar flow in any section of the column, the total vapor-liquid flow can be estimated.

Pumparounds dominate and control the separation between different components as they control the vapor-liquid flow throughout the column. If the rate of removal of heat by the Pumparound is decreased, then a load of vapors at the top of the column will increase. Hence the total reflux at the top should also be increased to have a constant temperature at the top. If no Pumparounds are used, then a load of liquid at the top is highest due to high total reflux. Moreover, the quantity of heat rejected to the environment by the top condenser would be maximum.

The installation cost or fixed capital investment by the addition of

Pumparounds is increased due to the power requirement for exchangers and pumps. The extra use of trays for reflux purposes than fractionation also accelerates the working cost. Usually, three pumparounds are used by the crude distillation column because this is the most cost-effective configuration. But two or four can also be used.

2.4.1 PUMPAROUND HEAT REMOVAL

An illustration of pumparounds, also known as circulating flux is shown here.

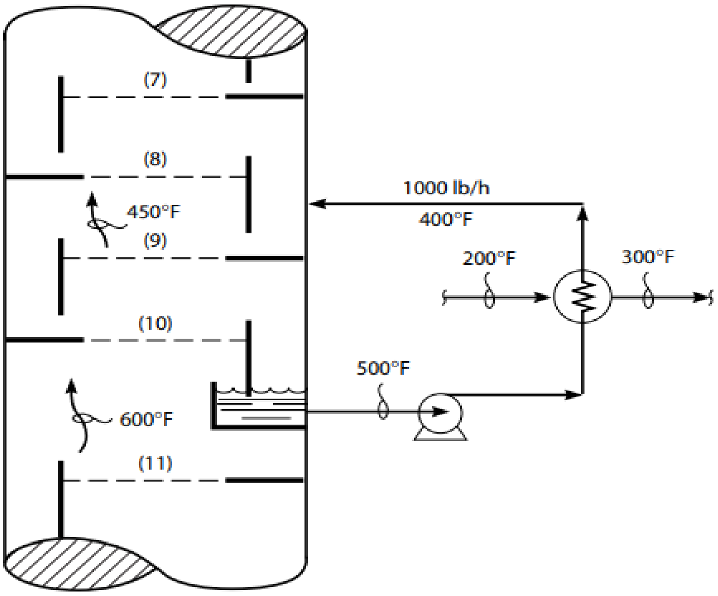


Figure 13: Pumparounds

Heated liquid at 500^0F is extracted from tray no.10, known as the **pumparound of the draw tray. The liquid in the draw stream is cooled to 400^0F and is sent back to the distillation tower to the elevated tray which is tray no. 9. Tray no.9 is called a pumparound return tray.**

The objective of pumparound is to cause cooling and partial condensation of rising vapors. The vapors at the pumparound draw stage are at 600^0F and the pumparound return tray vapors are at 450^0F. in this case, the number of pumparound trays is two. It may vary from two to five depending upon the other operating conditions.

For the calculation of heat removal in pumparound assume the specific heat of the liquid being pumparound is 0.7 Btu/lb.^{0}F:

$$(\mathbf{500}^{\circ}\mathrm{F}-\mathbf{400}^{\circ}\mathrm{F})\boldsymbol{x}(\mathbf{0.7})\boldsymbol{x}\mathbf{1000}\frac{\boldsymbol{lb}}{\boldsymbol{h}}=\mathbf{70000}\frac{\boldsymbol{Btu}}{\boldsymbol{h}}$$

For some liquid flowing on the opposite side of the pumparound the flow rate can be calculated as:

$$\frac{\mathbf{70000}\frac{\boldsymbol{Btu}}{\boldsymbol{h}}}{(\mathbf{300}^{\circ}\mathrm{F}-\mathbf{200}^{\circ}\mathrm{F})\boldsymbol{x}\left(\frac{\mathbf{0.5Btu}}{\mathbf{lb}}\right)}=\mathbf{1400}\frac{\boldsymbol{lb}}{\boldsymbol{h}}$$

This is the quickest way to find the flow rate of fluid by using the heat balance of the heat exchangers.

2.4.2 OBJECTIVES OF A PUMPAROUND

The purpose of circulating reflux is to cool the rising vapors. In the figure given below, the flow of top reflux is controlling the top temperature of the tower. If the rate of pumparound circulation is reduced, then the amount of heat extracted from trays no.9 and 10 is less and it causes the opening of the valve of top reflux. The rate of top reflux would rise. On the top tray, the amount of vaporization would be enhanced. Thus, overhead condenser duty will increase.

If we wish to decrease the heat duty of pumparound coolers it will lead to an increase in the overall heat duty of the top condenser. Thus, the total heat balance of the column is conserved. In the condenser recovered heat is lost to the cooling water while in the case of the condenser heat can be recovered to any process steam. If the outlet temperature of cooling water in the condenser is 140^{0}F then this is not good because the deposition of calcium carbonate began to occur inside of tubes at this elevated temperature. So, it is advisable to keep the outlet temperature of cooling water less than 125^{0}F. The outlet temperature of cooling water can be decreased by increasing the heat removal of pumparound.

Another main objective of pumparound is to lessen the chances of top tray flooding. If flooding happened at tray no.1or 2 then the following conditions will be observed:

- Abrupt increase in tower top temperature.

- Contamination of distillate products with heavier compounds
- Increase of pressure drop across some trays
- Rise in the level of liquid in the reflux drum

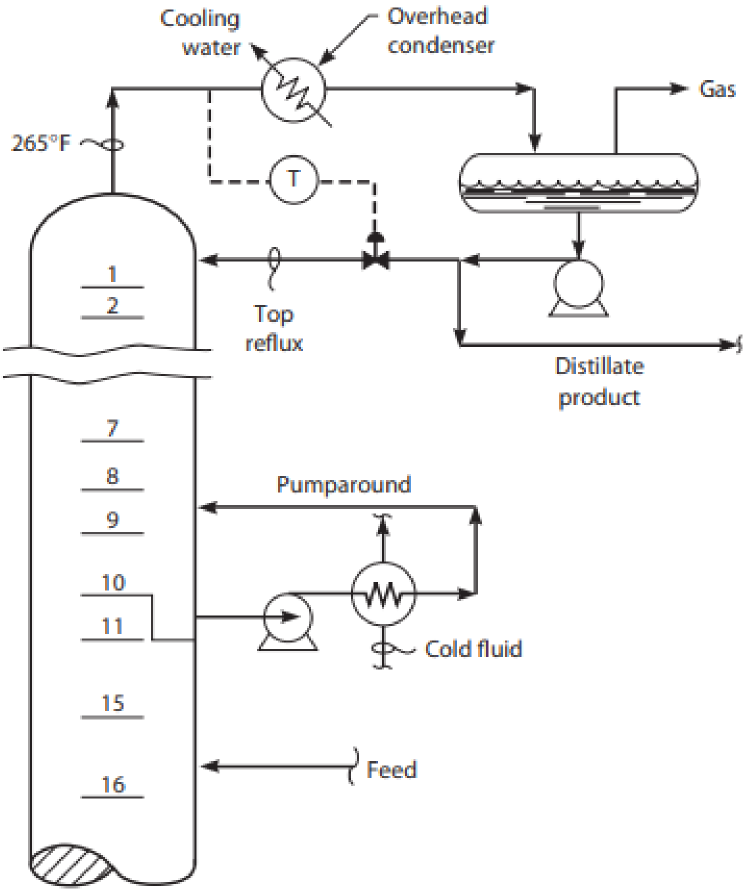
If we increase the reflux rate to lower the top tower temperature, then it will increase the top temperature than decrease it and this is the alarm of flooding at top trays. The corrective action is to first decrease the top flooding by increasing the duty of the pumparound by increasing the flow rate of cold liquid through the heat exchangers of the pumparound or by simply just increasing the flow rate of the pumparound. In such a scenario, there will decrease in the flow of vapors from tray no.1 to 7 and vapor velocity is reduced which will eventually decrease the pressure drop across the trays. The entrainment ability of vapors to hold liquid is reduced, the level of liquid in the downcomer is lowered and the flooding will be decreased.

By increasing the heat duty of the pumparound, the overhead condenser will be unloaded, and this will cool down the reflux drum. A cooler reflux drum can imbibe more gases to distillate and thus fewer gases will be vented. This is often desirable in many refinery operations.

2.4.3 DOES PUMPAROUND AIDS IN FRACTIONATION?

In the process designing it is generally assumed that pumparounds only extract heat from the column and do not expect that the exchange of heat between heated vapors and cold liquid across the trays aids in fractionation. But this is not true.

Figure 14: Pumparounds working



Referring to the figure above, the temperature of vapors rising from tray 9 is 450^{o}F and the temperature of the liquid flowing from tray 10 is 450^{o}F. This kind of temperature difference is evidence that fractionation is happening across the pumparound trays.

$$\Delta T = (\boldsymbol{temperature\ of\ liquid\ leaving\ lower\ tray}) - (\boldsymbol{temperature\ of\ vapors\ leaving\ higher\ tray})$$

This ΔT measures the quantity of fractionation. Higher is this ΔT, more will be the fractionation across the pumparound trays. Studies show that increasing the rate of pumparound circulations rise this ΔT up to a certain point. But if the pumparound circulation rate is increased beyond this, it will decrease the ΔT. This idea is pictured here:

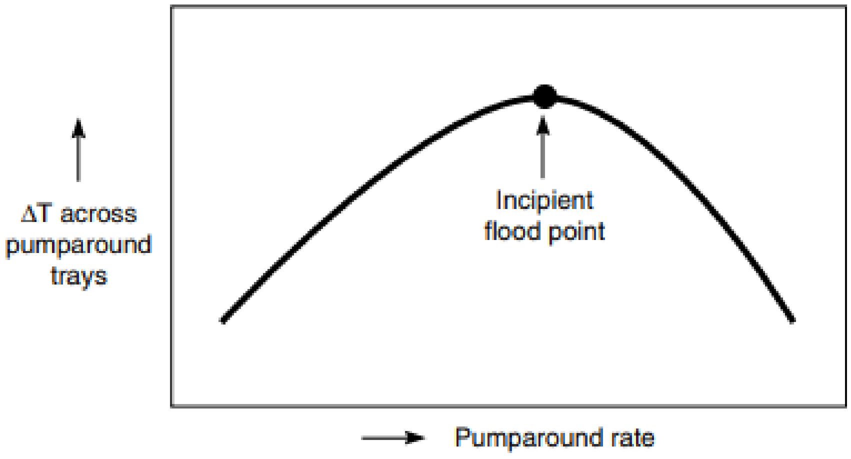


Figure 15: Fractionation

At the same point, there is a large increase in the liquid flow rate of pumparound, this causes the downcomers to be flooded with liquid, and the efficiency of the tray is reduced. At this stage, ΔT becomes smaller. in the case of pumparound trays, this point is known as the **incipient flood point**. [7]

2.5 PRODUCTS OF VACUUM DISTILLATION

2.5.1 HEAVY VACUUM GAS OIL

Heavy Vacuum Gas Oil is a complex mixture of heavy hydrocarbons obtained because of vacuum distillation of the atmospheric column residue. It consists of hydrocarbons with carbon numbers ranging from C20 to C50 with boiling points from 350^0C to 600^0C.

HVGO is mainly used as a feedstock for cracking units to produce valuable middle distillates.

Two main uses in this case are:

- **Gasoline Production:** HVGO is cracked in Fluidized Catalytic Cracker, in the presence of a catalyst to obtain gasoline rich product mixture.

- **Diesel/Kerosene Production:** HVGO is cracked in Hydrocracker using high pressured Hydrogen to obtain a mixture of diesel and kerosene products.
- HVGO can be used as it is without cracking by blending it with residual fuel oil although this use of HVGO is not as valuable as when it is used in cracking.
- HVGO can also be used as feedstock to obtain lube oil base stock after a series of solvent extraction processes.

2.5.2 ASPHALT

Asphalt is also known as bitumen. It is a sticky dark brown/black liquid with high viscosity. It can also be referred to as the semi-solid obtained from the vacuum column bottom residue.

Asphalt is processed to make the final material that is used for different purposes. This final asphalt is composed of 95 % aggregates, which are crushed stones, sand, and gravel, and 5 % bitumen which is responsible for holding the aggregates together.

- The main use of asphalt is in the construction of roads, where asphalt is used to bind aggregates to form asphalt concrete, which is the material that is used on top of the roads as paving.
- The other use of asphalt is in waterproofing products using bitumen called bituminous waterproofing where its waterproofing properties are used extensively.
- Another use of asphalt is in sealing and insulating various building materials like pipe coating and paint etc. It also includes sealing flat roofs and production of roofing felt.

2.5.3 LIGHT VACUUM GAS OIL

Light Vacuum Gas Oil is the lightest fraction obtained from vacuum distillation of atmospheric residue. It is a complex mixture of light hydrocarbons ranging from C25 to C35.

- The main use of LVGO is to produce light and middle distillates through FCC and hydrocracking.
- Hydrocracking of LVGO also produces LPG in addition to light distillates.

2.5.4 SLOP OIL

Slop Oil is a mixture of oil, fine solids, and water. It is a waste stream because of the emulsion of oil with water and fine solids hence it cannot be sold as it is. Slop oil is not eco-friendly and current methods of disposing of it are expensive.

Slop oil needs to be purified before it can be used. After purification:

- Slop oil is used as a fuel in waste oil-fired boilers and incinerators.
- Slop oil can also be sold to refineries. [8]

CHAPTER 3

PROCESS DESCRIPTION

3.1 PROCESS DESCRIPTION

The solution to our problem consists firstly of splitting the Furnace Fuel Oil stream, coming from the atmospheric tower bottom, in such a way that more FFO is sent to Vacuum Distillation Tower for producing Asphalt. This is done by increasing the fraction of the FFO stream entering the Vacuum Column from 0.45 to 0.90. In the splitter, the FFO stream is divided into 3 streams: first stream is cooled down in a heat exchanger and stored. The second stream is used for industrial purposes like in furnace and the third stream is sent to the vacuum column. The existing system is known to be operating at maximum capacity due to certain limitations like capacity of furnace and pumps whose seals cannot handle the increased flow rate and temperature and pressure conditions. After designing the equipment to accommodate the new flow rate, the increased flow rate of FFO is sent to Preheater before Vacuum Column to provide sufficient temperature conditions to the stream. After achieving the required conditions, the FFO stream enters the vacuum tower in which vacuum is created using steam ejectors at the top of the column so that separation of high boiling hydrocarbons can be achieved without thermal decomposition.

After the separation in the vacuum column, the lightest fraction obtained is the Light Vacuum Gas Oil which goes through pump around cooler into the column again to increase productivity and some is stored after cooling it down in a heat exchanger. The next fraction obtained is the Heavy Vacuum Gas Oil which is also pumped around into the column and then stored after heat exchanging. The fraction after HVGO is the slop oil which is simply a waste stream consisting of oil/water mixture. The bottom product obtained from the vacuum column is our desired product asphalt which is also cooled using heat exchanger and stored for further use.

3.2 PROCESS FLOW DIAGRAM

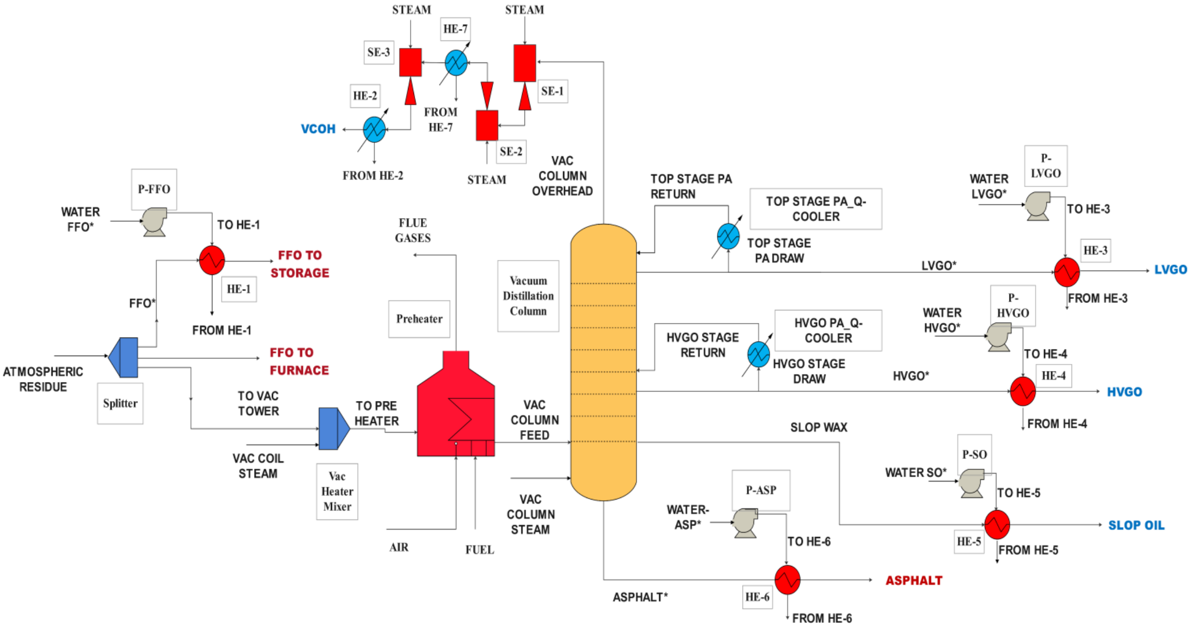


Figure 16: Main process flow diagram

3.3 PREHEATER/FURNACE

The furnace before the Vacuum Column is the most important part of our process because it determines whether the flow rate of FFO to the Vacuum Column can be increased or not depending on the capacity of the furnace. Furnaces usually are preheaters which reheat the atmospheric residue to 380-450°C before sending it to Vacuum Column for separation. The requirement of preheating is to recover heat as well as to achieve the desired vapor and liquid composition below and above the feed tray of the vacuum column. Furnace used here is the fired furnace which uses Methane to Pentane in specific ratios as fuel. The furnace increases the temperature of the stream from 318°C to 404°C before sending it to Vacuum Column. The outlet temperature of furnace is determined based on the coking tendency or thermal reactivity of crude oil and the level of separation required in the column.

3.4 STEAM EJECTORS

Steam ejectors are employed right on top of vacuum column to produce a vacuum inside the column using steam. Steam ejectors basically extract air and non-condensable gases from the column using high-pressure steam. The ejectors are employed at the top of the column rather than at the bottom for safety and energy efficiency purposes. The location at the top allows sufficient vertical space for evacuating the material that is sucked from the column to avoid liquid head buildup which causes more energy to be consumed.

Steam Ejectors are mainly composed of converging nozzle, diffuser throat, diverging nozzle, inlet, and outlet pipes, and condensers inside the assembly to change the phase of steam and other gases and separate them from non-condensable gases.

First of all, high pressure steam, called motive fluid, is passed through convergence and divergence type arrangement which causes the steam pressure to drop and its velocity to increase. This produces low pressure in the chamber. Due to pressure difference, non-condensable gases and hydrocarbons are sucked from the column into the ejector chamber effectively producing a vacuum in the column, which are then carried along with steam into inter condensers which condense the steam and hydrocarbons and send them to another condenser which further condenses the mixture and separates the non-condensable gases and condensate steam. The condensate stream is sent back for reuse using steam traps. This steam ejector assembly effectively produces 0.06 bar of pressure at the column top thus creating the required vacuum in the column.

3.5 VACUUM COLUMN

Our main equipment of focus is the Vacuum Column which is responsible for Asphalt production from FFO separation. The Vacuum Column operates at vacuum condition of 0.06 bar pressure at the top of the column and 0.08 bar

pressure at the bottom. This vacuum is generated using a steam ejector assembly at the top of the column. Due to this pressure difference in the column and the cooling and pumping around of the LVGO and HVGO streams from the side of the column, sufficient temperature gradient is created between the top of the column and the bottom of the column which increases the heat exchange throughout the column thus increasing efficiency of column so that better fractionation can occur without providing too much energy to the column. The temperature and pressure inside the Vacuum Column also depend on the fact that whether steam is used in the column or separation is done without steam introduction as done in dry towers. The current vacuum column also uses steam which is introduced from bottom along with FFO stream for better temperature and pressure conditions inside the column so that duty of the column is reduced, and better separation is achieved. The operating conditions
selected are tabulated below.

Table 1: Vacuum distillation column conditions

Column Parameters	
Top Temperature (^oC)	65.5
Bottom Temperature (^oC)	371.1
Top Pressure (bar)	0.06
Bottom Pressure (bar)	0.08

3.6 HEAT EXCHANGER NETWORK

There are number of heat exchangers being used in the process to cool down the products from Vacuum Column and also some coolers and condensers to assist in the process. The HE-1 is responsible for cooling down a portion of FFO from Atmospheric Bottom so that it can be stored.

The HE-3 cools down the LVGO from Vacuum Column side draw to 65°C. HE-4 is employed to cool down HVGO stream from Vacuum Column down to 86°C. HE-5 is responsible for cooling down Slop Oil from Vacuum Column side draw to 134°C. HE-6 is the last heat exchanger to cool down the vacuum column products. It cools down our desired product Asphalt from Vacuum Column Bottom to 172°C so that it can be used further.

The two pump-around coolers on the side of Vacuum Column are responsible for cooling a fraction of LVGO and HVGO streams and sending them back to vacuum column to increase efficiency of separation.

On the top of Vacuum Column, there are two condensers in the steam ejector assembly to condense the steam and some volatile hydrocarbons from the top of the Vacuum Column. All these heat exchangers make use of a single cooling water stream to reduce utility costs.

CHAPTER 4

MATERIAL BALANCE

4.1 SPLITTER

So, applying the law of conservation of mass, we obtain:

$mass\ flow\ rate\ of\ component\ in = mass\ flow\ rate\ of component\ out$

Since in splitter we have 3 streams, 1st stream is going to Vacuum Column named as **To Vacuum Tower**, 2nd stream one goes to Storage of FFO (Furnace Fuel Oil) named as **FFO*** and the last stream goes to furnace named as **FFO To Furnace**. We have provided split ratios for each stream. Split ratios are:

- **0.9** for To **Vacuum Tower**
- **0.05** for **FFO***
- **0.05** for **FFO To Furnace**

Thus $\dot{m}_1(y_{i1}) = \dot{m}_2(y_{i2}) + \dot{m}_3(y_{i3}) + \dot{m}_4(y_{i4})$

Since $\dot{m}_1$ = 346,909 kg/h, $\dot{m}_2$ = 312,219 kg/h, $\dot{m}_3$ = 17,345 kg/h, $\dot{m}_4$ = 17,345 kg/h

Results are tabulated as follows

IN			OUT			
Components	**Atmospheric Residue**			**To Vacuum Tower**	**FFO***	**FFO To Furnace**
(kg/h)	Yi	Mass flow	yi	Mass flow	Mass flow	**Mass flow**
Methane	0.00	0.00	0.00	0.00	0.00	**0.00**

Ethane	0.00	0.00	0.00	0.00	0.00	**0.00**
Propane	0.00	0.00	0.00	0.00	0.00	**0.00**
i-Butane	0.00	0.00	0.00	0.00	0.00	**0.00**
n-Butane	0.00	0.00	0.00	0.00	0.00	**0.00**
i-Pentane	0.00	0.00	0.00	0.00	0.00	**0.00**
n-Pentane	0.00	0.00	0.00	0.00	0.00	**0.00**
H2O	0.00034	119.3976	0.00034	107.43	5.969	**5.969**
C5-C13 (hydrocarbons)	0.00066	248.989	0.00066	224.09	12.449	**12.449**
C15-C24 (hydrocarbons)	0.17671	61377.379	0.17671	55210.63	3068.8	**3068.8**
C25-C33 (hydrocarbons)	0.3007	104437.9	0.3007	93994.11	5221.8	**5221.8**
C34-C43 (hydrocarbons)	0.3965	137672	0.3965	123904.8	6883.6	**6883.6**
C45-C51 (hydrocarbons)	0.124	43054	0.124	38778.7	2152.6	**2152.6**
O_2	0.00	0.00	0.00	0.00	0.00	**0.00**
N_2	0.00	0.00	0.00	0.00	0.00	**0.00**
CO	0.00	0.00	0.00	0.00	0.00	**0.00**
CO_2	0.00	0.00	0.00	0.00	0.00	**0.00**
TOTAL	-	**346,909**	-	**312,219**	**17,345**	**17,345**
TOTAL IN	**346,909**		**TOTAL OUT**	**346,909**		

Table 2: Splitter

4.2 VAC HEATER MIXER

Introducing the steam stream named as **Vac Steam** along with the stream named as **To Vacuum Tower** in the mixer and then applying the law of conservation
of mass we get:

$$mass\ flow\ rate\ of\ component\ in = mass\ flow\ rate\ of component\ out$$

Thus $\dot{m}2(yi2) + \dot{m}5(yi5) = \dot{m}6(yi6)$

Since $\dot{m}2$ = 312,219 kg/h, $\dot{m}5$= 9072 kg/h, then $\dot{m}6$= 321,291 kg/h

Results are tabulated as follows:

	IN				**OUT**	
COMPONENTS	**TO VAC TOWER**		**VAC COIL STEAM**		**TO VAC HEATER**	
(kg/h)	yi	Mass flow	yi	Mass flow	yi	**Mass flow**
Methane	0.00	0.00	0.00	0.00	0.00	**0.00**
Ethane	0.00	0.00	0.00	0.00	0.00	**0.00**
Propane	0.00	0.00	0.00	0.00	0.00	**0.00**
i-Butane	0.00	0.00	0.00	0.00	0.00	**0.00**
n-Butane	0.00	0.00	0.00	0.00	0.00	**0.00**
i-Pentane	0.00	0.00	0.00	0.00	0.00	**0.00**
n-Pentane	0.00	0.00	0.00	0.00	0.00	**0.00**
H2O	0.00034	107.43	1.00	9072	0.028	**9179.5**
C5-C13 (hydrocarbons)	0.0006	224.09	0.00	0.00	0.0007	**224.9**
C15-C24 (hydrocarbons)	0.1767	55239.63	0.00	0.00	0.17	**55396.4**
C25-C33 (hydrocarbons)	0.3007	93994.11	0.00	0.00	0.29	**93171.5**
C34-C43 (hydrocarbons)	0.3965	123875.8	0.00	0.00	0.38	**124558.4**

C45-C51 (hydrocarbons)	0.124	38778.7	0.00	0.00	0.12	**38760.5**
O_2	0.00	0.00	0.00	0.00	0.00	**0.00**
N_2	0.00	0.00	0.00	0.00	0.00	**0.00**
CO	0.00	0.00	0.00	0.00	0.00	**0.00**
CO_2	0.00	0.00	0.00	0.00	0.00	**0.00**
TOTAL	-	**312,219**	-	**9072**	-	**321,291**
TOTAL IN	**321,291**				**TOTAL OUT**	**321,291**

Table 3: Vac heater mixer

4.3 VACUUM DISTILLATION COLUMN

New steam stream was introduced named as **Vac Column Steam** along with the previous stream coming from preheater which is being provided heat named as **Vac Column Feed**. After entering into the vacuum distillation column, 5 main products are produced using the same procedure as used before (law of conservation of mass).

Applying the law of conservation of mass, we get:

$$mass\ flow\ rate\ of\ component\ in = mass\ flow\ rate\ of\ component\ out$$

Since $\dot{m}_7$ = 321,291 kg/h and $\dot{m}_8$ = 9072 kg/h then output products will be $\dot{m}_9$ = 19,356 kg/h, $\dot{m}_{10}$ = 12,796 kg/h, $\dot{m}_{11}$ = 34,046 kg/h, $\dot{m}_{12}$ = 1094 kg/h and $\dot{m}_{13}$ = 263,080 kg/h respectively.

$$\dot{m}_7(y_{i7}) + \dot{m}_8(y_{i8}) = \dot{m}_9(y_{i9}) + \dot{m}_{10}(y_{i10}) + \dot{m}_{11}(y_{11}) + \dot{m}_{12}(y_{i12}) + \dot{m}_{13}(y_{i13})$$

The results are tabulated as follows:

	IN				OUT	
COMPONENTS	**VAC COLUMN FEED**		**VAC COLUMN STEAM**		**VAC COLUMN OVERHEAD**	
(kg/h)	yi	Mass flow	yi	Mass flow	yi	**Mass flow**
Methane	0.00	0.00	0.00	0.00	0.00	**0.00**
Ethane	0.00	0.00	0.00	0.00	0.00	**0.00**
Propane	0.00	0.00	0.00	0.00	0.00	**0.00**
i-Butane	0.00	0.00	0.00	0.00	0.00	**0.00**
n-Butane	0.00	0.00	0.00	0.00	0.00	**0.00**
i-Pentane	0.00	0.00	0.00	0.00	0.00	**0.00**
n-Pentane	0.00	0.00	0.00	0.00	0.00	**0.00**
H2O	0.028	9179.5	1.00	9072	0.94	**18246.7**
C5-C13 (hydrocarbons)	0.0006	224.9	0.00	0.00	0.01	**207.3**
C15-C24 (hydrocarbons)	0.17	55396.4	0.00	0.00	0.04	**901.8**
C25-C33 (hydrocarbons)	0.29	93171.5	0.00	0.00	0.00	**0.00**
C34-C43 (hydrocarbons)	0.38	124558.4	0.00	0.00	0.00	**0.00**
C45-C51 (hydrocarbons)	0.12	38760.5	0.00	0.00	0.00	**0.00**
O_2	0.00	0.00	0.00	0.00	0.00	**0.00**
N_2	0.00	0.00	0.00	0.00	0.00	**0.00**
CO	0.00	0.00	0.00	0.00	0.00	**0.00**
CO_2	0.00	0.00	0.00	0.00	0.00	**0.00**
TOTAL	**-**	**321,291**	**-**	**9072**	**-**	**19,356**

Table 4: Vacuum distillation column (i)

	OUT					
COMPONENTS	**LVGO***		**HVGO***		**SLOP WAX**	
(kg/h)	Yi	Mass flow	Yi	Mass flow	yi	**Mass flow**
Methane	0.00	0.00	0.00	0.00	0.00	**0.00**
Ethane	0.00	0.00	0.00	0.00	0.00	**0.00**
Propane	0.00	0.00	0.00	0.00	0.00	**0.00**
i-Butane	0.00	0.00	0.00	0.00	0.00	**0.00**
n-Butane	0.00	0.00	0.00	0.00	0.00	**0.00**
i-Pentane	0.00	0.00	0.00	0.00	0.00	**0.00**
n-Pentane	0.00	0.00	0.00	0.00	0.00	**0.00**
H2O	0.00	0.00	0.00	0.00	0.00	**0.00**
C5-C13 (hydrocarbons)	0.0012	16.07	0.00	0.00	0.00	**0.00**
C15-C24 (hydrocarbons)	0.98	12630.9	0.89	30525	0.38	**410.7**
C25-C33 (hydrocarbons)	0.01	148.6	0.1	3510.3	0.53	**590.6**
C34-C43 (hydrocarbons)	0.00	0.00	0.00027	9.2	0.09	**92.4**
C45-C51 (hydrocarbons)	0.00	0.00	0.00	0.00	0.00	**0.00**
O_2	0.00	0.00	0.00	0.00	0.00	**0.00**
N_2	0.00	0.00	0.00	0.00	0.00	**0.00**
CO	0.00	0.00	0.00	0.00	0.00	**0.00**
CO_2	0.00	0.00	0.00	0.00	0.00	**0.00**
TOTAL	**-**	**12,796**	**-**	**34,039**	**-**	**1094**

Table 5: Vacuum distillation column (ii)

IN	OUT	
COMPONENTS	**Asphalt***	
(kg/h)	yi	**Mass flow**
Methane	0.00	**0.00**
Ethane	0.00	**0.00**
Propane	0.00	**0.00**
i-Butane	0.00	**0.00**
n-Butane	0.00	**0.00**
i-Pentane	0.00	**0.00**
n-Pentane	0.00	**0.00**
H2O	0.00001	**2.89**
C5-C13 (hydrocarbons)	0.00	**0.00**
C15-C24 (hydrocarbons)	0.04	**10729.9**
C25-C33 (hydrocarbons)	0.34	**89756.9**
C34-C43 (hydrocarbons)	0.47	**123785.5**
C45-C51 (hydrocarbons)	0.147	**38795.9**
O_2	0.00	**0.00**
N_2	0.00	**0.00**
CO	0.00	**0.00**
CO_2	0.00	**0.00**
TOTAL	**-**	**263,071**

TOTAL IN = 330,363 kg/h
TOTAL OUT = 330,363 kg/h

Table 6: Vacuum distillation column (iii)

CHAPTER 5

ENERGY BALANCE

5.1 PUMPS

Six main pumps were utilized for this process:

Water supply pump to the product heat exchangers network,

The equations required were:

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

Important assumptions are being made;

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

$$\eta Q_{p1} = \dot{m} * \frac{\Delta P}{\rho}$$

Required data and results tabulated as follows:

Pump		
Pump	P-FFO	**P-VCOH**
Temperature (^{o}C)	15.0	**15.0**
Specific Enthalpy (kJ/kg)	-15930	**-15930**
Inlet Pressure (bar_g)	3.417E-03	**3.417E-03**
Outlet Pressure (bar_g)	0	**0**
Density (kg/m^3)	1015	**1015**
Mass Flow Rate (kg/h)	1802	**1802**
Pump Efficiency	0.75	**0.75**
Pump Duty (kW)	**2.247**	**2.247**

Table 7: Pumps (i)

Pump		
Pump	P-SO	**P-ASP**
Temperature (^{o}C)	15.0	**15.0**
Specific Enthalpy (kJ/kg)	-15930	**-15930**
Inlet Pressure (bar_g)	3.417E-03	**3.417E-03**
Outlet Pressure (bar_g)	0	**0**
Density (kg/m^{3})	1015	**1015**
Mass Flow Rate (kg/h)	1802	**1802**
Pump Efficiency	0.75	**0.75**
Pump Duty (kW)	**2.247**	**2.247**

Table 8: Pumps (ii)

Table 9: Pumps (iii)

Pump		
Pump	P-LVGO	**P-HVGO**
Temperature (^{o}C)	15.0	**15.0**
Specific Enthalpy (kJ/kg)	-15930	**-15930**
Inlet Pressure (bar_g)	3.417E-03	**3.417E-03**
Outlet Pressure (bar_g)	0	**0**
Density (kg/m^{3})	1015	**1015**
Mass Flow Rate (kg/h)	1802	**1802**
Pump Efficiency	0.75	**0.75**
Pump Duty (kW)	**2.247**	**2.247**

5.2 PREHEATER

There is only one main preheater within the proposed process, and it is responsible for providing the heat to the FFO (Furnace Fuel Oil) and Vacuum Steam mixture to our main equipment which is Vacuum Distillation Column.

The basic equation in use for the following case is:

$$Q = \dot{m}\; c_p \Delta T \text{ (SF)}$$

Where SF is the safety factor usually varies from 1 to 2.

Stream		To Vac Heater	Air	Fuel	Flue Gas	**Vac Column Feed**
Vapor Fraction	Units	0.4649	1.000	1.000	1.000	**0.5820**
Temperature	*^{o}C*	318.6	25.0	25.0	537.6	**403.8**
Pressure	*psi*	36.7	14.5	14.5	14.5	**36.7**
Mass Flow	*kg/hr*	321291	109631	83808	193439	**321291**
Heat Flow	*kJ/h*	-5.7E+08	-3.06E+04	-3.71E+08	-4.68E+08	**-4.81E+08**
Mass Heat Capacity	*kJ/kg-^{o}C*	2.834	1.012	1.778	2.277	**3.024**
Q	***kW***	**24877**				

Table 10: Preheater

5.3 HEAT EXCHANGERS

To obtain the heat loads involved for every heat exchanger used in the proposed process, we performed material balance.

The basic equation in use for all the following cases is:

$$Q = \dot{m}\; c_p \Delta T$$

Table 11: HE-1

		Shell side		Tube side	
Stream		To HE-1	From HE-1	FFO*	**FFO To Storage**
Vapor Fraction	Units	0.000	0.000	0.000	**0.000**
Temperature	^oC	15.00	36.00	323.50	**320.000**
Pressure	*psi*	14.7	9.4	36.7	**45.00**
Mass Flow	*kg/hr*	1802	1802	17345	**17345**
Heat Flow	*kJ/h*	-2.87E+07	-2.85E+07	-2.51E+07	**-2.53E+07**
Mass Heat Capacity	*kJ/kg-^oC*	4.314	4.314	2.889	**2.877**
Q	***kW***	**45.3**		**-45.3**	

		Shell side		Tube side	
Stream		To HE-2	From HE-2	Vac Column Overhead	**VCOH**

Vapor Fraction	Units	0.000	0.000	1.000	**0.9995**
Temperature	*oC*	15.00	45.55	65.56	**60.00**
Pressure	*Psi*	14.7	7.5	0.9659	**0.9652**
Mass Flow	*kg/hr*	1802	1802	19356	**19356**
Heat Flow	*kJ/h*	-2.87E+07	-2.84E+07	-2.45E+08	**-2.45E+08**
Mass Heat Capacity	*kJ/kg-oC*	4.316	4.313	1.868	**1.867**
Q	***kW***	**66.1**		**-66.1**	

Table 12: HE-2

		Shell side		**Tube side**	
Stream		To HE-3	From HE-3	LVGO*	**LVGO**
Vapor Fraction	Units	0.000	0.4665	0.000	**0.000**
Temperature	*oC*	15.00	97.19	181.4	**95.00**
Pressure	*Psi*	14.7	13.3	1.019	**8.50**
Mass Flow	*kg/hr*	1802	1802	12796	**12796**
Heat Flow	*kJ/h*	-2.87E+07	-2.60E+07	-2.29E+07	**-2.55E+07**
Mass Heat Capacity	*kJ/kg-oC*	4.316	2.973	2.505	**2.137**
Q	***Kw***	**728**		**-728**	

Table 13: HE-3

		Shell side		**Tube side**	
Stream		To HE-4	From HE-4	HVGO*	**HVGO**
Vapor Fraction	Units	0.000	0.000	0.050	**0.000**
Temperature	*oC*	15.00	72.00	221.5	**216.5**
Pressure	*psi*	14.7	7.44	1.09	**10.00**
Mass Flow	*kg/hr*	1802	1802	34046	**34046**
Heat Flow	*kJ/h*	-2.87E+07	-2.82E+07	-5.75E+07	**-5.8E+07**
Mass Heat Capacity	*kJ/kg-oC*	4.314	4.346	2.637	**2.612**
Q	***Kw***	**123**		**-123**	

Table 14: HE-4

		Shell side		**Tube side**	
Stream		To HE-5	From HE-5	Slop Wax	**Slop Oil**
Vapor Fraction	Units	0.000	0.000	0.0002	**0.000**
Temperature	*oC*	15.00	70.00	255.3	**94.15**
Pressure	*psi*	14.7	13.5	1.143	**10.00**
Mass Flow	*kg/hr*	1802	1802	1094	**1094**
Heat Flow	*kJ/h*	-2.87E+07	-2.82E+07	-1.75E+06	**-2.18E+06**

Mass Heat Capacity	*kJ/kg-oC*	4.316	4.317	2.796	**2.510**
Q	***kW***	**62.6**		**-62.6**	

Table 15: HE-5

		Shell side		**Tube side**	
Stream		To HE-6	From He-6	Asphalt*	**Asphalt**
Vapor Fraction	Units	0.000	1.000	0.0001	**0.000**
Temperature	*oC*	15.00	95.00	277.5	**270.8**
Pressure	*Psi*	14.7	8.00	1.196	**8.00**
Mass Flow	*kg/hr*	1802	1802	263071	**263071**
Heat Flow	*kJ/h*	-2.87E+07	-2.40E+07	-4.16E+08	**-4.2E+08**
Mass Heat Capacity	*kJ/kg-oC*	4.316	1.891	2.725	**2.702**
Q	***kW***	**1321**		**-1321**	

Table 16: HE-6

5.4 VACUUM DISTILLATION COLUMN

To obtain the power to be provided to the column for the separation, difference of heat given and heat removed will give us the separation power. Since pumparounds are also included in the proposed design so we also will have to get the power consumption of the tower only. Vacuum is being created by steam ejectors (3 are connected) while consuming the same power.

The results are tabulated as follows:

Property Table			
IN		**OUT**	
Stream	**Heat Flow (MJ/h)**	**Stream**	**Heat Flow (MJ/h)**
Vac	117925	Vac Column	**22956**

Column Feed		Overhead	
Vac Column Steam	459800	LVGO*	**245539**
-	-	HVGO*	**57512**
-	-	Slop Wax	**415256**
-	-	Asphalt*	**1724**
-	-	-	**-**
-	-	-	**-**
TOTAL IN	**577,726**	**TOTAL OUT**	**742,988**

Table 17: Vacuum distillation column (iv)

Property	**Property**
Steam Ejector Consumption Power (For each steam ejector) (MJ/h)	**Column Separation Power (MJ/h)**
352	**165,182**

Table 18: Vacuum distillation column (v)

Sub-equipment	**Property**
Pumparounds	**Heat Flow (MJ/h)**
Top Stage Pumparound Cooler	**136186**
HVGO Pumparound Cooler	**28020**
TOTAL	**164,206**

Table 19: Vacuum distillation column (vi)

CHAPTER 6

EQUIPMENT DESIGN

6.1 HEAT EXCHANGER

6.1.1 LVGO HEAT EXCHANGER

The following calculations are done to design the LVGO heat exchanger which is obtained from vacuum distillation column and then is passed through heat exchanger for specific temperature and then goes to storage.

Tube Side

Mass flow of hot fluid = 12796 kg/h. Inlet and outlet temperatures of hot fluid are 183.3 °C and 95 °C respectively.

Shell Side

Mass flow of cold fluid = 1802 kg/h. Inlet and outlet temperatures of cold fluid are 15 °C and 97.19 °C respectively.

Specifications

- Tube outer diameter = 0.75 in.
- Pitch = 1 in. square
- Type of tubes = 15 BWG
- Length of a tube = 20 ft.
- No. of tubes = 150
- No. of passes (Tube side) = 6
- Shell inner diameter = 17 ¼ in.
- Baffle spacing = 5 in.
- No. of passes (Shell side) = 47

1) Heat Balance

$$Q = m\ Cp\ \Delta T = \ 2561000\ kJ/h = 2427000\ Btu/h$$

2) LMTD Calculation

Hot fluid		Cold fluid	Difference
181.4	Higher Temperature	97.19	**86.11**
95	Lower Temperature	15.0	**80.0**
86.4	**Difference**	**82.19**	**4.21**

LMTD = $\frac{\Delta T1 - \Delta T2}{\ln\left(\frac{\Delta T1}{\Delta T2}\right)}$ = 82.069 °C = 147.7 °F

R = $\frac{T1 - T2}{T2 - T1}$ = 2.8

S = $\frac{t2 - t1}{T1 - t1}$ = 0.17

F_T = 0.789

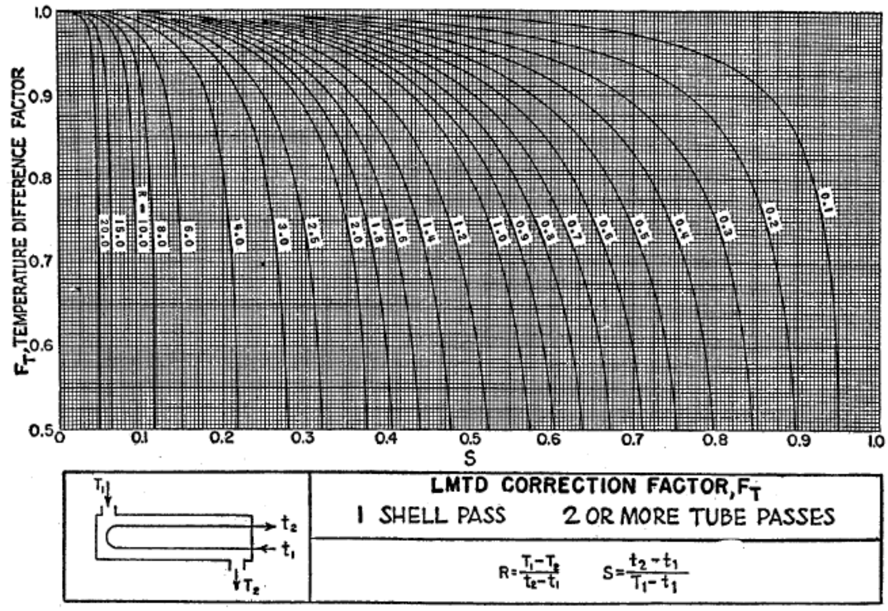


Figure 17: F_T factor graph

$\Delta t = F_T \times \mathrm{LMTD} = 64.75\ °\mathrm{C} = 116.5\ °\mathrm{F}$

3) Calorific Temperature

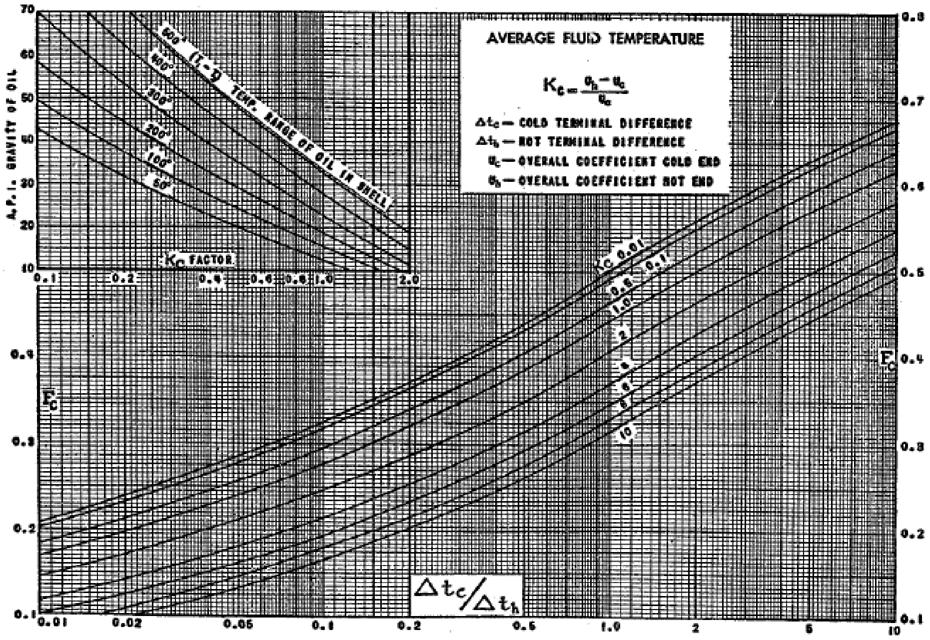


Figure 18: Calorific temperature graph

$\frac{\Delta tc}{\Delta th} = 0.92, \qquad K_c = 0.005,\ F_c = 0.8$

$Tc = T2 + Fc \times (T2 - T1) = 89.76\ °\mathrm{C} = 193.56\ °\mathrm{F}$

$tc = t2 + Fc \times (t2 - t1) = 80.7\ °\mathrm{C} = 177.2\ °\mathrm{F}$

1) SHELL SIDE CALCULATION

4) Flow Area

$a_s = \frac{\mathrm{I.D} * C'' * B}{144 * \mathrm{Pr}}$

$a_s = 0.149\ ft^2$

5) Mass Velocity

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

$G_s = 26657\ {}^{lb}/_{ft^2.hr}$

6) Reynolds Number

At $T_c = 322.8$ °F

$\mu = 1.032\ {}^{lb}/_{ft.hr}$

$D_e = 0.079 ft$

$Re_s = \frac{De*Gs}{\mu}$

$Re_s = 2039.6$

7) j_H Factor

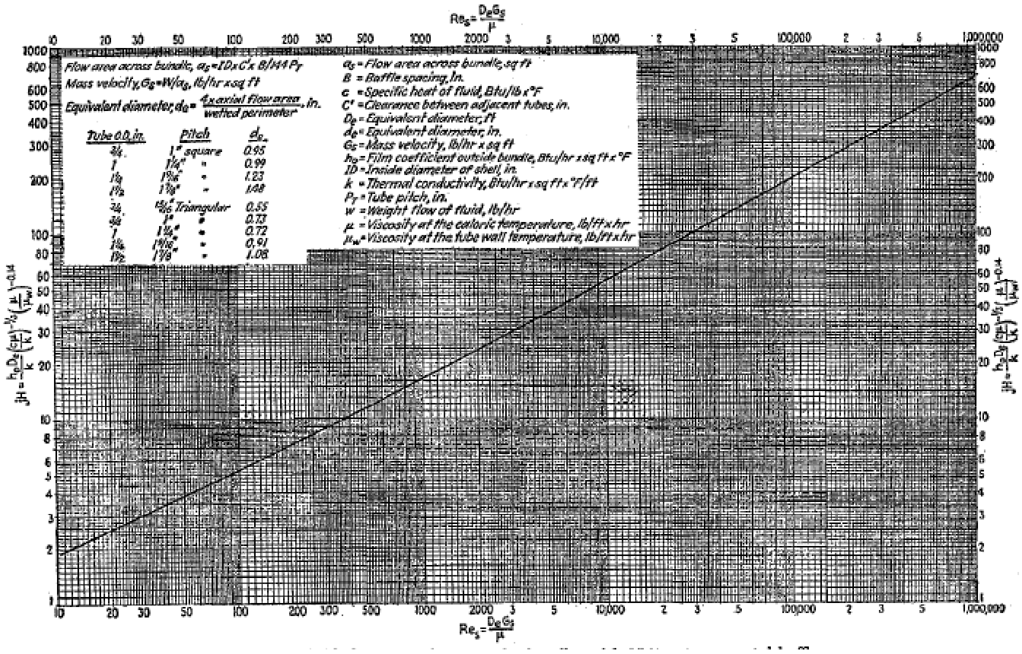


Figure 19: Reynolds number graph versus jH factor (for shell side)

$j_H = 30.4$

8) $(c*\mu/k)^{1/3}$

At $T_c = 193.56$ °F

$c = 0.521 \, {}^{Btu}/_{lb.}$°F

$k = 0.224 \, {}^{Btu}/_{hr.ft.}$°F

$(c*\mu/k)^{1/3} = 1.33$

9) Heat Transfer Coefficient

$(h_0/Ø_s) = (j_H * k / D_e) \, (c*\mu/k)^{1/3}$

$(h_0/Ø_s) = 114.6$

10) Tube Wall Temperature

$t_w = t_c + ((h_0/Ø_s)/ ((h_0/Ø_s) + (h_i/Ø_t)) * (T_c - t_c)$

$t_w = 180.1$°F

11) $(\mu / \mu_w)^{0.14}$

At $t_w = 155.0$°F

$\mu_w = 0.734\ ^{lb}/_{ft.hr}$

$Ø_s = (\mu / \mu_w)^{0.14}$

$Ø_s = 1.476$

12) Corrected Coefficient

$h_0 = (h_0/Ø_s) * Ø_s$

$h_0 = 168.4\ ^{Btu}/_{hr.ft^2.°F}$

2) TUBE SIDE CALCULATION

4) Flow Area

$a_t' = 0.289\ ft^2$

$a_t = \frac{N*at'}{144*n}$

$a_t = 0.15\ ft^2$

5) Mass Velocity

$G_t = \frac{W}{at}$

$G_t = 188066\ ^{lb}/_{ft^2.hr}$

6) Reynolds Number

At $t_c = 101.6$ °F

$\mu = 0.6377\ ^{lb}/_{ft.hr}$

$D_e = 0.071 ft$

$Re_t = \frac{De*Gs}{\mu}$

$Re_t = 20938.8$

7) j_H Factor

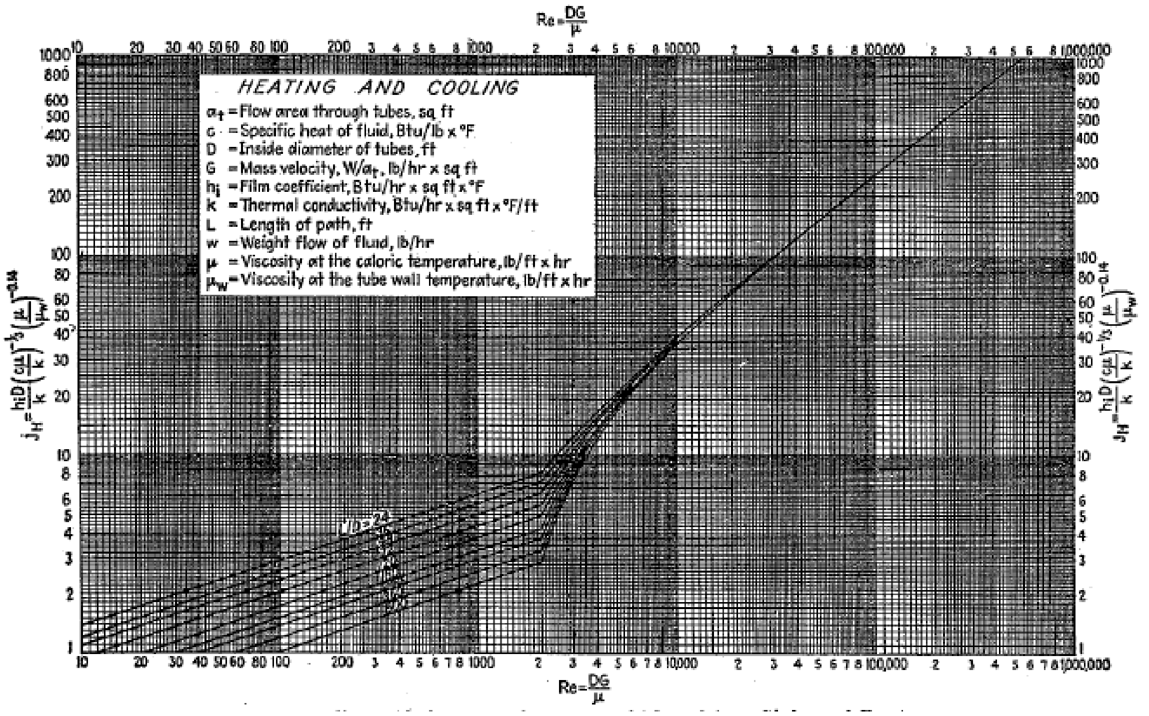


Figure 20: Reynolds number graph versus jH factor (for tube side)

$j_H = 135$

L/D = 281.6

8) $(c*\mu/k)^{1/3}$

At t_c = 177.2 °F

$$c = 0.2143\, {}^{Btu}/_{lb.°F}$$

$$k = 0.363\, {}^{Btu}/_{hr.ft.°F}$$

$(c*\mu/k)^{1/3} = 0.72$

9) Heat Transfer Coefficient

$(h_i/\O_i) = (j_H * k / D_e)\,(c*\mu/k)^{1/3}$

$(h_i/Ø_i) = 500$

$(h_{io}/Ø_t) = (h_i/Ø_i) * (ID/OD)$

$(h_{io}/Ø_t) = 430.2$

10) $(\mu / \mu_w)^{0.14}$

At $t_w = 180.1°F$

$\mu_w = 0.768\ {}^{lb}/_{ft.hr}$

$Ø_s = (\mu / \mu_w)^{0.14}$

$Ø_s = 0.83$

11) Corrected Coefficient

$h_{io} = (h_{io}/Ø_t) * Ø_t$

$h_{io} = 353.02\ {}^{Btu}/_{hr.ft^2.°F}$

1) Clean Overall Coefficient

$U_c = (hi_0 * h_0) / hi_0 + h_0$

$U_c = 114.3\ {}^{Btu}/_{hr.ft^2.°F}$

2) Design Overall Coefficient

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

Total Surface Area A = N * L * a'' = 588.3 ft^2

$U_d = (Q / A.\Delta T) = 23.12\ {}^{Btu}/_{hr.ft^2.°F}$

3) Dirt Factor

$R_d = (U_c - U_d) / (U_c * U_d)$

$R_d = 0.034\ hr.ft^2.°F\ /\ ^{Btu}$

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

4) Shell Side Pressure Drop

Following calculations are carried to find out pressure drop across shell side of heat exchanger.

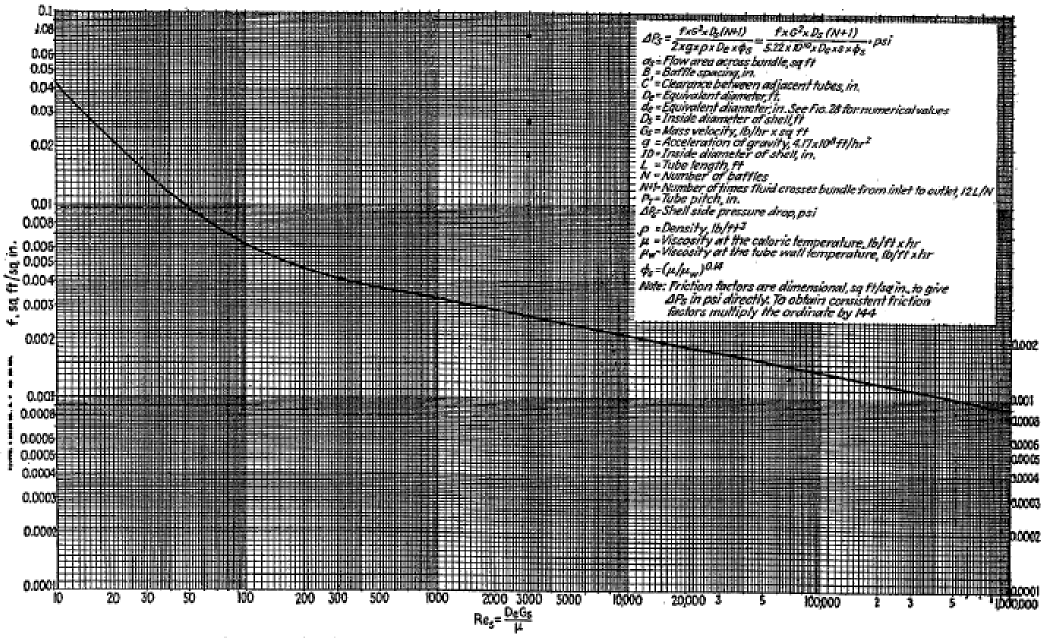


Figure 21: Reynolds number versus f factor graph (for shell side)

$Re_s = 2039.1$

$f = 0.17$

$D_S = 1.43 ft$

$s = 1.015$

N+1 = 12 * (L/B)

N+1 = 48

$\Delta P_S = (f * (G_S)^2 * D_s * (N+1)) / (5.22*10^{10} * D_e * s * Ø_s)$

$\Delta P_S = 1.3$ psi

5) Tube Side Pressure Drop

Following calculations are carried to find out pressure drop across tube side of heat exchanger.

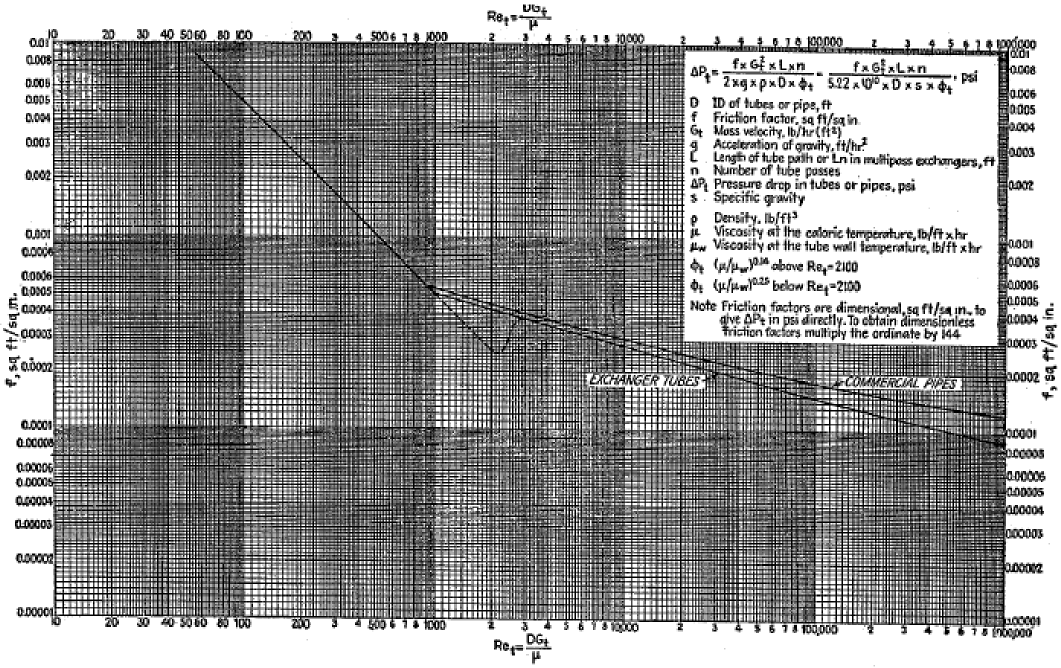


Figure 22: Reynolds number versus f factor graph (for tube side)

$Re_s = 20938.8$

f = 0.009

$s = 0.767$

$\Delta P_t = (f * (G_t)^2 * L * n) / (5.22*10^{10} * D * s * Ø_t)$

$\Delta P_t = 5.39$ psi

$G_t = 188066\ {}^{lb}/_{ft^2.\,hr}$

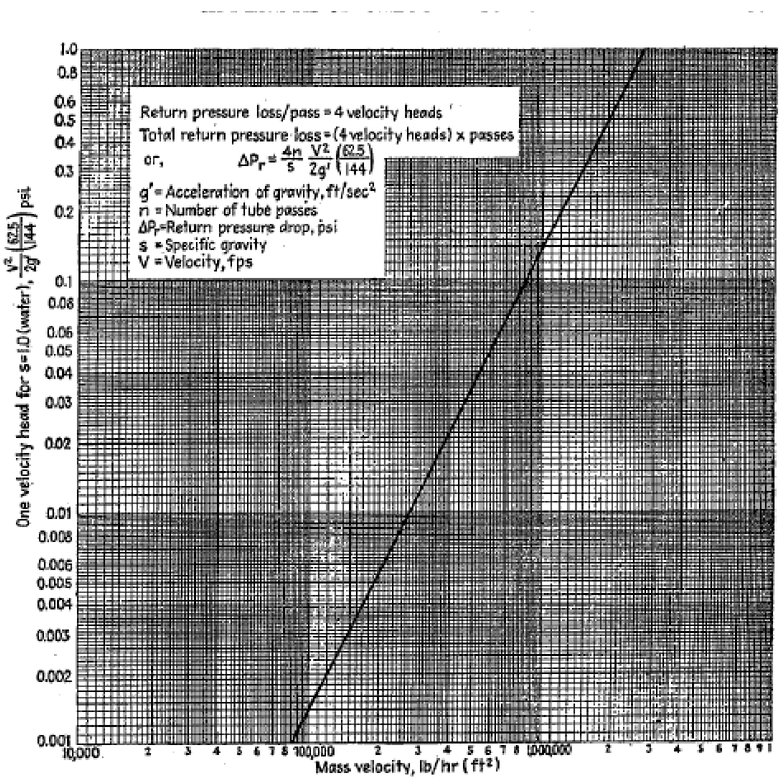


Figure 23: Mass velocity graph versus ($v^2/2g'$)

$(v^2/2g') = 0.191$

$\Delta P_r = (4n*v^2/ s*2g')$

$\Delta P_r = 1.98$ psi

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

$\Delta P_T = 7.4$ psi [9, 10]

6.2 VACUUM DISTILLATION COLUMN

1) Light Key and Heavy Key

Light key (LK) : NBP [0]114

Heavy Key (HK) : NBP [0]215

2) Type of tray

Sieve tray

3) Calculation of relative volatility

$$\alpha_i = \frac{K_i}{K_{HK}}$$

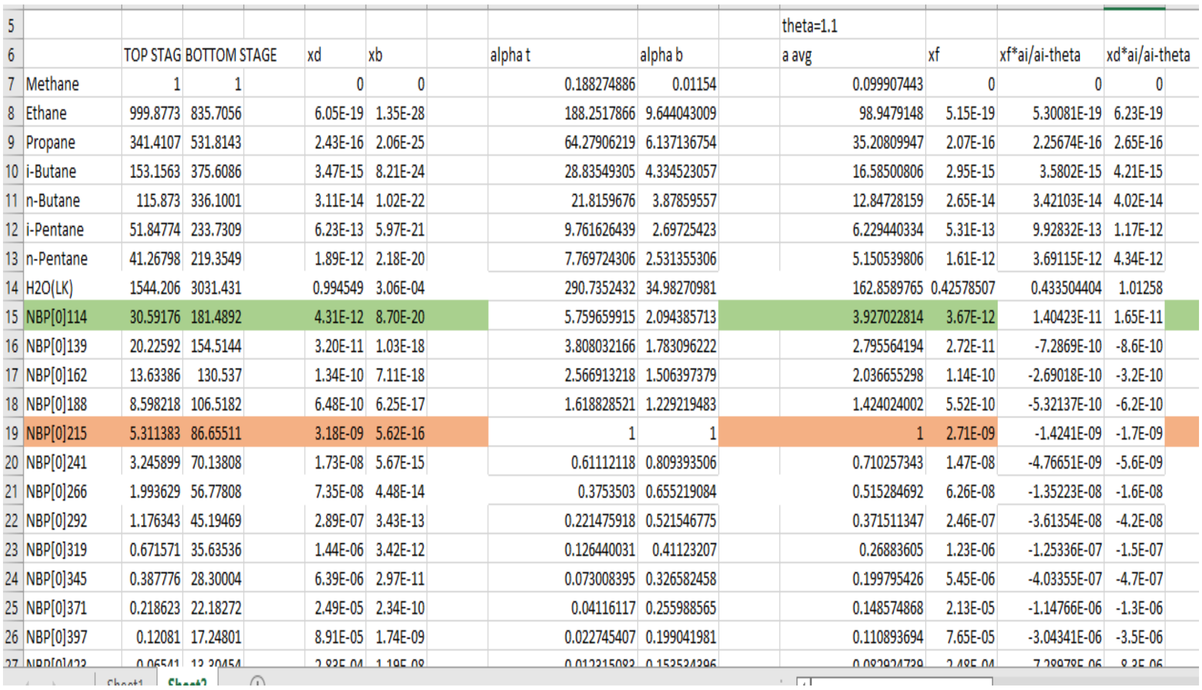
$$\alpha_{avg} = \frac{\alpha_{top} + \alpha_{bottom}}{2}$$

Where:

K_i value of component i

K_{HK} value of heavy component α from figure given below

(Calculations are made by using an Excel sheet).



5								theta=1.1			
6		TOP STAG	BOTTOM STAGE	xd	xb	alpha t	alpha b	a avg	xf	xf*ai/ai-theta	xd*ai/ai-theta
7	Methane	1	1	0	0	0.188274886	0.01154	0.099907443	0	0	0
8	Ethane	999.8773	835.7056	6.05E-19	1.35E-28	188.2517866	9.644043009	98.9479148	5.15E-19	5.30081E-19	6.23E-19
9	Propane	341.4107	531.8143	2.43E-16	2.06E-25	64.27906219	6.137136754	35.20809947	2.07E-16	2.25674E-16	2.65E-16
10	i-Butane	153.1563	375.6086	3.47E-15	8.21E-24	28.83549305	4.334523057	16.58500806	2.95E-15	3.5802E-15	4.21E-15
11	n-Butane	115.873	336.1001	3.11E-14	1.02E-22	21.8159676	3.87859557	12.84728159	2.65E-14	3.42103E-14	4.02E-14
12	i-Pentane	51.84774	233.7309	6.23E-13	5.97E-21	9.761626439	2.69725423	6.229440334	5.31E-13	9.92832E-13	1.17E-12
13	n-Pentane	41.26798	219.3549	1.89E-12	2.18E-20	7.769724306	2.531355306	5.150539806	1.61E-12	3.69115E-12	4.34E-12
14	H2O(LK)	1544.206	3031.431	0.994549	3.06E-04	290.7352432	34.98270981	162.8589765	0.42578507	0.433504404	1.01258
15	NBP[0]114	30.59176	181.4892	4.31E-12	8.70E-20	5.759659915	2.094385713	3.927022814	3.67E-12	1.40423E-11	1.65E-11
16	NBP[0]139	20.22592	154.5144	3.20E-11	1.03E-18	3.808032166	1.783096222	2.795564194	2.72E-11	-7.2869E-10	-8.6E-10
17	NBP[0]162	13.63386	130.537	1.34E-10	7.11E-18	2.566913218	1.506397379	2.036655298	1.14E-10	-2.69018E-10	-3.2E-10
18	NBP[0]188	8.598218	106.5182	6.48E-10	6.25E-17	1.618828521	1.229219483	1.424024002	5.52E-10	-5.32137E-10	-6.2E-10
19	NBP[0]215	5.311383	86.65511	3.18E-09	5.62E-16	1	1	1	2.71E-09	-1.4241E-09	-1.7E-09
20	NBP[0]241	3.245899	70.13808	1.73E-08	5.67E-15	0.61112118	0.809393506	0.710257343	1.47E-08	-4.76651E-09	-5.6E-09
21	NBP[0]266	1.993629	56.77808	7.35E-08	4.48E-14	0.3753503	0.655219084	0.515284692	6.26E-08	-1.35223E-08	-1.6E-08
22	NBP[0]292	1.176343	45.19469	2.89E-07	3.43E-13	0.221475918	0.52154677 5	0.371511347	2.46E-07	-3.61354E-08	-4.2E-08
23	NBP[0]319	0.671571	35.63536	1.44E-06	3.42E-12	0.126440031	0.41123207	0.26883605	1.23E-06	-1.25336E-07	-1.5E-07
24	NBP[0]345	0.387776	28.30004	6.39E-06	2.97E-11	0.073008395	0.326582458	0.199795426	5.45E-06	-4.03355E-07	-4.7E-07
25	NBP[0]371	0.218623	22.18272	2.49E-05	2.34E-10	0.04116117	0.255988565	0.148574868	2.13E-05	-1.14766E-06	-1.3E-06
26	NBP[0]397	0.12081	17.24801	8.91E-05	1.74E-09	0.022745407	0.199041981	0.110893694	7.65E-05	-3.04341E-06	-3.5E-06

Figure 24: Excel sheet calculation for distillation column

4) Determination of minimum Reflux ratio R_m

$$\sum \frac{\alpha_i x_{i,d}}{\alpha_i - \theta} = R_m + 1 \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots . (1)$$

Where:

α = average Relative volatility of any component.

x = mole fraction of component.

θ = constant.

R_m= minimum Reflux ratio

$$\sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 1 - q \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots .. \ldots\ldots . (2)$$

Where:

x_{if}=mole fraction of component in feed.

q = Feed quality.

$$q = \frac{H_G - H_F}{H_G - H_L} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (3)$$

Where:

H_G= Enthalpy of gas at the feed (KJ/Kmol)

H_L= Enthalpy of liquid at the feed (KJ/Kmol)

H_F= Enthalpy of feed (KJ/Kmol)

By putting values in eq (3)

$$q = \frac{(-248123.8) - (-401692)}{(-248123.8) - (-615508)} = 0.418$$

Substitute in equation (2) to find θ:

$$\sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 1 - 0.418$$

$$\sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 0.58$$

From excel sheet, value of θ = 2.9. Substitute this value in equation (1) to find . From excel sheet:

$$\sum \frac{\alpha_i x_{i,d}}{\alpha_i - \theta} = 0.0935 = R_m + 1$$

$$\boldsymbol{R_m = 0.9065}$$

5) Calculation of Actual Reflux Ratio (R)

The rule of thumb is:

R = (1.2 -1.5) $\boldsymbol{R_m}$

R =1.35* 0.9065

R = 1.22

6) Calculation of Minimum number of stages $\boldsymbol{N_{min}}$

$$N_{min} = \frac{\log\left[\frac{x_{LK}}{x_{HK}}\right]_d \log\left[\frac{x_{HK}}{x_{LK}}\right]_b}{\log \alpha_{LK}}$$

Where:

x_{LK}= mole fraction of light key.

x_{HK}= mole fraction of heavy key.

α_{LK}= average relative volatility of light key.

$$\boldsymbol{N_{min} = 4.042\ stages}$$

7) Calculation of theoretical number of stages

$$\frac{N - N_{min}}{N+1} = 0.75\left[1 - \left(\frac{R - R_{min}}{R+1}\right)^{0.566}\right] = 0.75[1 - \left(\frac{(1.22 - 0.9065)}{(1.22 + 1)}\right)^{0.566}]$$

$$\boldsymbol{N = 9.09\ stages}$$

8) Calculation of the column efficiency $(\boldsymbol{E_O})$

$$E_O = 51 - 32.5\log(\mu_a \alpha_a) \dots\dots\dots\dots\dots (4)$$

Where:

μ_a= the molar average liquid viscosity = 0.092 mNs/m^2

α_a= average relative volatility of the light key = 3.92 mNs/m^2

By using eq (4)

$$E_O = 51 - 32.5\log(0.092 \times 3.92)$$

$$E_O = 65\ \%$$

9) Calculation of actual number of stages

$$N_a = \frac{N}{E_O}$$

$$= \frac{9.09}{0.65} = \mathbf{13.98}\ \boldsymbol{stages}$$

10) Calculation of column Height ($\boldsymbol{H_t}$)

$$H_t = \left[N_a \times C + \frac{(N_a - 1)}{10}\right] + (0.2 \times H_t)$$

$$H_t = \left[14 \times 0.73 + \frac{(14 - 1)}{10}\right] + (0.2 \times H_t)$$

$$\boldsymbol{H_t = 14.4\ m}$$

11) Calculation of the feed plate location

$$\log\frac{N_r}{N_s} = 0.206 \log\left[\left(\frac{B}{D}\right)\left(\frac{x_{f,HK}}{x_{f,LK}}\right)\left(\frac{x_{b,LK}}{x_{d,HK}}\right)^2\right]$$

Where:

N_r= number of stages above the feed

N_s= number of stages below the feed

B = molar flow bottom product

$x_{f,HK}$ = concentration of the heavy key in the feed

$x_{f,LK}$ = concentration of the light key in the feed

$x_{b,LK}$= concentration of the light key if in the bottom product

$x_{d,HK}$ = concentration of the heavy key in the top product

$$\log\frac{N_r}{N_s} = 0.206\log\left[\left(\frac{510.6}{1018}\right)\left(\frac{2.71\times\ 10^{-9}}{3.67\times\ 10^{-12}}\right)\left(\frac{8.7\times\ 10^{-20}}{3.18\times\ 10^{-9}}\right)^2\right] = 0.3$$

$$\frac{N_r}{N_s} = 1.99\ \ldots.(3)$$

As we know total number of plates are 14, so,

$$N_r + N_s = 14\ \ldots\ldots(4)$$

Solving above equations simultaneously

$N_r = 9.3$

$$\boldsymbol{N_s = 4.68}$$

The feed enters the column at tray no 5 from the bottom.

12) Calculation of diameter of the column

The following areas terms are use in the design:

A_c= total column cross sectional area

A_d= cross sectional area of down comer

A_n= net area available for vapour-liquid disengagement, normally equal to $A_c - A_d$

13) For a single pass plate

A_a= active area, equal to $A_c - 2A_d$for single-pass plates

A_h= hole area, the total area of all the active holes

A_P =perforated area (including blanked areas),

A_{ap}= the clearance area under the downcomer apron top diameter calculation

$$U_f = K\sqrt{\frac{\rho_L - \rho_V}{\rho_V}}\ \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots..(5)$$

Where:

U_f= flooding vapor velocity m/s, based on the net column cross-sectional area A_n

K = constant obtained from graph

$$F_{LV} = \frac{L_w}{V_w}\sqrt{\frac{\rho_V}{\rho_L}} \dots\dots\dots\dots\dots\dots..(6)$$

Where:

F_{LV} = The liquid-vapor flow factor

L_w= liquid mass flow rate, kg/s,

V_w = vapor mass flow rate, kg/s.

Top diameter calculations:

$$\frac{L_w}{V_w} = \frac{R}{R+1} = \frac{1.22}{1.22+1} = 0.37$$

Values of densities are:

$$\rho_V = 4.46\frac{kg}{m^3}$$

$$\rho_L = 716.4\frac{kg}{m^3}$$

By putting values in eq (6)

$$F_{LV} = 0.37\sqrt{\frac{4.46}{716.4}} = 0.029$$

From figure given below

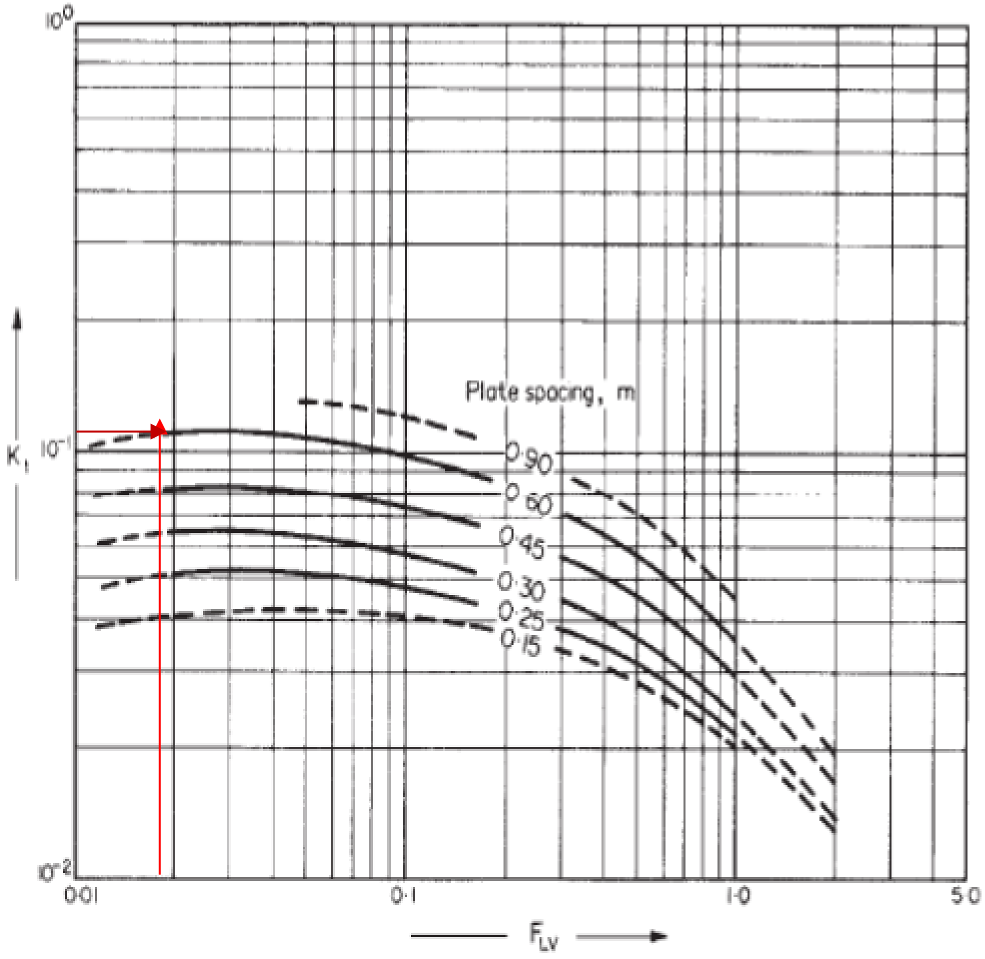


Figure 25: Plate spacing graph

K = 0.13

By putting value of k in eq (5)

$$U_f = 0.13\sqrt{\frac{716.4 - 4.46}{4.46}} = 1.64$$

Design velocity = 80% of U$_f$

U = 0.8 ×1.64= 1.313 m/s

Net column area used in separation is

$$A_n = \frac{Q_v}{U}$$

Q_v = Volumetric flow rate of vapors

$$Q_v = \frac{mass\ vapor\ flow\ rate}{3600 \times vapor\ density}$$

$$Q_v = \frac{68120}{3600 \times 4.46} = 4.2\ \frac{m^3}{s}$$

Now, net area

$$A_n = \frac{Q_v}{U} = \frac{4.2}{1.313} = 3.1\ m^2$$

Assume that down comer occupies 12% of cross-sectional Area (A_c) of column

$$A_d = 0.12\ \ A_c$$

$$A_c = \ A_n + \ A_d$$

$$A_n = \ A_c - 0.12$$

$$A_c = 0.88\ \ A_c$$

$$A_c = \frac{A_n}{0.88} = \frac{3.1}{0.88} = 3.5\ m^2$$

As we know,

$$A_c = \frac{\pi}{4} \times {D_c}^2$$

So,

$$\boldsymbol{D_c} = \left(\frac{4\ A_c}{\pi}\right)^{0.5} = \left(\frac{4 \times 3.5}{\pi}\right)^{0.5} = \mathbf{2.1}\ \boldsymbol{m}$$

And

$A_d = 0.12\ \ A_c\ \ = 0.12 \times 3.5 = 0.42\ m^2$

14) Plate Design

$$A_a = \ A_c - 2A_d$$

$$= 3.5 - (2 \times 0.42) = 2.66\ m^2$$

Hole area A_h take 10% A_a

$$A_h = 0.1 \times 2.66 = 0.2\ m^2$$

15) Weir length ($\boldsymbol{L_w}$)

$$\frac{A_d}{A_c} = \frac{0.42}{3.5} = 0.12$$

From Figure given below:

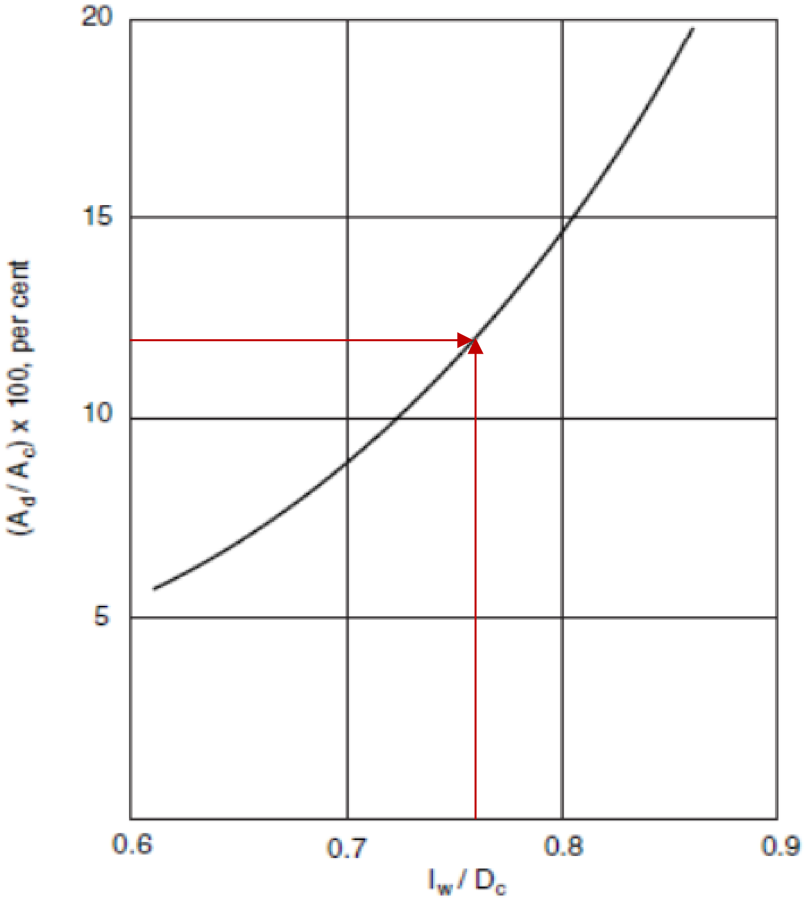


Figure 26: Weir length graph

$$\frac{L_w}{D_d} = 0.77 \dots \dots \quad (7)$$

$$D_d = \left(\frac{4\ A_d}{\pi}\right)^{0.5} = \left(\frac{4 \times 0.42}{\pi}\right)^{0.5} = 0.73\ m$$

By using eq (7)

$$L_w = 0.56\ m$$

16) Determination of entrainment correlation (ψ)

From figure given below:

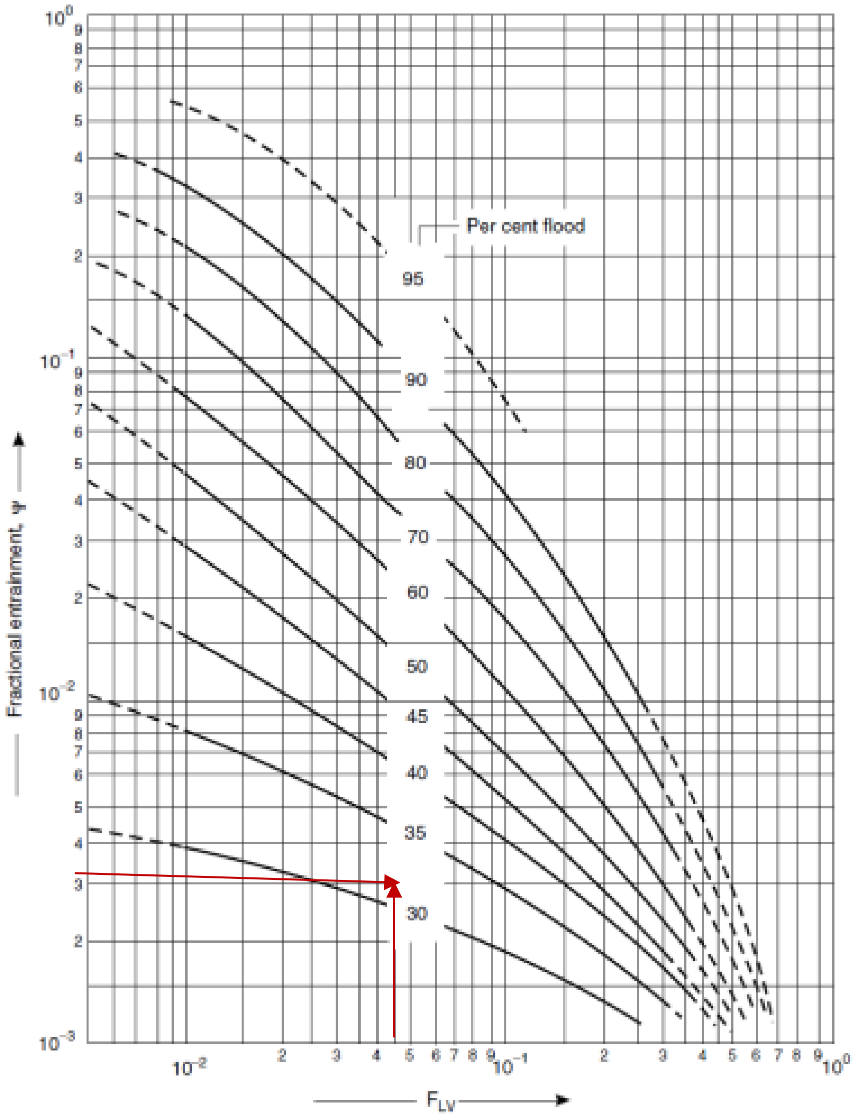


Figure 27: Entrainment graph

At $F_{LV} = 0{,}029$ and flooding 30 %:

$$\psi = 3 \times 10^{-3}$$

Now the entrainment rate is

$$\varphi = \frac{\psi \times L}{1 - \psi}$$

$$= \frac{3 \times 10^{-3} \times 10263.7}{1 - 3 \times 10^{-3}} = \mathbf{30.88\ Kg/h}$$

17) Weeping point

Weeping will occur when $U_{Omin} < U_{omincal}$

$$U_O = \frac{V}{\rho_v \times A_h} = \frac{19670}{4.46 \times 0.2 \times 3600} = 6\ \frac{m}{s}$$

Taking 70% turn down

$$U_{omin} = 0.7 \times U_o$$

$$= 0.7 \times 6 = 4.28\ \frac{m}{s}$$

Now, U_{omin} is calculated as:

$$U_{Omincal} = \frac{K_2 - 0.9(25.4 - d_0)}{\rho_v{}^{0.5}} \ldots\ldots\ldots\ldots\ldots..(8)$$

Here,

$d_0 = 5\ mm$

K_2 is a function of $(h_w + h_{owmin})$

$h_w = weir\ length = 23\ mm$

$$h_{owmin} = minimum\ weir\ crust = 750\left(\frac{L_{min}}{\rho_L \times L_w}\right)^{2/3} \ldots..(9)$$

$$L_{min} = 0.7 \times 10263.7 = 7184.5\frac{Kg}{h}$$

By putting value in eq (9)

$$h_{owmin} = 750\left(\frac{7184.5}{716.4 \times 0.56 \times 3600}\right)^{2/3} = 21.85\ mm$$

$$h_w + h_{owmin} = 21.85 + 23 = 44.85\ mm$$

From figure given below:

$$K_2 = 29.8$$

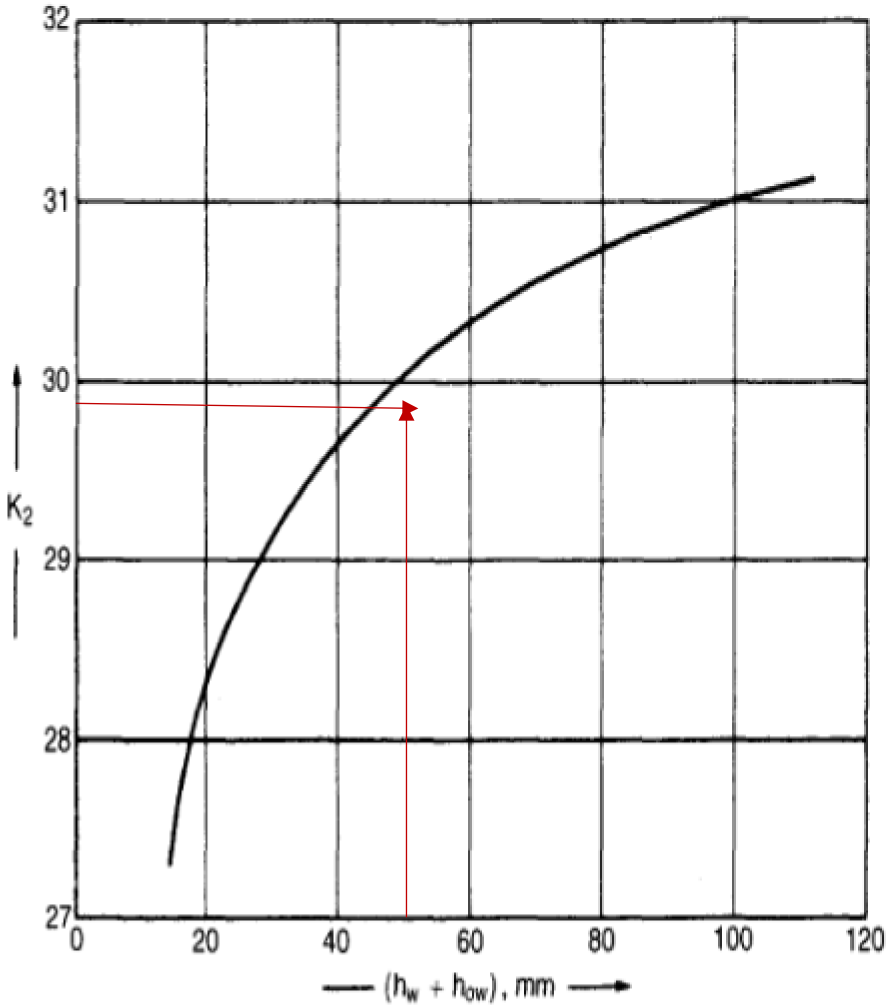


Figure 28: Weeping graph

Now by using eq (8)

$$U_{Omincal} = \frac{29.8 - 0.9(25.4 - 0.005)}{(4.46)^{0.5}} = 3.2\,\frac{m}{s}$$

As $\boldsymbol{U_{Omin} < U_{omincal}}$, so weeping will not occur.

18) Perforated area

By using the $\frac{L_w}{D_c} = 0.89$ and $\frac{L_h}{D_c} = 0.27$

the value of $\theta = 100°$ obtained from figure given below:

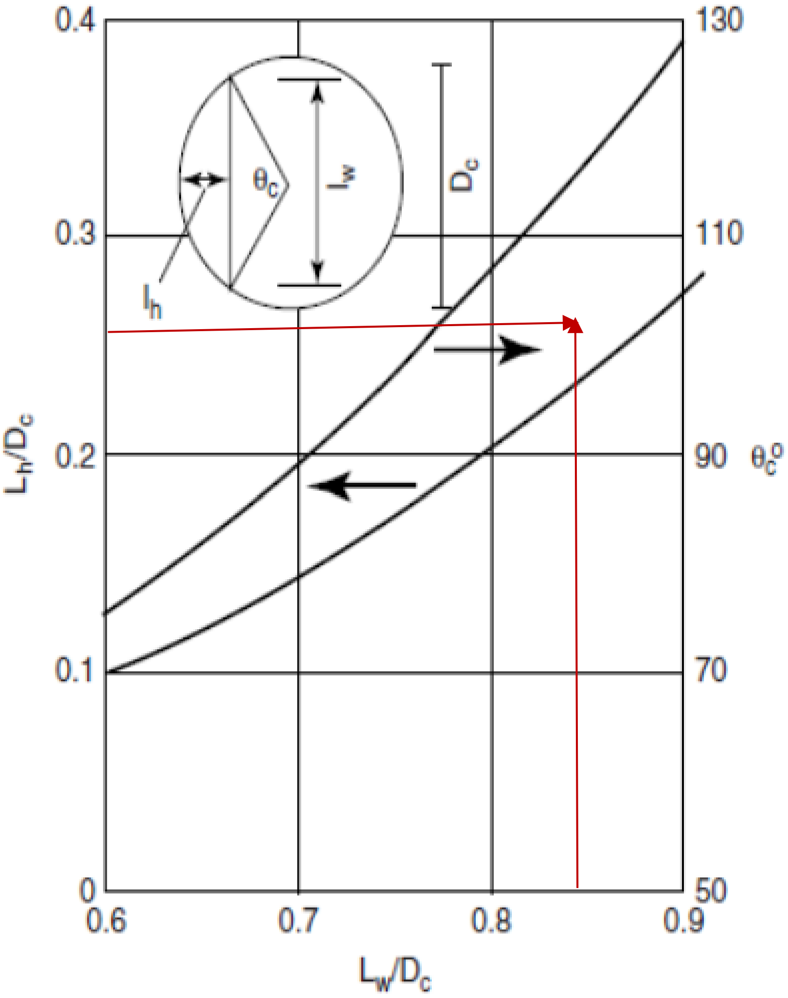


Figure 29: Perforated area graph

Angle substances at plate edge by imperforated strip = 180-100 =80˚

Hole size: 5 mm is the preferred size

Calming zone width = 50 mm

Mean Length imperforated = $(0.73 - 0.05)\pi \times \frac{80}{180} = 0.94$m

Area of imperforated = 0.05 × 0.94 = 0.047 m^2

Mean length of calming zone = $(0.73 - 0.05)\ sin\frac{100°}{2}$= 0.52 m^2

Area of calming zone = 2(0.52 × 0.05) = 0.052 m^2

Total area for perforations,

$$\boldsymbol{A_P = 3.5 - 0.047 - 0.052 = 3.4\,m}$$

19) Pressure drop calculations

$$\Delta P = 9.81 \times h_t \times 10^{-3} \times \rho_L \ldots\ldots(A)$$

ΔP = total plate pressure drop, Pa (N/m2),

$h_t = total\ plate\ drop, mm\ liquid$

$$h_t = h_d + h_w + h_{ow} + h_r \ldots\ldots(10)$$

$$h_d = 51\left(\frac{U_0}{C_o}\right)^2 \frac{\rho_v}{\rho_L} = \cdots\ldots\ldots(11)$$

Now to calculate C_o

$\frac{A_h}{A_P} = \frac{0.2}{3.4}$ = 5.86%

Plate thickness = 50 mm

Hole diameter = $d_h = 4\left(\frac{A_h}{\pi}\right)^{0.5} = 4\left(\frac{0.2}{\pi}\right)^{0.5} = 1.009\ m$

$\frac{Plate\ thickness}{Hole\ diameter} = \frac{0.05}{1.009} = 0.049\ m$By using $\frac{A_h}{A_P}$ and $\frac{Plate\ thickness}{Hole\ diameter}$, value of C_o can

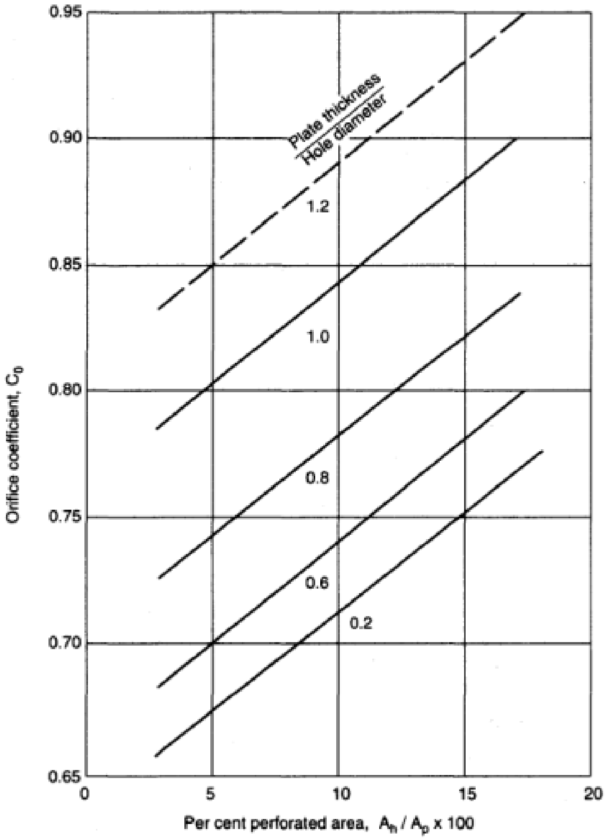


Figure 30: Orifice coefficient graph

be calculated from figure given below by using extrapolation.

$$C_o = 0.67$$

By using eq (11)

$$h_d = 51\left(\frac{6}{0.67}\right)^2 \frac{4.46}{716.4} = 25.4\,mm$$

$$h_r = \frac{12500}{716.4} = 17.44\,mm$$

$$h_w = 23mm$$

$$h_{ow} = 21.85\,mm$$

By using eq (11)

$$h_t = 24.4 + 17.44 + 23 + 21.85 = 86.69\,mm$$

By using eq (A)

$$\boldsymbol{\Delta P = 9.81 \times 86.69 \times 10^{-3} \times 716.4 = 0.609\ kpa}$$

20) **Down comer liquid back up**

For safe design and to avoid flooding

$$h_b < \frac{1}{2}(C + h_w) \ldots\ldots.(12)$$

Here

C = tray spacing

$$h_b = h_t + h_w + h_{ow} + h_{dc} \ldots\ldots(13)$$

$h_b =$ down comer back – up, measured from plate surface, mm

$h_{dc} =$ head loss in the down comer, mm

$$h_{dc} = 166\left(\frac{L}{\rho_L A_m}\right) \ldots\ldots(14)$$

A_m= either the down comer area Ad or the clearance area under the downcomer Aap; whichever is the smaller,[2]

L = liquid flow rate in down comer, kg/s

$$A_m = A_{ap} = h_{ap} \times L_w \ldots..(15)$$

$$h_{ap} = h_w - 5\,mm = 23 - 5 = 18\,mm$$

By using eq (15)

$$A_{ap} = 0.56 \times 18 \times 10^{-3} = 0.01\,m2$$

By using eq (14)

$$h_{dc} = 166\left(\frac{10263.7}{716.4 \times 0.01 \times 3600}\right) = 66\,mm$$

By using eq (13)

$$h_b = 86.69 + 23 + 21.85 + 66 = 197.75\,mm = 0.197\,m$$

By using eq (12)

$$= \frac{1}{2}(0.73 + 23 \times 10^{-3}) = 0.37$$

As $h_b < \frac{1}{2}(C + h_w)$, so flooding will not occur.

21) Down comer residence time:

$$t_r = \frac{A_d h_b \rho_L}{L} = \frac{0.42 \times 0.197 \times 716.4}{10263.\frac{7}{3600}} = \mathbf{20.79\,s}$$

Where:

t_r= residence time in seconds. [10]

6.3 PREHEATER

The major parts of a furnace include a Radiant section, Convection section, stack, tubes, piping and instrumentation, burners, and insulations. The furnace design relates the required heater duty to the required number of the burners, coils, dimensions, insulation, etc.

The total heat transfer to the cold surface is given by the following relation:

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

Where,

$\boldsymbol{f}$ is the overall exchange factor.

$\boldsymbol{T_G}$ is the temperatures of the flue gases in ^{o}R

$\boldsymbol{T_S}$ is the surface temperature of the tubes in ^{o}R

$\boldsymbol{A_{cp}}$ is the surface area of cold surface, ft^2

$\boldsymbol{\alpha_{cp}}$ is a factor by which the cold surface area must be reduced to get the effective area

In the case of natural convection usually, we assume, A ≈ $2\alpha_{cp}A_{cp}$, h ≈ 2, and f ≈ 0.57. Therefore, the above equation is reduced to the following form:

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

To design the furnace, the following parameters should be known:

Total required heater duty (Btu/hr)	**8.44 x 10^7**
Efficiency, η	95 %
Fuel value [Btu/hr].	2 x 10^4
The temperature of inlet air [^{o}F].	77
Amount of atomizing steam and ratio [lb steam/lb fuel].	0 because we are not using steam in our design.
Tubes diameters, d_o (in)	10.56 (0.88 ft)
tubes center-to-center distance, ctc, [in.]	10.92 (0.91 ft)
Exposed tube length, L [ft].	55
The total length of both radiative and convective section, [ft]	98

Average flux for both radiant and convective sections, q [Btu/hr.ft2].	3.4 x 10^3

Table 20: Preheater specifications

The First step is to calculate $(A/F)_{st}$ by using the following fuel composition:

Component	**No. of moles**
$\boldsymbol{CH_4}$	0.3
$\boldsymbol{C_2H_6}$	0.3
$\boldsymbol{C_3H_8}$	0.1
$\boldsymbol{C_4H_{10}}$	0.2
$\boldsymbol{C_5H_{12}}$	0.1
	Total = 1 mole of fuel

Balanced chemical equations for combustion of all the above components are as follows:

$$0.3CH_4 + 0.6(O_2 + 3.76N_2) \rightarrow 0.3CO_2 + 0.6H_2O$$

$$0.3C_2H_6 + 1.05(O_2 + 3.76N_2) \rightarrow 0.6CO_2 + 0.9H_2O$$

$$0.1C_3H_8 + 0.5(O_2 + 3.76N_2) \rightarrow 0.3CO_2 + 0.4H_2O$$

$$0.2C_4H_{10} + 1.3(O_2 + 3.76N_2) \rightarrow 0.8CO_2 + 1H_2O$$

$$0.1C_5H_{12} + 0.8(O_2 + 3.76N_2) \rightarrow 0.5CO_2 + 0.6H_2O$$

$n_{air,st} = (0.6 + 1.05 + 0.5 + 1.3 + 0.8)(1 + 3.76) = 20.23$

So,

$$\left[\frac{A}{F}\right]_{st} = \frac{\mathbf{20.23}}{\mathbf{1}} = \mathbf{20.23}$$

To get the estimated T_G, let

$$\frac{\boldsymbol{\Sigma Q}}{\boldsymbol{\alpha_{cp} A_{cp} f}} = \mathbf{2} \times \boldsymbol{q}(\boldsymbol{Avg\ Flux}) \quad = 2 \times 3.4 \times 10^3\ \frac{Btu}{hr.ft^2} = 6.8 \times 10^3\,\frac{Btu}{hr.ft^2}$$

Now assume T_S = 600°F and check the following graph for T_S and $\frac{\Sigma Q}{\alpha_{cp} A_{cp} f}$ to find T_G,

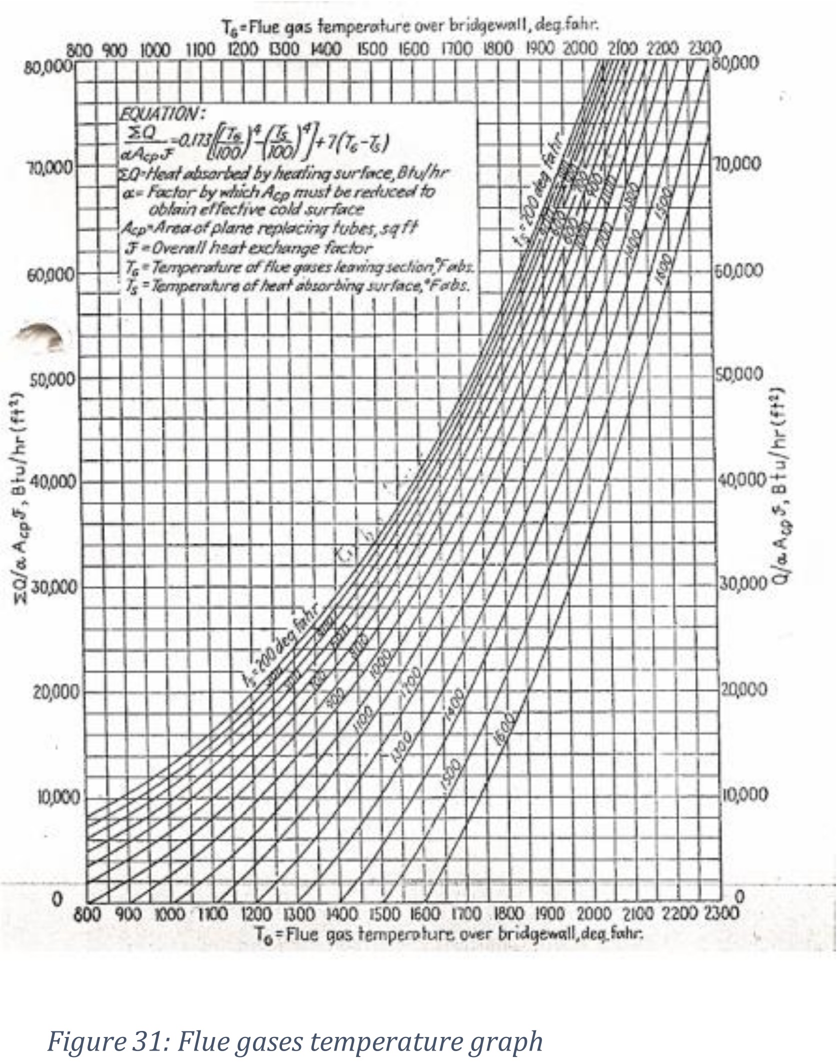


Figure 31: Flue gases temperature graph

From here estimated **T_G= 945 °F or 1405 °R.**

As now we have found out, this value, we will carry out further calculations, including the whole design of the furnace to check whether the T_G calculated this way is accurate. If it becomes accurate then all the values mentioned above will be taken as correct.

Now a detailed design will be done for finding out whether our above assumptions are true or not.

The heat released by the burners in total would be calculated by dividing heater duty by the efficiency:

$$Q_F = \frac{Heater\ Duty}{\eta} = \frac{8.44 \times 10^7}{0.95} = \mathbf{8.8 \times 10^7}\ \frac{\boldsymbol{Btu}}{\boldsymbol{hr}}$$

Now from the already known values we will find out the mass of the fuel required:

$$m_{fuel} = \frac{Q_F}{Fuel\ Value} = \frac{8.8 \times 10^7}{2 \times 10^4} = \mathbf{4400}\ \frac{\boldsymbol{lb}}{\boldsymbol{h}}$$

For the above found out mass we will now the corresponding value of the mass of air required for the reaction.

$$m_{air,st} = (\frac{A}{F})_{st} \times m_{fuel}$$

$$= (20.23) \times (4400) = 8.8 \times 10^4 \frac{lb}{h}$$

As we are not using percent excess air here so

$$m_{air} = \mathbf{8.8 \times 10^4} \frac{\boldsymbol{lb}}{\boldsymbol{h}}$$

The heat liberated by the inlet air will be now calculated. This will be calculated by the following equation:

$$Q_{air} = m_{air} \times C_{p,air} \times (T_{air} - T_{ref})$$

Here $T_{ref}\ is\ taken\ as$ 60°F and $C_{p,air} is\ taken\ as$ 0.24 $\frac{Btu}{lb.°F}$.

So,

$$Q_{air} = 8.8 \times 10^4 \times 0.24 \times (77 - 60)$$
$$= \mathbf{3.66 \times 10^5}\ \frac{\boldsymbol{Btu}}{\boldsymbol{hr}}$$

Now the heat which is taken up by the walls of the furnace will be calculated by the following equation:

$$Q_{wall} = 2\%\ Q_F$$
$$= 0.02 \times (8.8 \times 10^7) = \mathbf{1.76 \times 10^6} \frac{\boldsymbol{Btu}}{\boldsymbol{hr}}$$

The heat released through the exhaust gases will now be calculated by the following equation:

$$Q_{exhaust} = m_{fuel} \times (1 + G') \times C_{p,avg} \times (T_G - 520) \ldots. (1)$$

The T_G used here will used in Rankine (^OR) and G' will be the air to fuel ratio. To find $C_{p,avg}$:

$$C_{p,avg} = \Sigma x_i C_{p,i}$$

Here x usually refers to moles of $CO_2\ and\ H_2O$. From the previous calculations the moles will be calculated as:

$$x = 0.3 + 0.6 + 0.6 + 0.9 + 0.3 + 0.4 + 0.8 + 1 + 0.5 + 0.6 = 6$$

$$x_{CO_2} = \frac{0.3 + 0.6 + 0.3 + 0.8 + 0.5}{6} = 0.41$$
$$x_{H_2O} = \frac{0.6 + 0.9 + 0.4 + 1 + 0.6}{6} = 0.58$$

At 945 °F,

$$C_p of\ CO_2 = 0.081\ \frac{Btu}{lb.°R}$$

$$C_p of\ H_2O = 1.03 \times 10^{-3}\ \frac{Btu}{lb.°R}$$

Hence,

$$C_{p,avg} = (0.41 \times 0.081) + (0.58 \times 1.03 \times 10^{-3})$$

$$= \ 0.03 \frac{Btu}{lb.°R}$$

By substituting values in eq (1)

$$Q_{exhaust} = 4400\ \frac{lb}{h} \times (1 + 20.23) \times 0.03 \frac{Btu}{lb.°R} \times (1405 - 520)°R$$

$$= \mathbf{2.4 \times 10^6}\ \boldsymbol{\frac{Btu}{hr}}$$

Now we will calculate the number of tubes required inside the furnace to carry the desired heating. The formula for it would be:

$$N_{tubes} = \frac{Q}{2\pi r L_T q}$$

$$= \frac{8.4 \times 10^7}{2\pi\left(\frac{0.88}{2}\right) \times 98 \times 3.4 \text{ x } 10^3} \approx \mathbf{91}$$

Now the cold surface area available for heat exchange will be calculated. The following formula would be used:

$$A_{cp} = ctc \times L \times N_{tubes}\ = 0.91 \times 55 \times 91 = \mathbf{4413}\ \boldsymbol{ft^2}$$

The ctc/diameter ratio will be calculated to find the value of α from the graph down below:

$$\frac{ctc}{d_o} = \frac{0.91}{0.88} = 1.03$$

From the followng graph, point out the intersection between the ctc/do and the direct one row curve to find value of α.

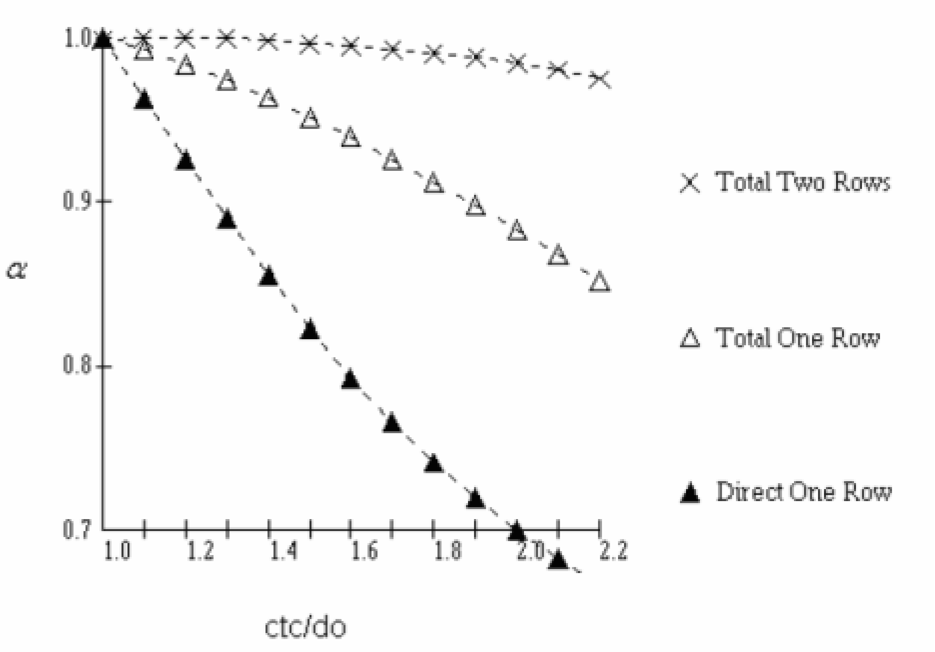


Figure 32: Effective area graph

The value of α from the graph above is found out to be **0.98**.

Now the total exposed area of the furnace will be calculated which taken in heat from the burners. The following ratio is considered:

Length : Width : Height

3 : 2 : 1

In our design, ratios will be as follows

Length : Width : Height

55 : 36.6 : 18.33

The area will be calculated by the following formula:

$$A_T = (2 \text{ x Area of left side OR right side}) + (\text{Area bottom})$$

$$+ (2 \text{ x Area of back OR front side})$$

$$= (2 \times 18.33 \times 36.6) + (36.6 \times 55) + (2 \times 18.33 \times 55)$$

$$= \mathbf{5371\ ft^2}$$

The total refracting surface will be given by the following formula that gives the effective refracting surface:

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

$$= 5371 - (0.98 \times 4413) = \mathbf{1046.2\ ft^2}$$

The gas emissivity ε of the flue gases will be calculated from the graph down below. This graph is against the emissivity and pL, which is the product of partial pressure of carbon dioxide and water and the length of the furnace.

At 940 °F

$$\boldsymbol{p_{CO_2} = 0.089\ atm}$$

$$\boldsymbol{p_{H_2O} = 0.12\ atm}$$

$$pL = [p_{CO_2} + p_{H_2O}] \times L$$

$$= (0.089 + 0.12) \times 18.33 = \mathbf{3.83\ atm.ft}$$

The units will be atm-ft. Knowing the pL product and the temperature of the hot flue gases we can find the emissivity through their intersection from the graph below:

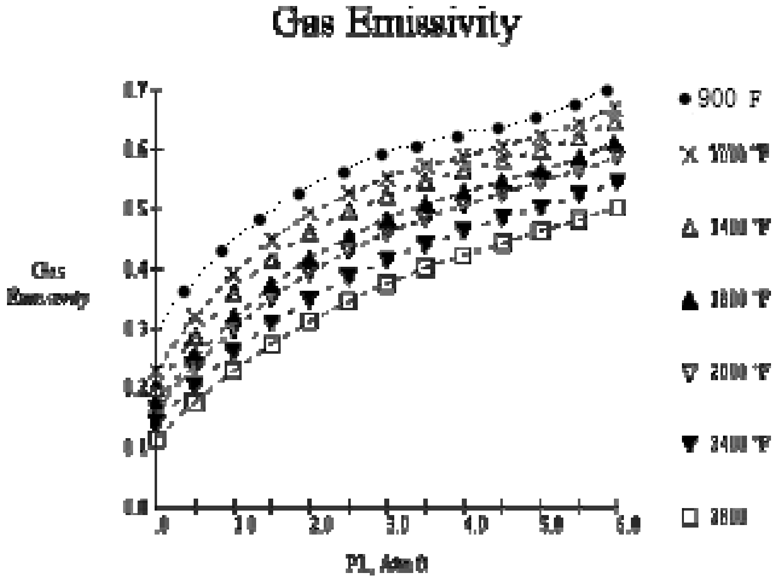


Figure 33: Emissivity graph

So value of **ε** calculated from above graph is **0.62.**

Now by incorporating this emissivity value and the factor of $\frac{A_R}{\alpha A_{cp}}$ we will obtain a factor "f" from the graph down below. It will be the overall heat exchange factor.

$$\frac{A_R}{\alpha A_{cp}} = \frac{1046.2}{0.62 \times 4413} = 0.38$$

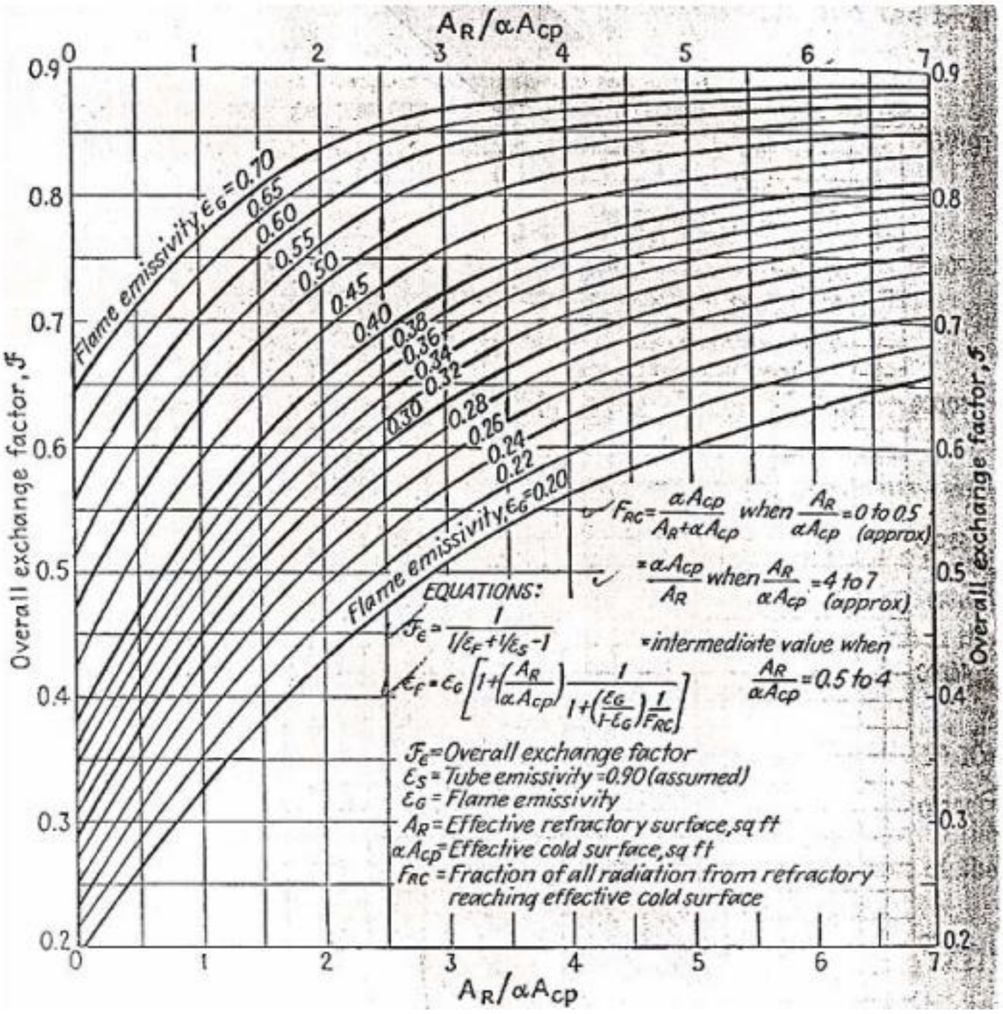


Figure 34: Overall exchange factor graph

The value of **f** obtained from the graph will be **0.67**.

The value of the following factor will now be calculated:

$$\frac{\boldsymbol{\Sigma Q}}{\boldsymbol{\alpha_{cp}A_{cp}f}} = \frac{8.4 \times 10^{7}}{0.98 \times 4413 \times 0.67} = \mathbf{1 \times 10^{4}}$$

Now we go back to the first graph to find the value of T_G. if this value comes close to the values that we assumed beforehand then our furnace has a feasible design otherwise the calculations will have to be repeated. So, from the graph below the value of T_G is 1000 °F which is close to our estimated value of 940°F.

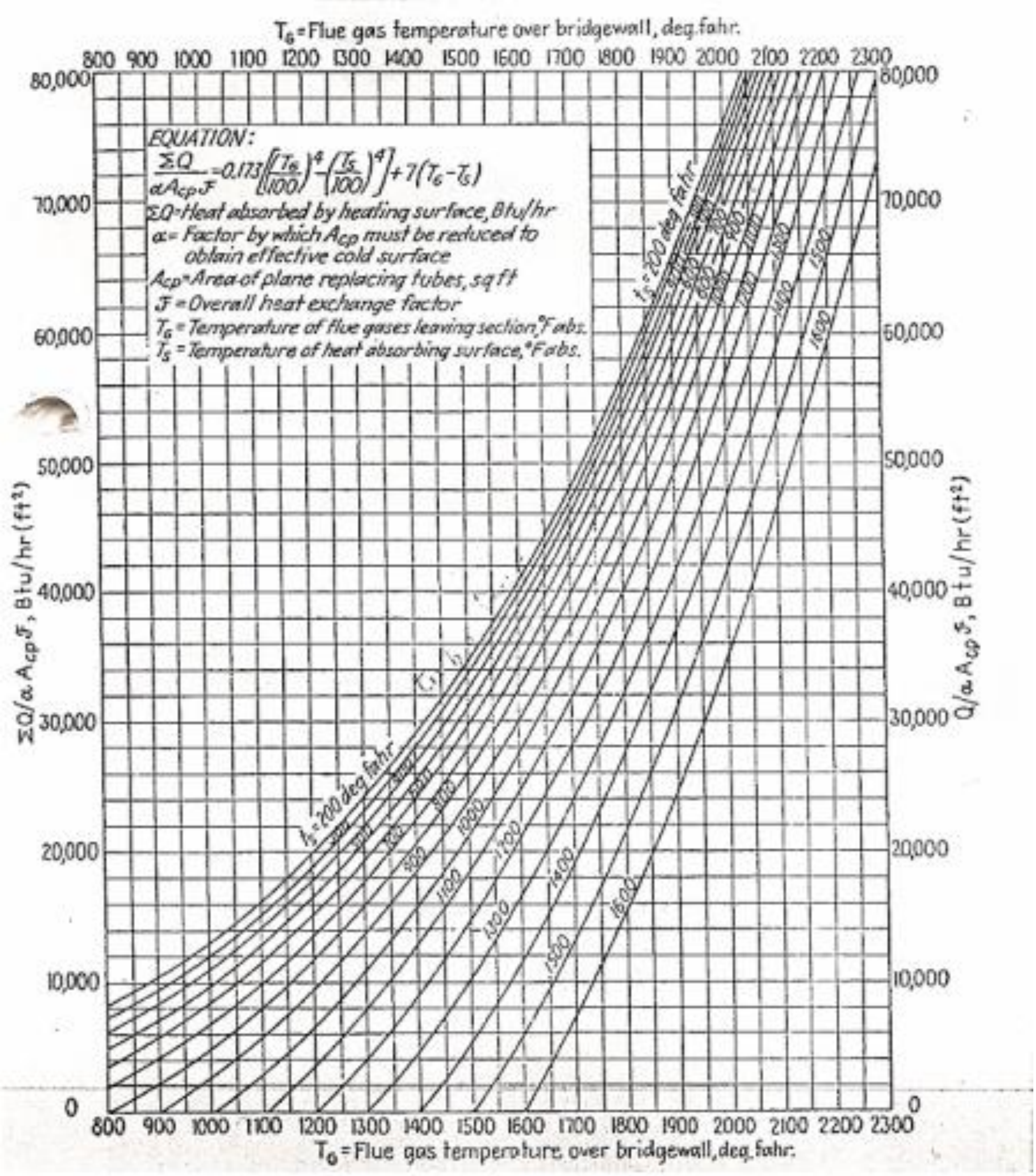


Figure 35: Flue gases temperature pointing values from graph

As these values are quite close hence we can say that our furnace has a feasible design. [11-13]

1) Burner Design

For the burner design we will take the whole burner duty and divide it by the number of burners to get the duty required of a single burner. With this we will find the mass of the fuel being injected and its release speed which will lead us to the diameter of the burner nozzle.

Before moving towards the design, we need to know the following parameters:

Duty of furnace(J/s)	**2.4 x10^7 (8.44 x 10^7 Btu/hr)**
No of Burners	21
Fuel Heating Value(J/Kg)	4.67 x 10^7
Fuel Injection Speed (m/s)	5

Table 21: Burner specifications

The following steps will be followed:

The duty of a single burner would be calculated as follows:

$$\boldsymbol{Duty\ of\ one\ Burner = \frac{Furnace\ Duty}{No.of\ Burners}}$$

$$\boldsymbol{= \frac{2.4 \times 10^7}{21} = 1.1 \times 10^7 \frac{J}{s}}$$

The fuel heating value has already been identified above which will give us the mass flow rate of the fuel through the following equation:

$$\boldsymbol{mass\ flow\ rate = \frac{Furnace\ Duty}{Fuel\ Heating\ Value}}$$

$$\boldsymbol{= \frac{2.4 \times 10^7}{4.67 \times 10^7} = 0.5\ \frac{Kg}{s}}$$

Now the volumetric flow rate will be obtained from the mass flow rate and the density of the fuel as follows:

$$\rho_{fuel} = \left(0.3 \times \rho_{CH_4}\right) + \left(0.3 \times \rho_{C_2H_6}\right) + \left(0.1 \times \rho_{C_3H_8}\right) + \left(0.2 \times \rho_{C_4H_{10}}\right) + (0.1 \times \rho_{C_5H_{12}})$$

Densities of above compunds, obtained from literature, are as follows:

$$\rho_{CH_4} = 0.668\ \frac{Kg}{m^3}$$

$$\rho_{C_2H_6} = 1.26 \frac{Kg}{m^3}$$

$$\rho_{C_3H_8} = 2.01 \frac{Kg}{m^3}$$

$$\rho_{C_4H_{10}} = 2.48 \ \frac{Kg}{m^3}$$

$$\rho_{C_5H_{12}} = 626 \ \frac{Kg}{m^3}$$

So,

$$\rho_{fuel} = (0.3 \times 0.668) + (0.3 \times 1.26) + (0.1 \times 2.01) + (0.2 \times 2.48) + (0.1 \times 626) = 63.87 \ \frac{Kg}{m^3}$$

Now

$$Volume = \frac{mass\ flow\ rate}{\rho_{fuel}}$$

$$= \frac{0.5}{63.87} = 0.0078 \frac{m^3}{s}$$

Now we will assume the speed of the flame release and fuel injection to be 5 m/s. Now the area will be calculated as follows:

$$Area = \frac{Volumetric\ flow\ rate}{Speed\ of\ fuel\ release}$$

$$= \frac{0.0078}{5} = 0.0015\, m^2$$

The diameter will be calculated as follows:

$$Diameter = 2 \times \sqrt{\frac{Area}{\pi}}$$

$$= 2 \times \sqrt{\frac{0.0015}{\pi}} = 0.04\, m = 4\, cm$$

The diameter of our multi nozzle burners will be 4 cm. There will be three nozzles on one burner and each will have a diameter of 4 cm. These nozzles will be inclined towards each other to have a radial and uniform flame envelope.[13]

CHAPTER 7

SIMULATION

7.1 SIMULATION MAIN FLOW SHEET

The proposed flow sheet is given as follows

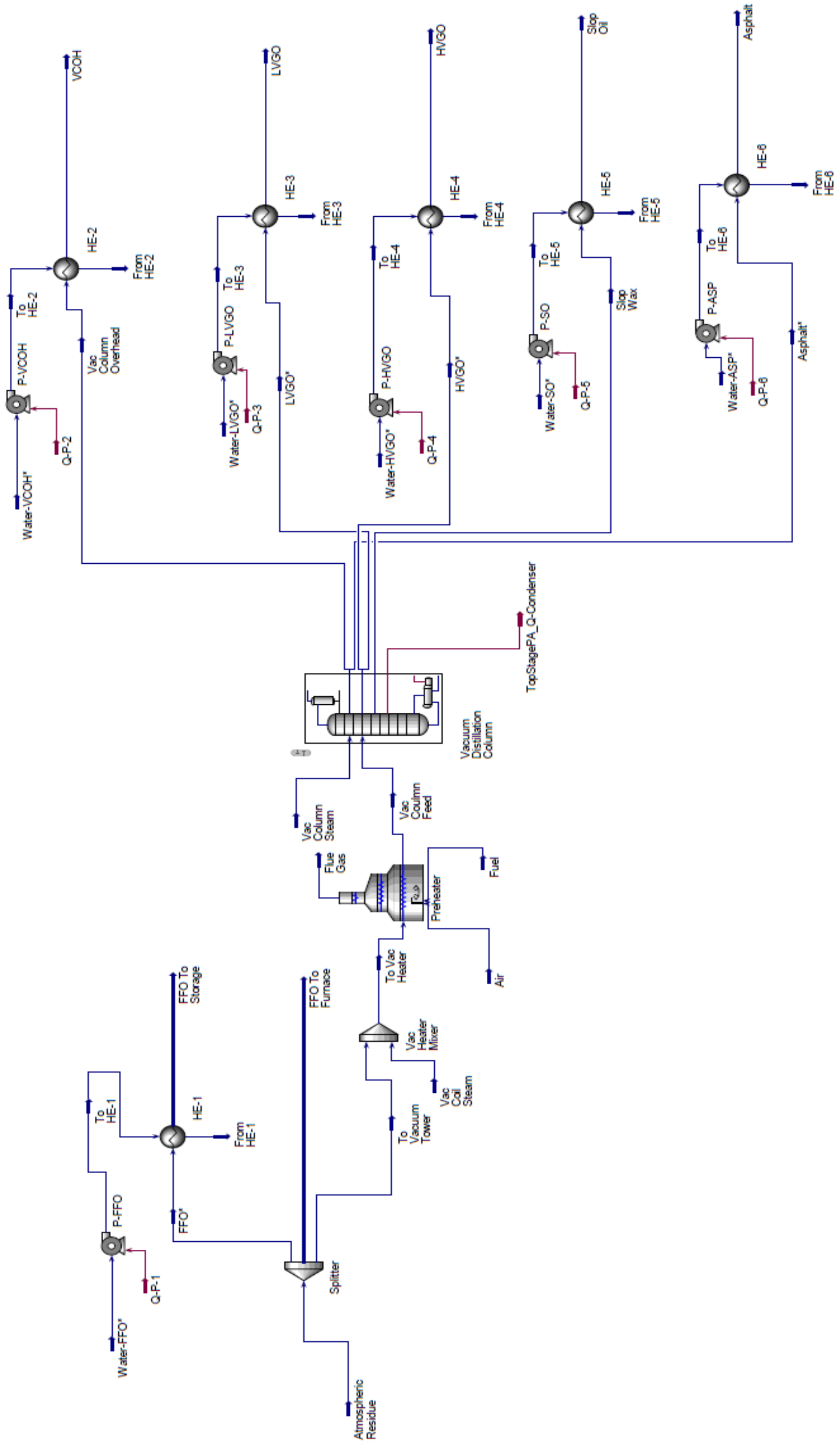


Figure 36: Main simulation sheet

7.2 LIGHTER AND HYPOTHETICAL COMPONENTS INPUT

For introducing the lighter and hypothetical components in the simulation for our proposed flow sheet, perform several steps on Aspen HYSYS:

- First of all, click and open **Aspen HYSYS V11**.
- Click on **Properties tab** at left bottom corner.
- Click on component list and add all the lighter components one by one.
- Add data given in table below.

Light Components	
Components	**Boiling Point (^{O}C)**
Methane	**-161.525**
Ethane	**-88.59999695**
Propane	**-42.10199585**
i-Butane	**-11.72998657**
n-Butane	**-0.501989746**
i-Pentane	**27.87801514**
n-Pentane	**36.05901489**
H_2O	**99.99801025**
O_2	**-182.9503997**
N_2	**-195.8001998**
CO	**-191.4503997**
CO_2	**-78.55199279**

Table 22: Light components

- Now click on **Petroleum Management tab** and click lighter Fractions.
- Add data given in table below.

Hypothetical components	Boiling Point (^{O}F)	Molecular weight(g/mol)	Density (kg/m3)
NBP[0]114	113.9	76.64	**679.52**
NBP[0]139	138.9	78.50	**683.41**
NBP[0]162	162.0	84.29	**695.02**
NBP[0]188	188.3	91.19	**708.26**
NBP[0]215	214.6	98.45	**720.86**
NBP[0]241	240.6	106.66	**733.42**
NBP[0]266	265.7	114.77	**746.09**
NBP[0]292	292.1	122.51	**758.68**

NBP[0]319	319.1	129.91	**768.47**
NBP[0]345	344.6	137.19	**776.14**
NBP[0]371	370.5	147.21	**785.33**
NBP[0]397	396.5	157.59	**794.08**
NBP[0]423	422.6	168.26	**802.29**
NBP[0]449	448.7	179.42	**810.39**
NBP[0]475	474.7	192.13	**819.14**
NBP[0]501	500.7	203.92	**826.91**
NBP[0]527	526.8	216.22	**834.49**
NBP[0]553	552.8	230.57	**842.69**
NBP[0]579	578.9	244.85	**850.19**
NBP[0]605	605.0	259.63	**857.31**
NBP[0]631	631.0	275.70	**864.76**
NBP[0]657	657.1	290.27	**871.26**
NBP[0]683	683.1	306.64	**878.39**
NBP[0]709	709.2	322.86	**885.57**
NBP[0]735	735.2	337.77	**892.22**
NBP[0]761	761.3	353.71	**898.95**
NBP[0]787	787.3	373.42	**906.24**
NBP[0]825	824.7	395.41	**913.76**
NBP[0]875	874.8	427.78	**925.03**
NBP[0]925	924.6	465.12	**938.85**
NBP[0]975	974.7	497.77	**951.83**
NBP[0]1024	1024.0	530.35	**965.29**
NBP[0]1074	1074.0	565.79	**980.13**
NBP[0]1124	1124.0	601.08	**994.09**
NBP[0]1174	1174.0	639.40	**1008.17**
NBP[0]1243	1243.0	683.59	**1023.18**
NBP[0]1343	1343.0	747.80	**1046.00**
NBP[0]1451	1451.0	812.21	**1068.81**
NBP[0]1511	1511.0	842.98	**1102.03**
NBP[0]1605	1605.0	855.75	**1130.02**
NBP[0]1734	1734.0	876.08	**1160.05**
NBP[0]1851	**1851.0**	**900.35**	**1186.49**

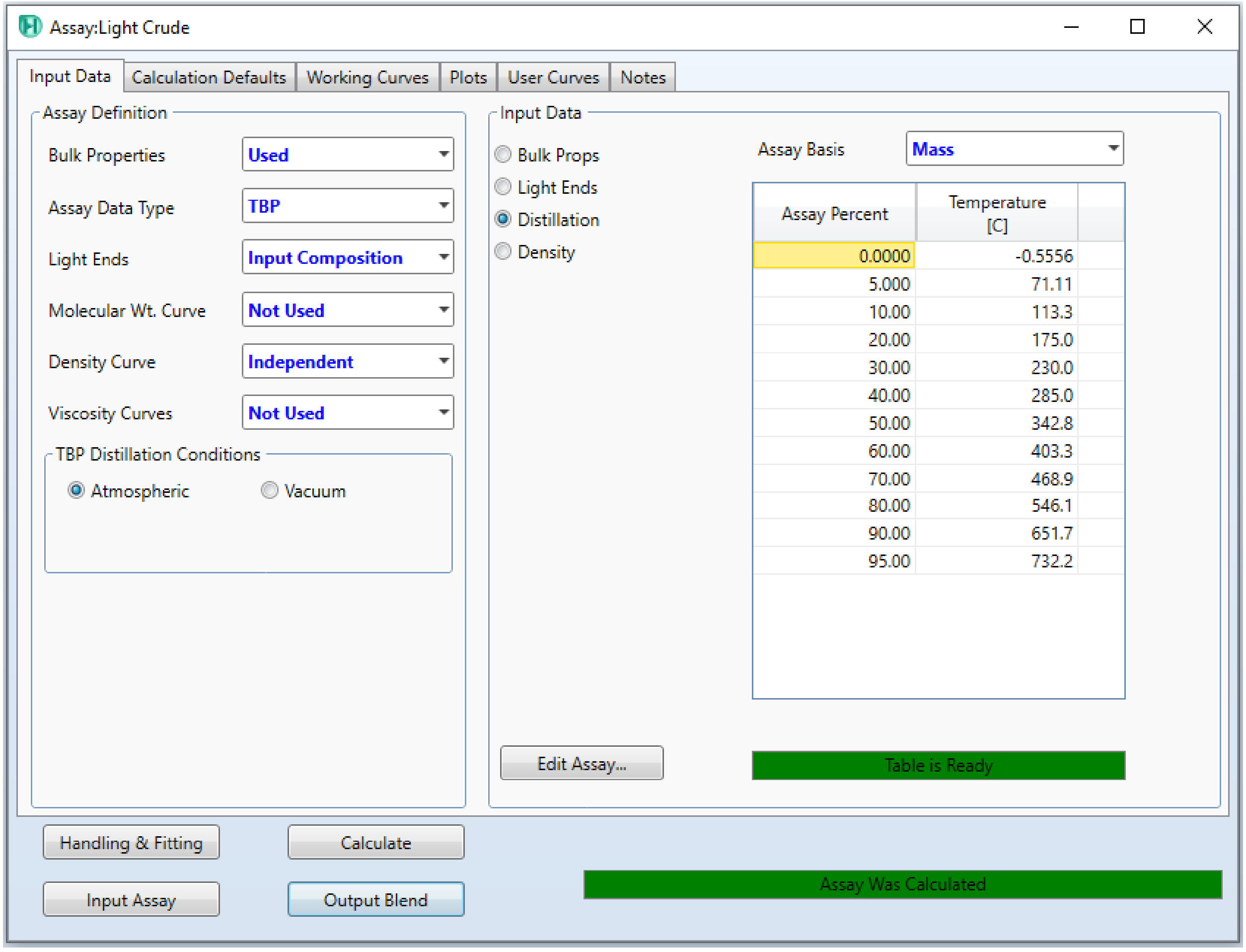
Table 23: Hypothetical components

7.3 ASSAY GENERATION AND OIL GENERATION

So after addition of all these types of data, following steps should be done.

- Click on **Oil Manager**.
- Click on Input Assay button and then go to Add button. Use the **Bulk properties** and specify Assay data type of TBP (True Boiling Point).
- Ensure that Assay Basis is Mass.
- Input the TBP curve on wt% basis for **Light crude**. Click on Edit assay button.
- Now add light ends composition. Light ends option must be active.
- Then add the density data. Make sure that **Density option** must be active.
- Add the data needed for oil characterization and allow HYSYS to create pseudo-components. Make the **Bulk props** option active.
- For oil characterization, **sulfur distribution** is required. Define the **User property** in appropriate button under Home Tab.
- After creating sulfur distribution, add the data and highlight sulfur in Available properties column and enter Add.
- In input assay, go to Light Crude, repeat the same steps for Medium and Heavy crude.
- After entering all assays, go to each crude oil and select **Calculate** button if there’s a problem that assay hasn’t been calculated.
- Develop a blend of three crudes (light, medium, and heavy), go to **Output Blend** and select Add button.

After installing crude oil you can analysis the product distribution and various other fluid properties.



Assay Definition

Bulk Properties: Used

Assay Data Type: TBP

Light Ends: Input Composition

Molecular Wt. Curve: Not Used

Density Curve: Independent

Viscosity Curves: Not Used

TBP Distillation Conditions

● Atmospheric ○ Vacuum

Input Data

○ Bulk Props
○ Light Ends
● Distillation
○ Density

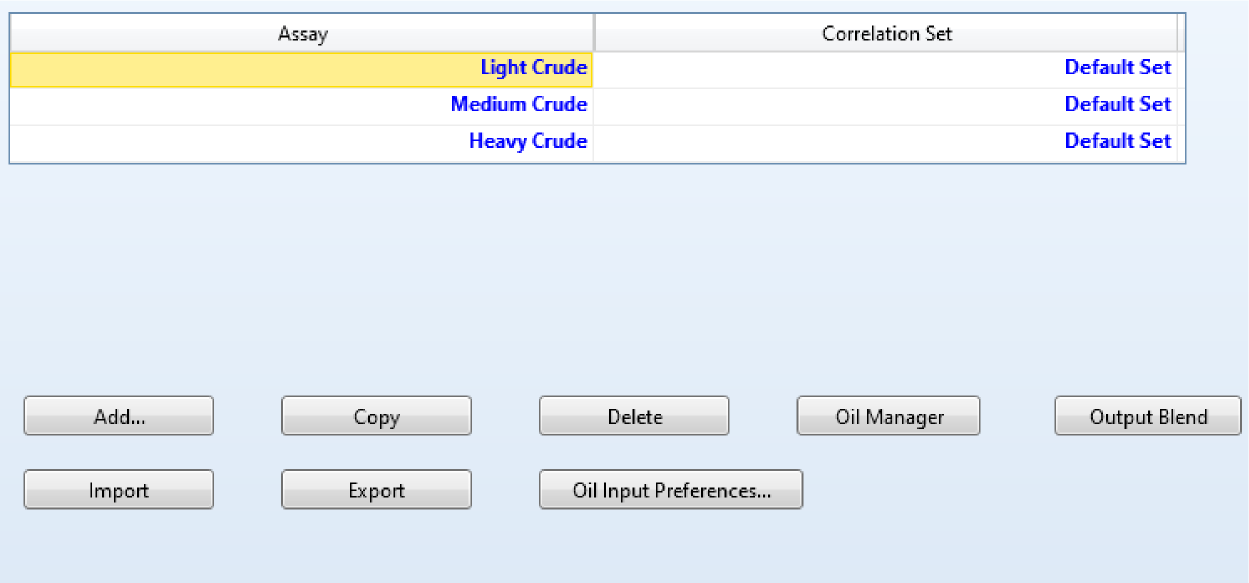
Assay Basis: Mass

Assay Percent	Temperature [C]
0.0000	-0.5556
5.000	71.11
10.00	113.3
20.00	175.0
30.00	230.0
40.00	285.0
50.00	342.8
60.00	403.3
70.00	468.9
80.00	546.1
90.00	651.7
95.00	732.2

Table is Ready

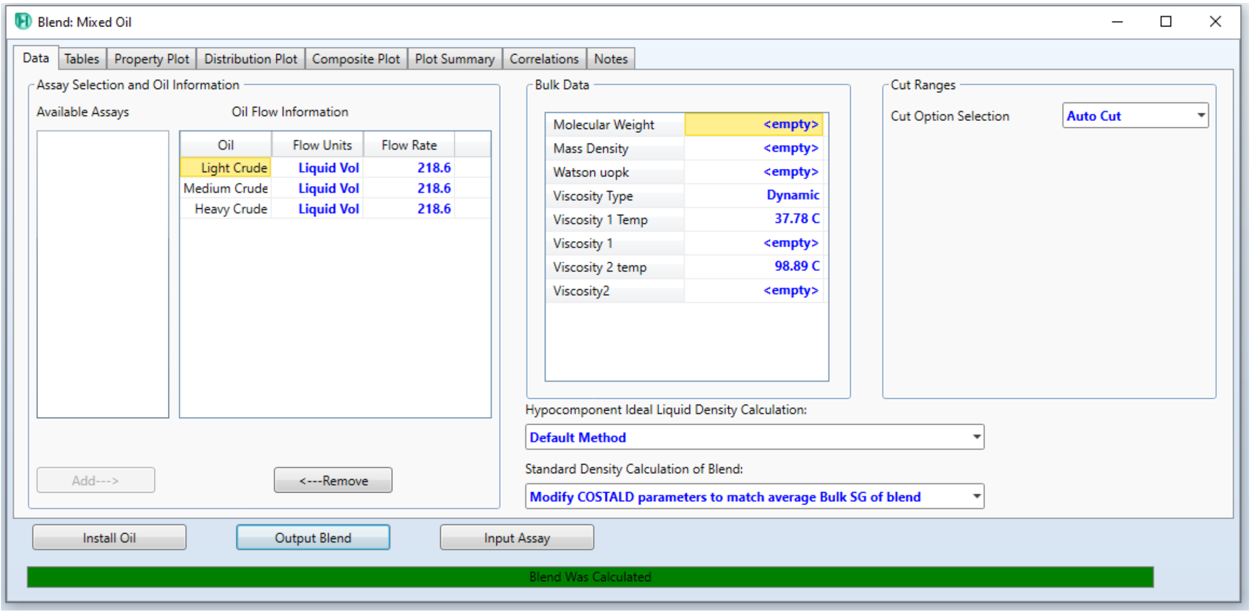
Assay Was Calculated

Figure 37: Assay Calculation



Assay	Correlation Set
Light Crude	Default Set
Medium Crude	Default Set
Heavy Crude	Default Set

Figure 38: Types of Crude



Oil	Flow Units	Flow Rate
Light Crude	Liquid Vol	218.6
Medium Crude	Liquid Vol	218.6
Heavy Crude	Liquid Vol	218.6

Molecular Weight	<empty>
Mass Density	<empty>
Watson uopk	<empty>
Viscosity Type	Dynamic
Viscosity 1 Temp	37.78 C
Viscosity 1	<empty>
Viscosity 2 temp	98.89 C
Viscosity2	<empty>

Cut Option Selection: Auto Cut

Hypocomponent Ideal Liquid Density Calculation: Default Method

Standard Density Calculation of Blend: Modify COSTALD parameters to match average Bulk SG of blend

Figure 39: Oil Installation

Products Analysis	
Products	**Temperature (^{o}C)**
Off Gas	10
Lt St Run	70
Naphtha	180
Kerosene	240
Light Diesel	290
Heavy Diesel	340
Atm Gas Oil	370
Residue	1200

Table 24: Product distribution

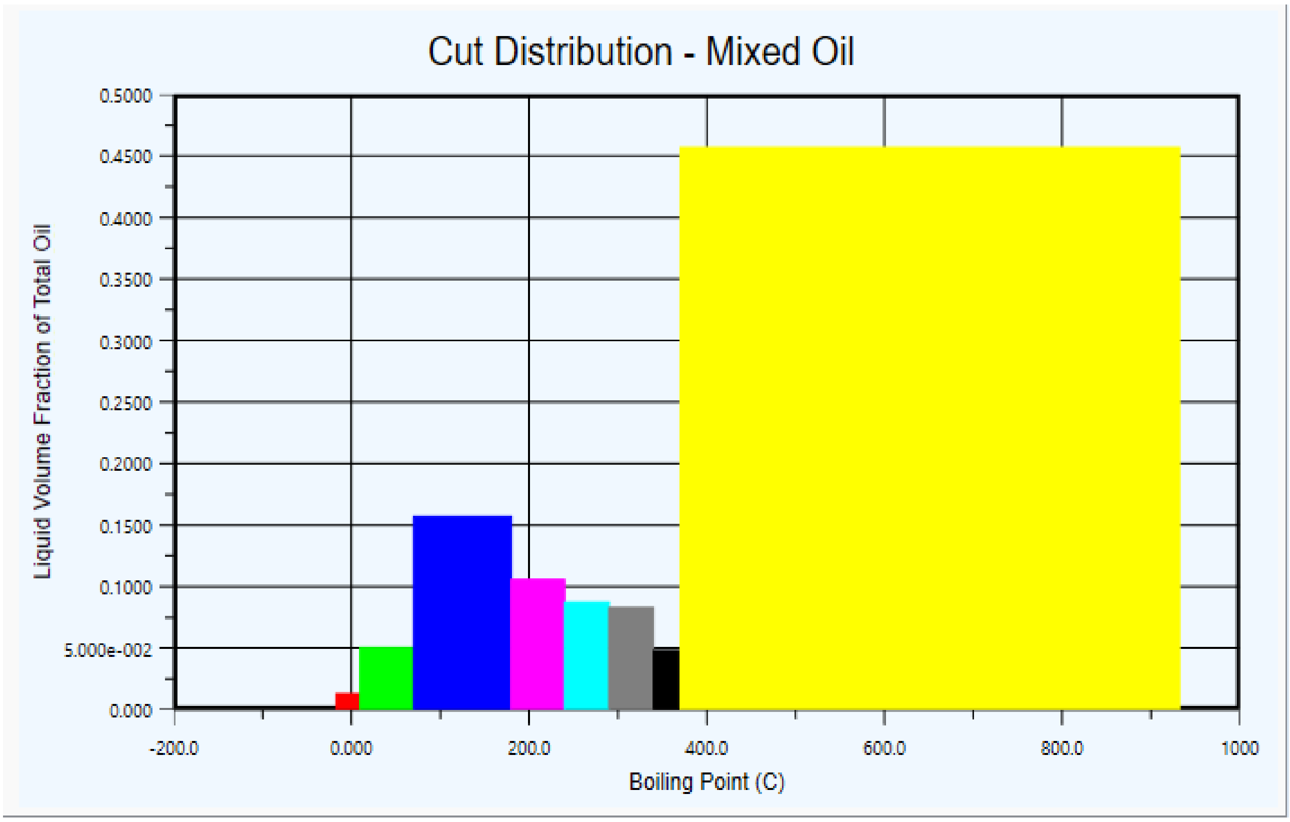


Figure 40: Cut distribution

True boiling point analysis guides us for estimating cut points for all the products and their flow rates.

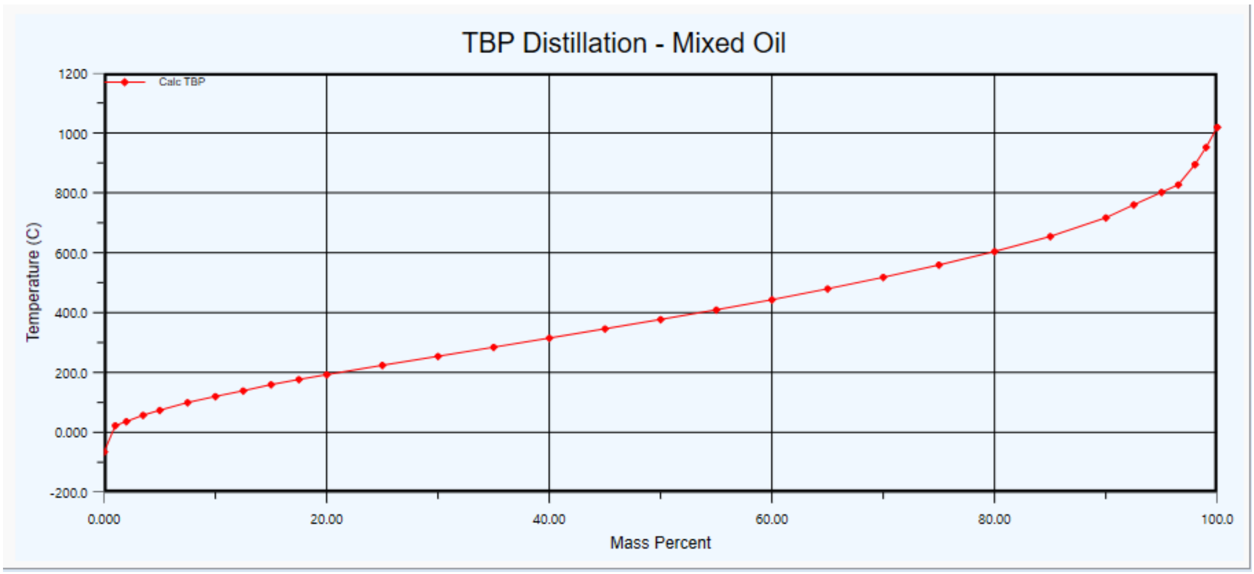


Figure 41: True boiling point curve

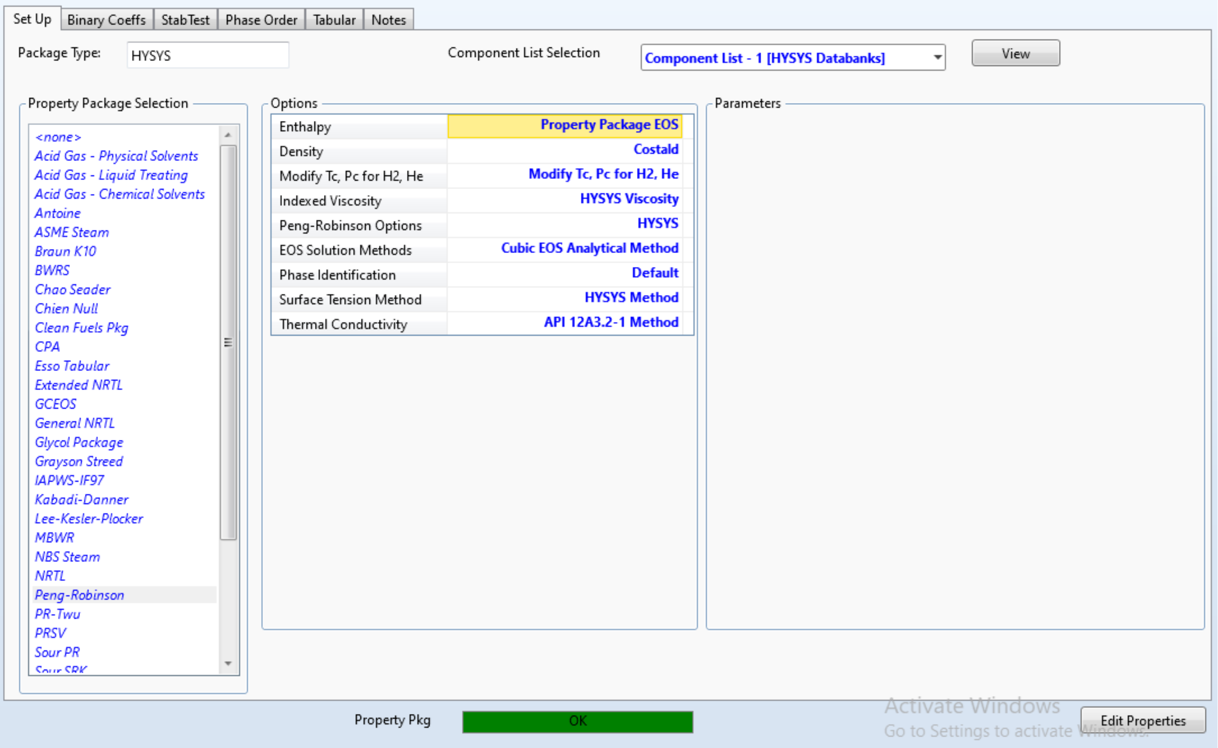
7.4 BINARY COEFFICIENTS

Table 25: Binary Coefficients

Components	Methane	Ethane	Propane	i-Butane	n-Butane	i-Pentane	n-Pentane	H_2O
Methane	---	0.00224	0.00683	0.0131	0.0121	0.0173	0.0176	0.5000
Ethane	0.00224	---	0.00126	0.0045	0.0046	0.0071	0.0076	0.5000
Propane	0.00683	0.00126	---	0.0010	0.0008	0.0025	0.0021	0.48000
i-Butane	0.01311	0.00457	0.00104	---	0.00001	0.0003	0.00034	0.48000
n-Butane	0.01230	0.00410	0.00082	0.0001	---	0.0005	0.0005	0.48000
i-Pentane	0.01763	0.00741	0.00258	0.0003	0.00035	---	0.0000	0.48000
n-Pentane	0.01793	0.00761	0.00270	0.0003	0.0004	0.0000	---	0.48000
H_2O	0.50000	0.50000	0.48000	0.4800	0.4800	0.48000	0.48000	---
NBP[0]114	0.01872	0.00813	0.00302	0.0005	0.0007	0.00002	0.00014	0.48000
NBP[0]139	0.02116	0.00979	0.00407	0.001	0.0012	0.00017	0.00021	0.48000
NBP[0]162	0.02314	0.01118	0.00498	0.0014	0.0017	0.00039	0.00045	0.48000

7.5 FLUID PACKAGE

In order to calculate the properties of the components, Cubic equation of state analytical method is used while their parameters are set as default. For calculation of Enthalpy, fluid package **Peng Robinson** is used.



Package Type: HYSYS

Component List Selection: Component List - 1 [HYSYS Databanks]

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 42: Fluid Package

It is a type of equation state which is mostly used as a function of temperature to calculate entropy, enthalpy at ideal conditions. Every simulation requires some relations to access the values and conditions of the existing flow sheet. It helps in determining relevant fluid properties at provided states. Fluid package selection depends on the type of components used. Peng Robinson is mostly used for hydrocarbon systems

7.6 COLUMN ENVIRONMENT

Vacuum Column environment is shows as follows but it does not have a condenser and reboiler. Only Top and Bottom stage is highly involved whose duties are controlling the reflux.

In the graph shown below, there is pressure drop seen linearly with Tray number of the column. However, it is not possible that graph must be linear, it is variably shown in reality which is impossible for us to design.

Vacuum Column is packed type for proposed simulation flow sheet.

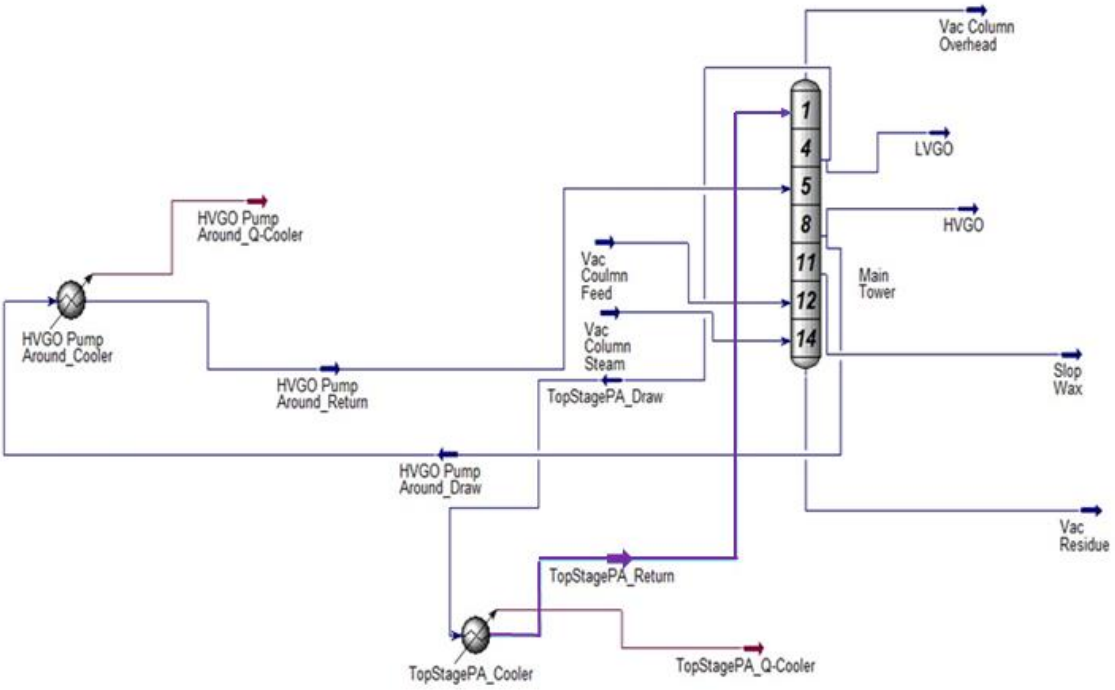
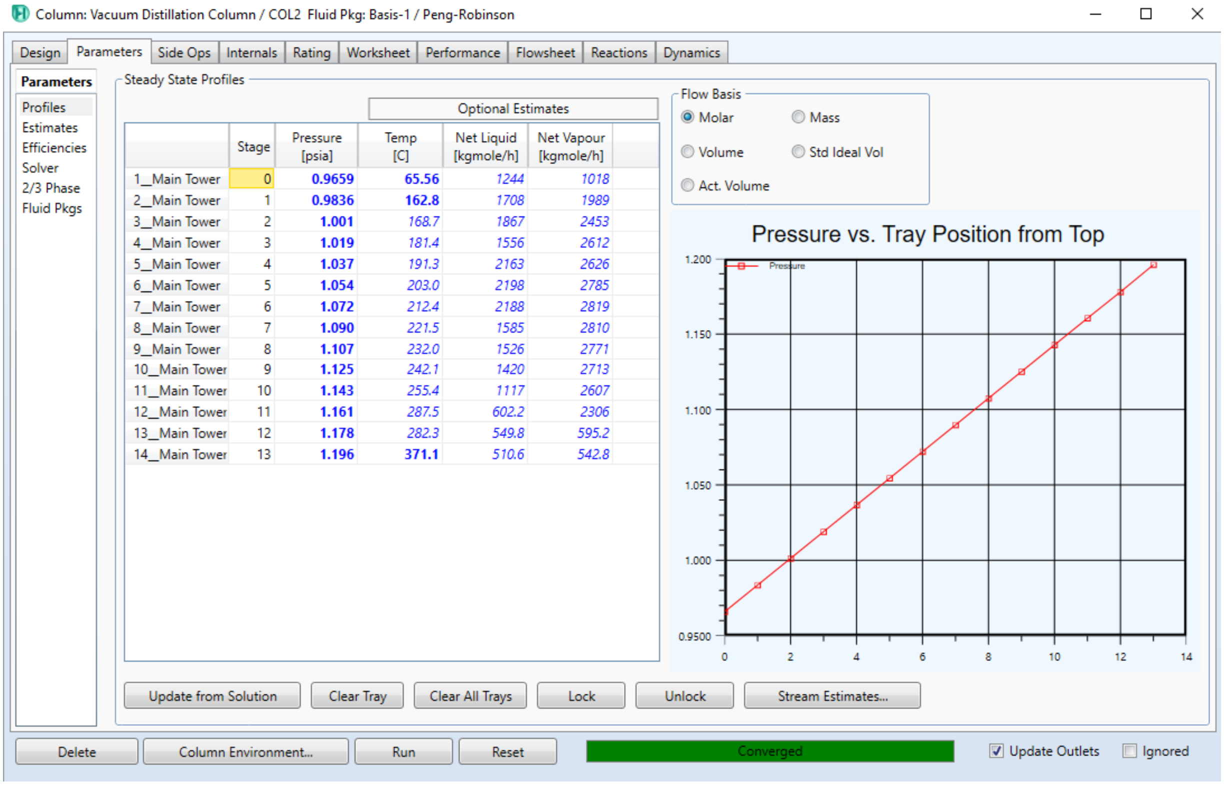


Figure 43: Sub Flow Sheet Vacuum Column Environment

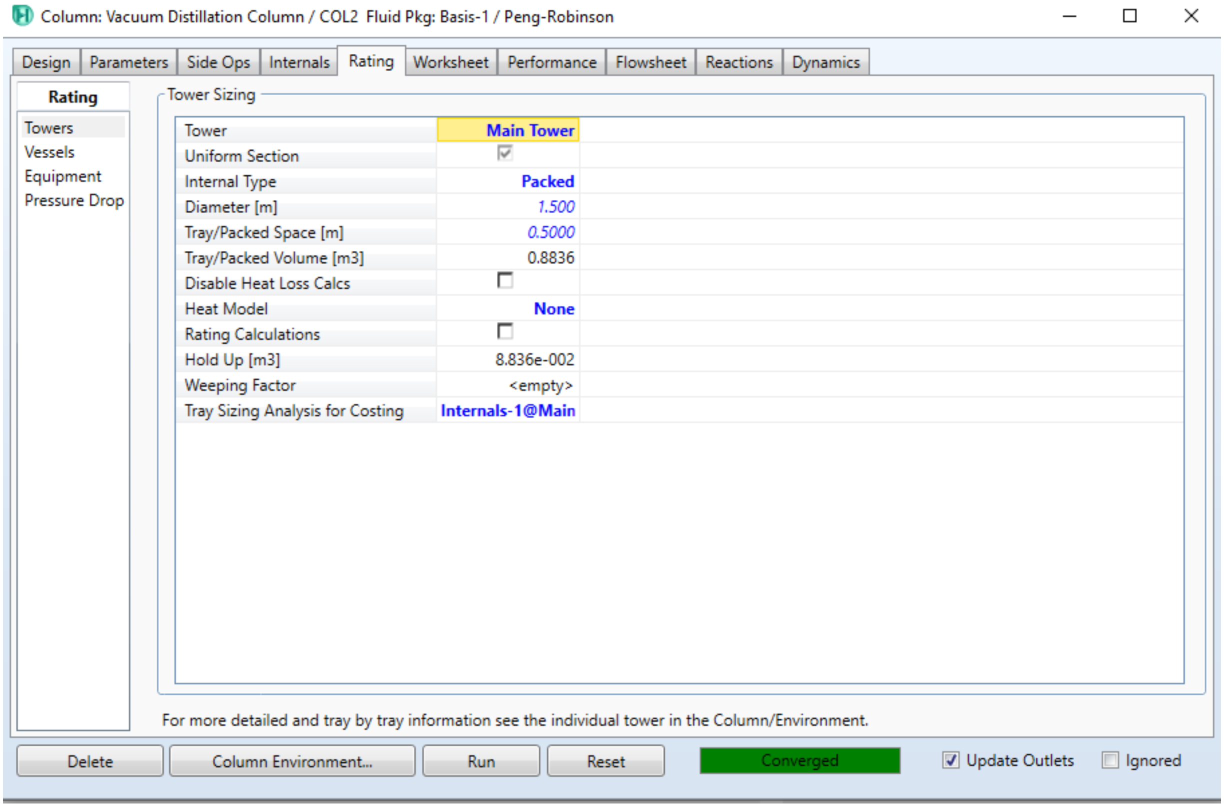


	Stage	Pressure [psia]	Temp [C]	Net Liquid [kgmole/h]	Net Vapour [kgmole/h]
1__Main Tower	0	0.9659	65.56	1244	1018
2__Main Tower	1	0.9836	162.8	1708	1989
3__Main Tower	2	1.001	168.7	1867	2453
4__Main Tower	3	1.019	181.4	1556	2612
5__Main Tower	4	1.037	191.3	2163	2626
6__Main Tower	5	1.054	203.0	2198	2785
7__Main Tower	6	1.072	212.4	2188	2819
8__Main Tower	7	1.090	221.5	1585	2810
9__Main Tower	8	1.107	232.0	1526	2771
10__Main Tower	9	1.125	242.1	1420	2713
11__Main Tower	10	1.143	255.4	1117	2607
12__Main Tower	11	1.161	287.5	602.2	2306
13__Main Tower	12	1.178	282.3	549.8	595.2
14__Main Tower	13	1.196	371.1	510.6	542.8

● Molar ○ Mass
○ Volume ○ Std Ideal Vol
○ Act. Volume

☑ Update Outlets ☐ Ignored

Figure 44: Vacuum Column Performance



Tower	Main Tower
Uniform Section	☑
Internal Type	Packed
Diameter [m]	1.500
Tray/Packed Space [m]	0.5000
Tray/Packed Volume [m3]	0.8836
Disable Heat Loss Calcs	☐
Heat Model	None
Rating Calculations	☐
Hold Up [m3]	8.836e-002
Weeping Factor	<empty>
Tray Sizing Analysis for Costing	Internals-1@Main

For more detailed and tray by tray information see the individual tower in the Column/Environment.

☑ Update Outlets ☐ Ignored

Figure 45: Vacuum Column Environment

7.7 EQUIPMENT DATA

Following were the equipment added in the proposed simulation flow sheet:

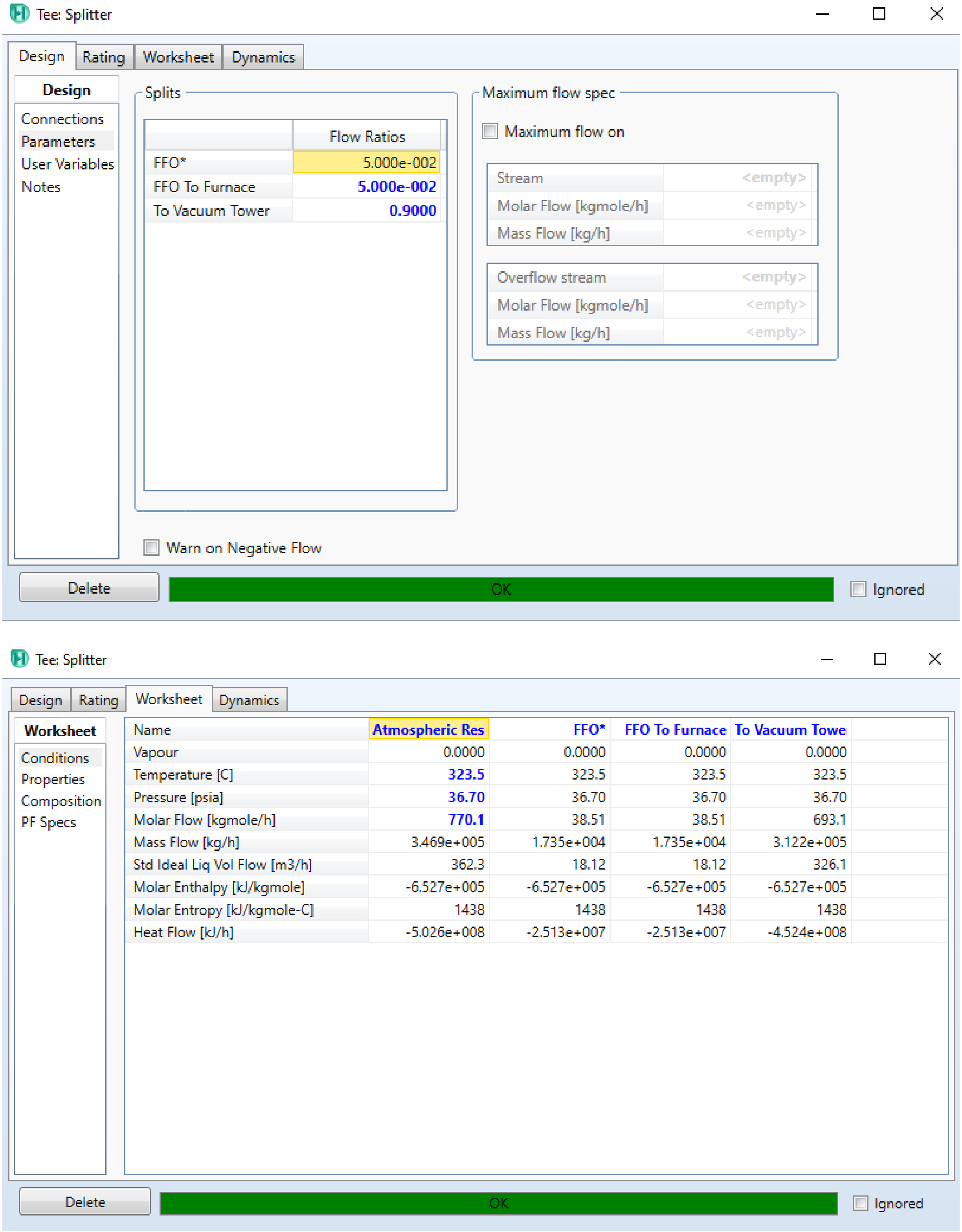
- Splitter
- Mixer
- Preheater
- Pump
- Heat Exchanger
- Vacuum Distillation Column

7.7.1 SPLITTER

Atmospheric Residue was introduced as feed, since in splitter we have 3 streams, 1st stream is going to Vacuum Column named as **To Vacuum Tower**, 2nd stream one goes to Storage of FFO (Furnace Fuel Oil) named as **FFO*** and the last stream goes to furnace named as **FFO To Furnace**. We have provided split ratios for each stream. Split ratios are:

- 0.9 for To **Vacuum Tower**
- 0.05 for **FFO***
- 0.05 for **FFO To Furnace**

The simulated results are tabulated as follows:



	Flow Ratios
FFO*	5.000e-002
FFO To Furnace	5.000e-002
To Vacuum Tower	0.9000

☐ Maximum flow on

Stream	<empty>
Molar Flow [kgmole/h]	<empty>
Mass Flow [kg/h]	<empty>

Overflow stream	<empty>
Molar Flow [kgmole/h]	<empty>
Mass Flow [kg/h]	<empty>

☐ Warn on Negative Flow

☐ Ignored

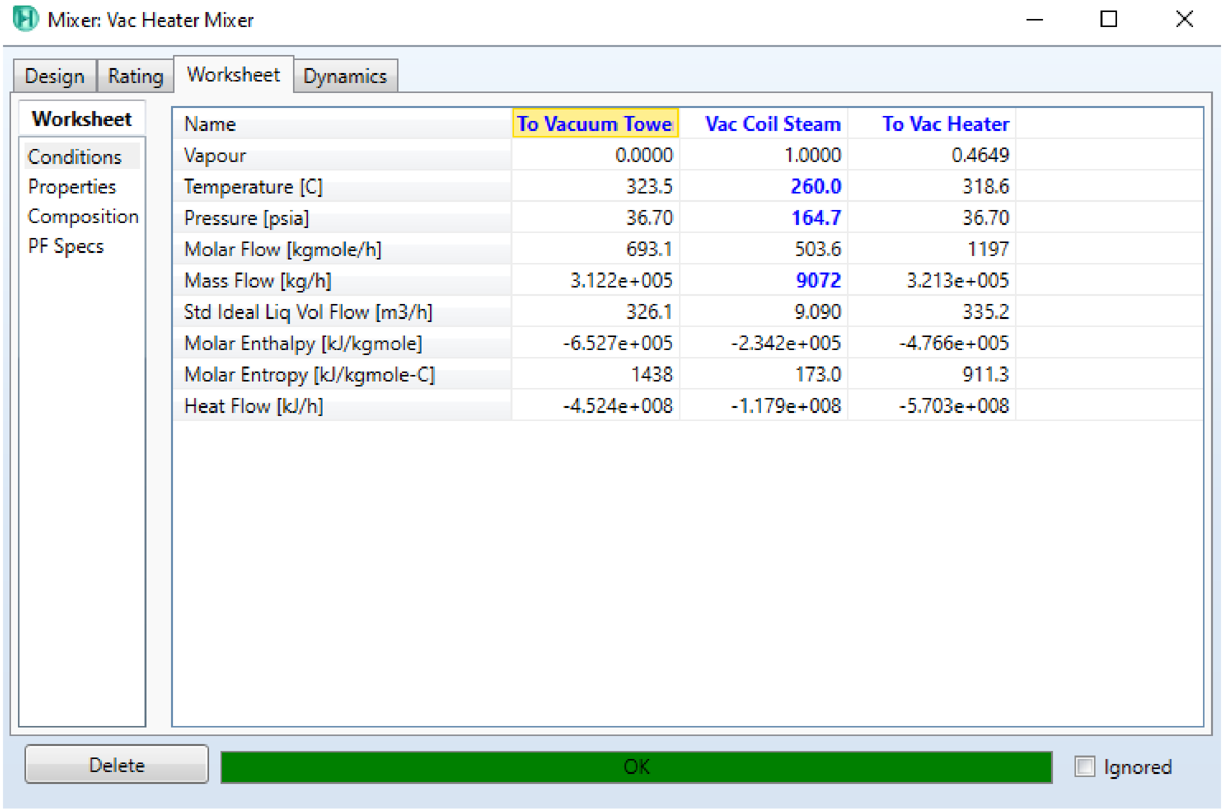
Name	Atmospheric Res	FFO*	FFO To Furnace	To Vacuum Towe
Vapour	0.0000	0.0000	0.0000	0.0000
Temperature [C]	323.5	323.5	323.5	323.5
Pressure [psia]	36.70	36.70	36.70	36.70
Molar Flow [kgmole/h]	770.1	38.51	38.51	693.1
Mass Flow [kg/h]	3.469e+005	1.735e+004	1.735e+004	3.122e+005
Std Ideal Liq Vol Flow [m3/h]	362.3	18.12	18.12	326.1
Molar Enthalpy [kJ/kgmole]	-6.527e+005	-6.527e+005	-6.527e+005	-6.527e+005
Molar Entropy [kJ/kgmole-C]	1438	1438	1438	1438
Heat Flow [kJ/h]	-5.026e+008	-2.513e+007	-2.513e+007	-4.524e+008

☐ Ignored

Figure 46: Splitter (i) (ii)

7.7.2 MIXER

Introducing the steam stream named as **Vac Steam** along with the stream named as **To Vacuum Tower** in the mixer. The simulated results are tabulated as follows:



Mixer: Vac Heater Mixer

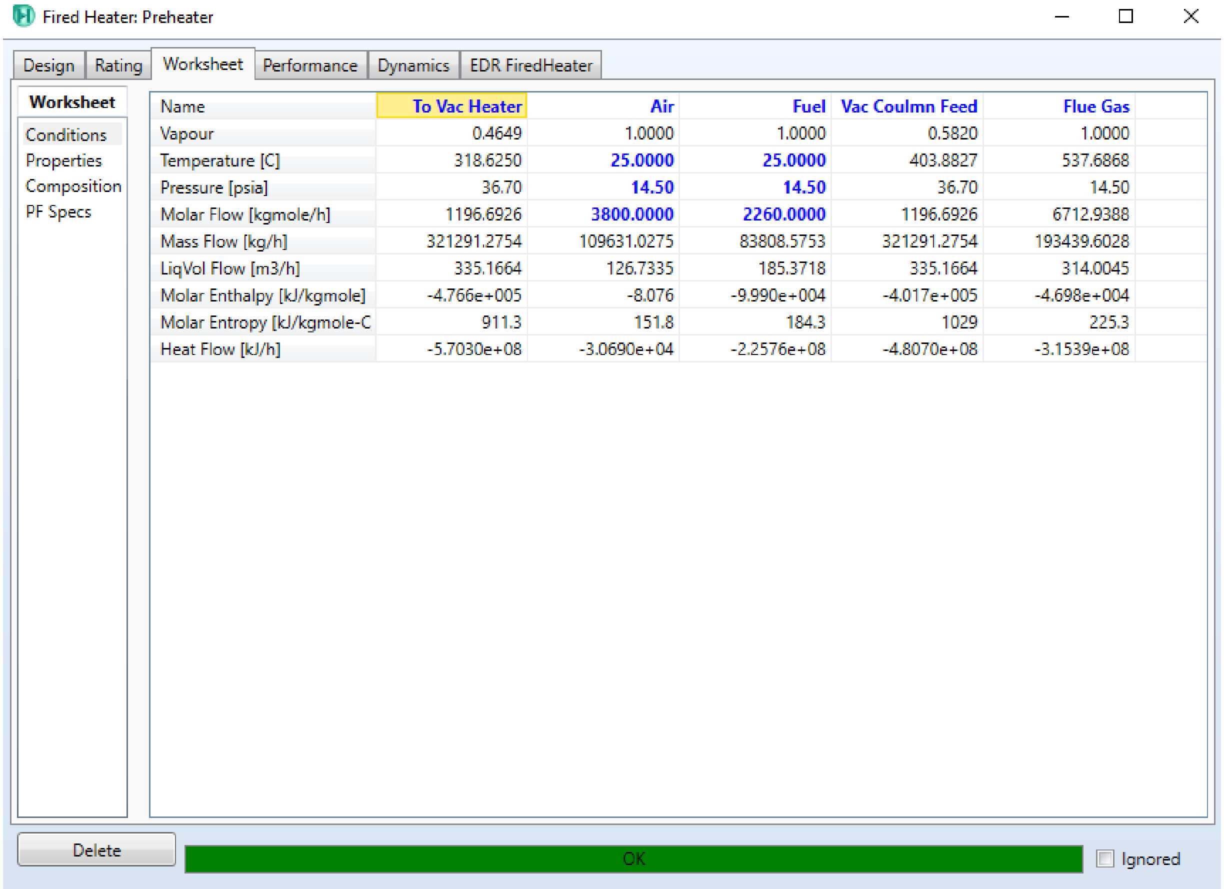
Name	To Vacuum Towe	Vac Coil Steam	To Vac Heater
Vapour	0.0000	1.0000	0.4649
Temperature [C]	323.5	260.0	318.6
Pressure [psia]	36.70	164.7	36.70
Molar Flow [kgmole/h]	693.1	503.6	1197
Mass Flow [kg/h]	3.122e+005	9072	3.213e+005
Std Ideal Liq Vol Flow [m3/h]	326.1	9.090	335.2
Molar Enthalpy [kJ/kgmole]	-6.527e+005	-2.342e+005	-4.766e+005
Molar Entropy [kJ/kgmole-C]	1438	173.0	911.3
Heat Flow [kJ/h]	-4.524e+008	-1.179e+008	-5.703e+008

Figure 47: Mixer

7.7.3 PREHEATER

There is only one main preheater within the proposed process, and it is responsible for providing the heat to the **FFO (Furnace Fuel Oil)** and **Vacuum Steam** mixture.

The simulated results are tabulated as follows:

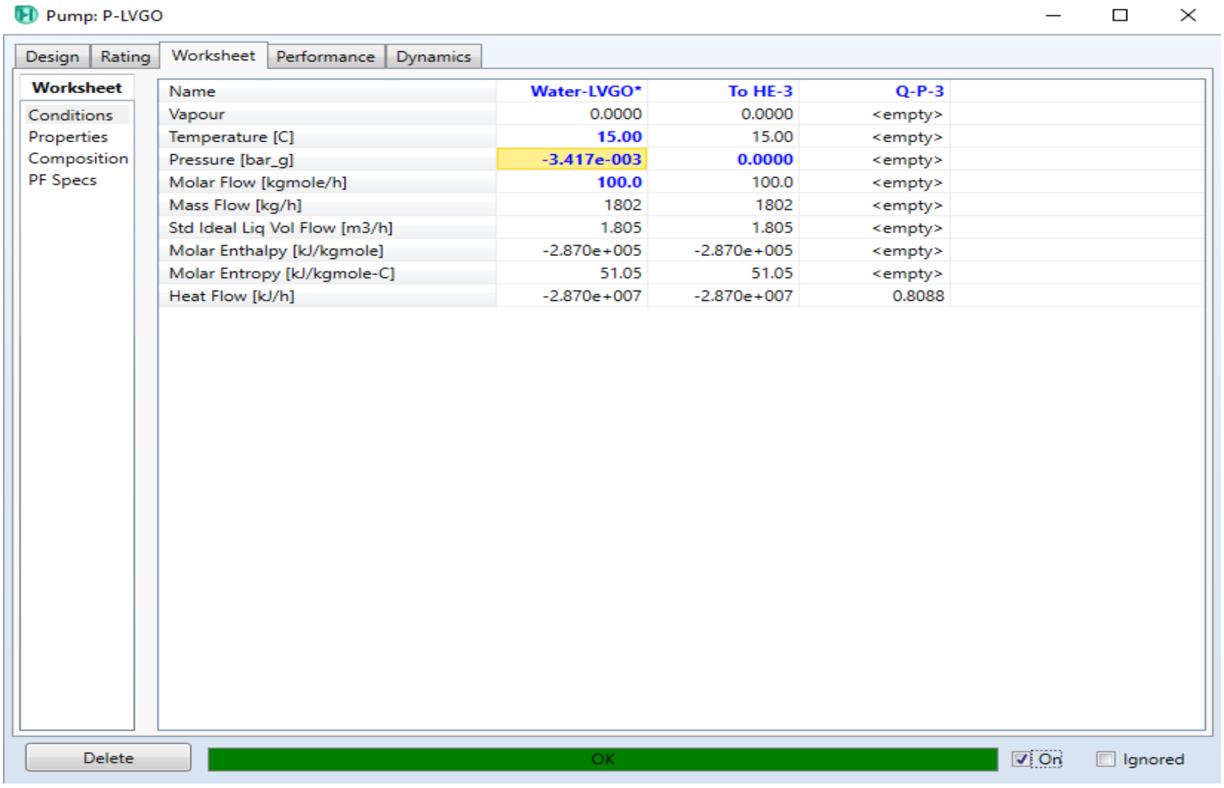


Name	To Vac Heater	Air	Fuel	Vac Coulmn Feed	Flue Gas
Vapour	0.4649	1.0000	1.0000	0.5820	1.0000
Temperature [C]	318.6250	25.0000	25.0000	403.8827	537.6868
Pressure [psia]	36.70	14.50	14.50	36.70	14.50
Molar Flow [kgmole/h]	1196.6926	3800.0000	2260.0000	1196.6926	6712.9388
Mass Flow [kg/h]	321291.2754	109631.0275	83808.5753	321291.2754	193439.6028
LiqVol Flow [m3/h]	335.1664	126.7335	185.3718	335.1664	314.0045
Molar Enthalpy [kJ/kgmole]	-4.766e+005	-8.076	-9.990e+004	-4.017e+005	-4.698e+004
Molar Entropy [kJ/kgmole-C	911.3	151.8	184.3	1029	225.3
Heat Flow [kJ/h]	-5.7030e+08	-3.0690e+04	-2.2576e+08	-4.8070e+08	-3.1539e+08

Figure 48: Preheater

7.7.4 PUMP

Six main pumps were utilized for the proposed simulation flow sheet; their function was to provide water to the heat exchangers product while using water as a source of utility. The simulated results are tabulated as follows:

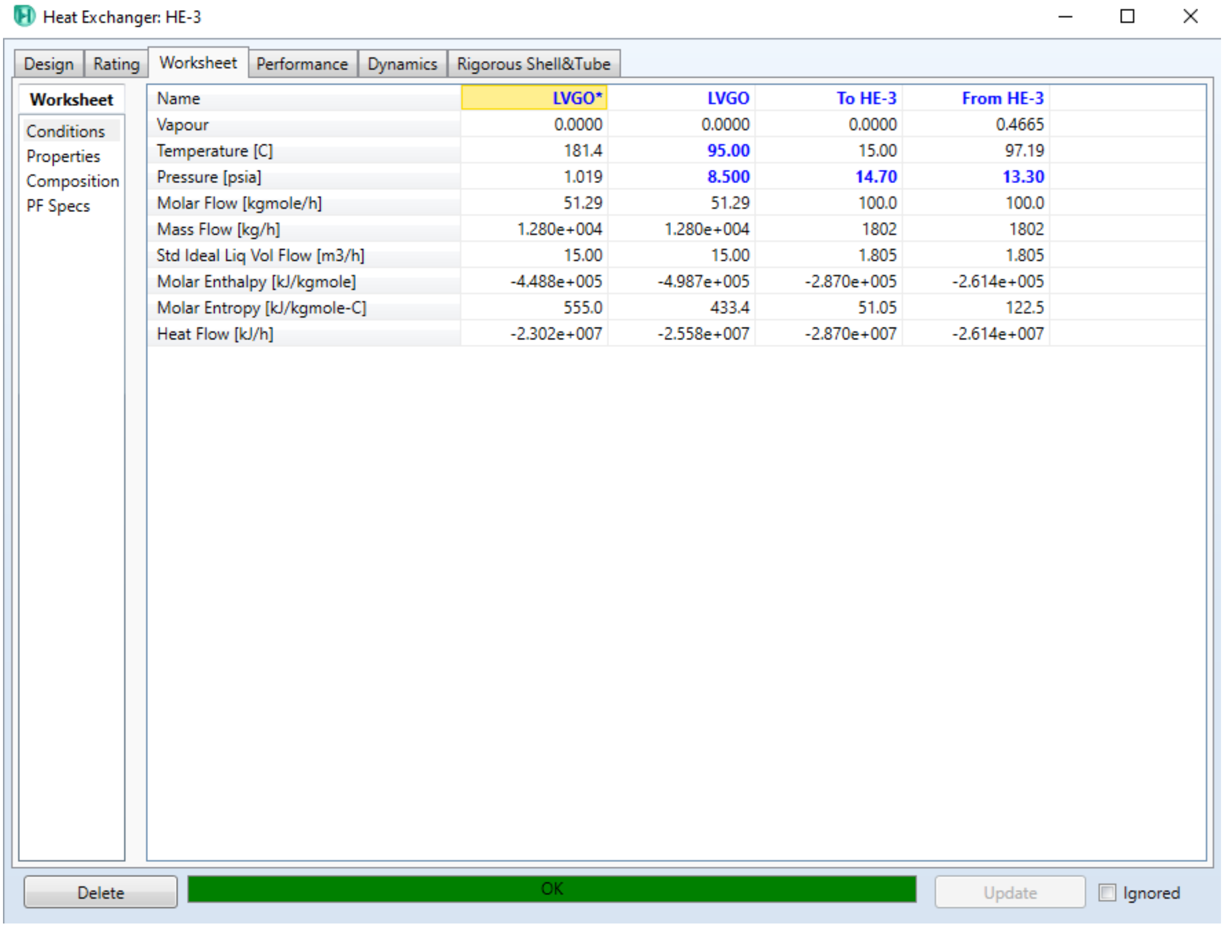


Name	Water-LVGO*	To HE-3	Q-P-3
Vapour	0.0000	0.0000	<empty>
Temperature [C]	15.00	15.00	<empty>
Pressure [bar_g]	-3.417e-003	0.0000	<empty>
Molar Flow [kgmole/h]	100.0	100.0	<empty>
Mass Flow [kg/h]	1802	1802	<empty>
Std Ideal Liq Vol Flow [m3/h]	1.805	1.805	<empty>
Molar Enthalpy [kJ/kgmole]	-2.870e+005	-2.870e+005	<empty>
Molar Entropy [kJ/kgmole-C]	51.05	51.05	<empty>
Heat Flow [kJ/h]	-2.870e+007	-2.870e+007	0.8088

Figure 49: LVGO pump

7.7.5 HEAT EXCHANGER

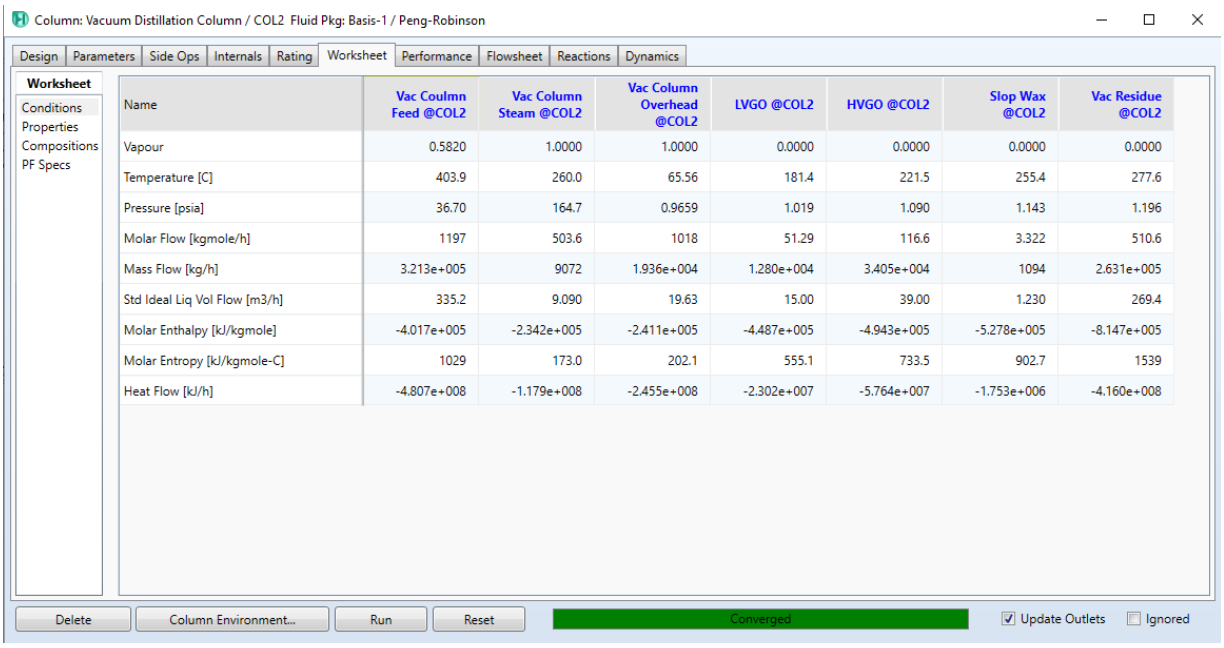
To obtain the heat loads involved for every heat exchanger used in the proposed simulation flow sheet, simulated results are tabulated as follows:



Name	LVGO*	LVGO	To HE-3	From HE-3
Vapour	0.0000	0.0000	0.0000	0.4665
Temperature [C]	181.4	95.00	15.00	97.19
Pressure [psia]	1.019	8.500	14.70	13.30
Molar Flow [kgmole/h]	51.29	51.29	100.0	100.0
Mass Flow [kg/h]	1.280e+004	1.280e+004	1802	1802
Std Ideal Liq Vol Flow [m3/h]	15.00	15.00	1.805	1.805
Molar Enthalpy [kJ/kgmole]	-4.488e+005	-4.987e+005	-2.870e+005	-2.614e+005
Molar Entropy [kJ/kgmole-C]	555.0	433.4	51.05	122.5
Heat Flow [kJ/h]	-2.302e+007	-2.558e+007	-2.870e+007	-2.614e+007

Figure 50: LVGO Heat Exchanger

7.7.6 VACUUM DISTILLATION COLUMN



Name	Vac Coulmn Feed @COL2	Vac Column Steam @COL2	Vac Column Overhead @COL2	LVGO @COL2	HVGO @COL2	Slop Wax @COL2	Vac Residue @COL2
Vapour	0.5820	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000
Temperature [C]	403.9	260.0	65.56	181.4	221.5	255.4	277.6
Pressure [psia]	36.70	164.7	0.9659	1.019	1.090	1.143	1.196
Molar Flow [kgmole/h]	1197	503.6	1018	51.29	116.6	3.322	510.6
Mass Flow [kg/h]	3.213e+005	9072	1.936e+004	1.280e+004	3.405e+004	1094	2.631e+005
Std Ideal Liq Vol Flow [m3/h]	335.2	9.090	19.63	15.00	39.00	1.230	269.4
Molar Enthalpy [kJ/kgmole]	-4.017e+005	-2.342e+005	-2.411e+005	-4.487e+005	-4.943e+005	-5.278e+005	-8.147e+005
Molar Entropy [kJ/kgmole-C]	1029	173.0	202.1	555.1	733.5	902.7	1539
Heat Flow [kJ/h]	-4.807e+008	-1.179e+008	-2.455e+008	-2.302e+007	-5.764e+007	-1.753e+006	-4.160e+008

Figure 51: Vacuum Distillation Column

Table 26: Overall material Balance by Aspen HYSYS

COMPONENTS	**IN**				**OUT**		
(kg/h)	**Atmospheric Residue**	**Water-FFO***	**Air**	**Fuel**	**From HE-1**	**Asphalt**	**Flue Gas**
Methane	0.00	0.00	0.00	25142.4	0.00	0.00	19537.3
Ethane	0.00	0.00	0.00	25142.4	0.00	0.00	19537.3
Propane	0.00	0.00	0.00	8380.8	0.00	0.00	6518.8
n-Butane	0.00	0.00	0.00	8380.8	0.00	0.00	6518.8
i-Butane	0.00	0.00	0.00	8380.8	0.00	0.00	6518.8
n-Pentane	0.00	0.00	0.00	8380.8	0.00	0.00	2340.6
i-Pentane	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	119.3976	1802.0	0.00	0.00	1802.0	2.89	25089
C5-C13 (hydrocarbons)	248.989	0.00	0.00	0.00	0.00	0.00	0.00
C15-C24 (hydrocarbons)	61377.379	0.00	0.00	0.00	0.00	10729.9	0.00
C25-C33 (hydrocarbons)	104437.9	0.00	0.00	0.00	0.00	89756.9	0.00
C34-C43 (hydrocarbons)	137672	0.00	0.00	0.00	0.00	123785.5	0.00
C45-C51 (hydrocarbons)	43054	0.00	0.00	0.00	0.00	38795.9	0.00
O_2	0.00	0.00	23022.0	0.00	0.00	0.00	0.00
N_2	0.00	0.00	86608.0	0.00	0.00	0.00	86505.9
CO	0.00	0.00	0.00	0.00	0.00	0.00	20910.7
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL	**346,909**	**1802**	**109,631**	**83,808**	**1802**	**263,071**	**193,439**

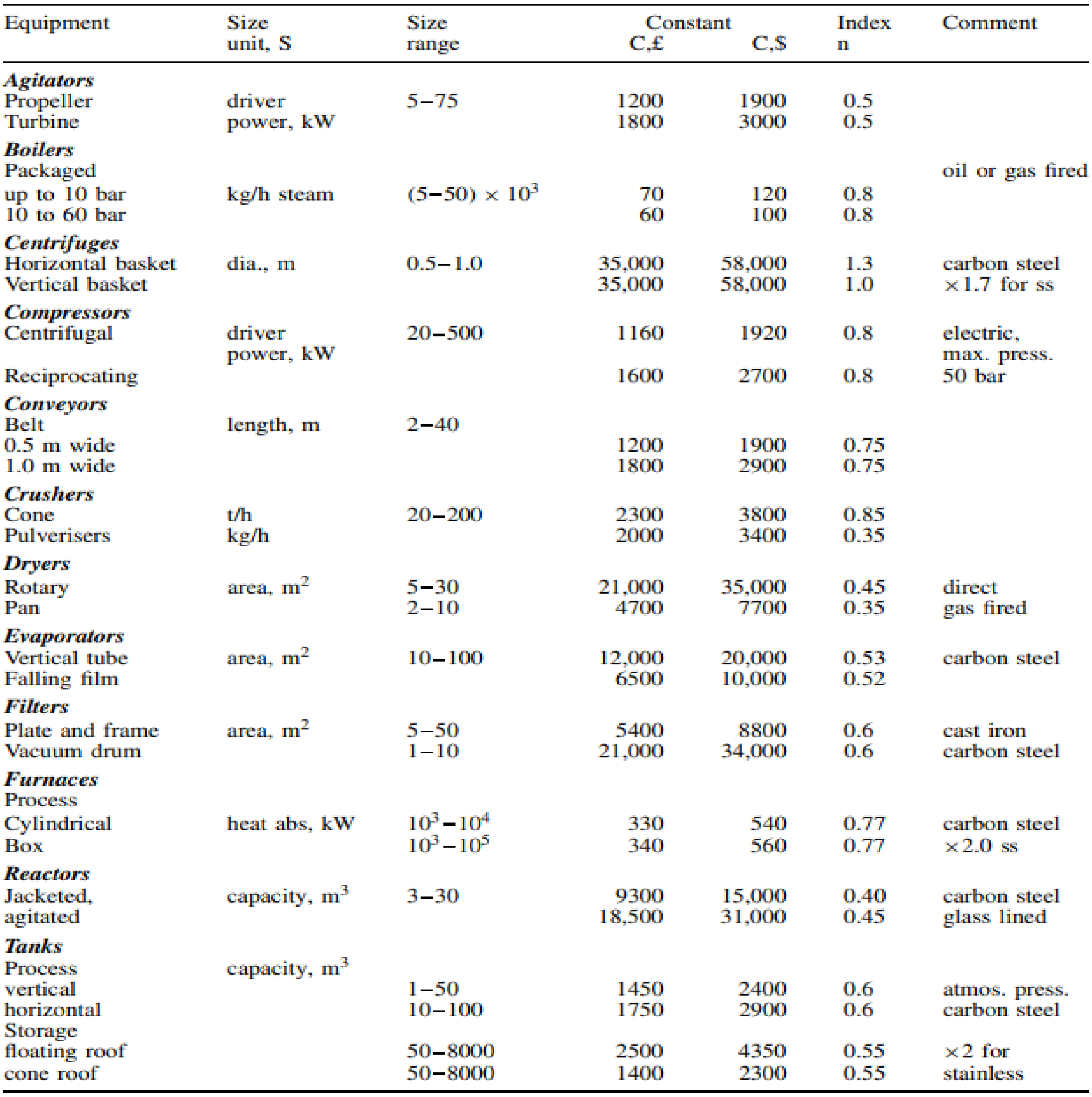
Table 27: Overall Energy Balance by Aspen HYSYS

COMPONENTS		**IN**				**OUT**		
	units	**Atmospheric Residue**	**Q-P-1**	**Q-P-2**	**Q-Vac Preheater**	**HVGO**	**Slop Oil**	**Asphalt**
Vapor Fraction		0.00	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	*C*	323.5	0.00	0.00	0.00	217.8	196.9	271.4
Pressure	*psi*	36.7	0.00	0.00	0.00	10.00	10.00	8.00
Heat Flow	*kJ/h*	-5.02E+08	0.00	0.00	0.00	-5.79E+07	-1.948E+06	-4.201E+08
Mass Heat Capacity	*kJ/kg-C*	2.88	0.00	0.00	0.00	2.61	2.51	2.71
Duty	*kW*		2.247	2.247	2.48E+04			

CHAPTER 8

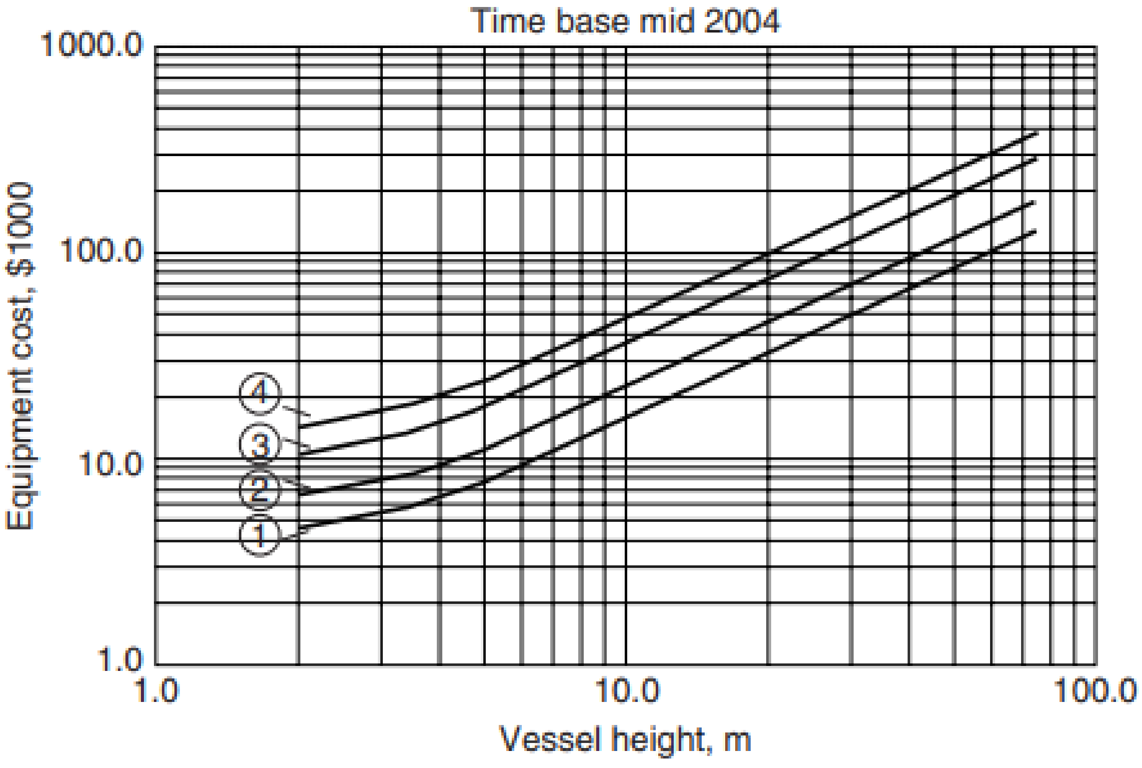
ECONOMIC ANALYSIS

Cost is one of the most important parameters in efficient management of a project. Project costing decides whether the project is feasible or not. Economic analysis is very important to financially evaluate the feasibility of project and all of its components. Since main goal of a project is to do business and generate revenue so, cost becomes the major factor for the decision of equipment installation. There is usually a trade-off between time and quality, and price of a project for desired results. We have selected Coulson and Richardson method for our Project's economic analysis.



Equipment	Size unit, S	Size range	Constant C,£	C,$	Index n	Comment
Agitators						
Propeller	driver	5–75	1200	1900	0.5	
Turbine	power, kW		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,	capacity, m^3	3–30	9300	15,000	0.40	carbon steel
agitated			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 52: Purchased equipment cost



(*b*) US dollars

Diameter, m	
① —0.5	③ —2.0
② —1.0	④ —3.0

Material factors	
C.S.	× 1.0
S.S.	× 2.0
Monel	× 3.4
S.S. clad	× 1.5
Monel clad	× 2.1

Pressure factors	
1–5 bar	× 1.0
5–10	× 1.1
10–20	× 1.2
20–30	× 1.4
30–40	× 1.6
40–50	× 1.8
50–60	× 2.2

Figure 53: Vessel PCE

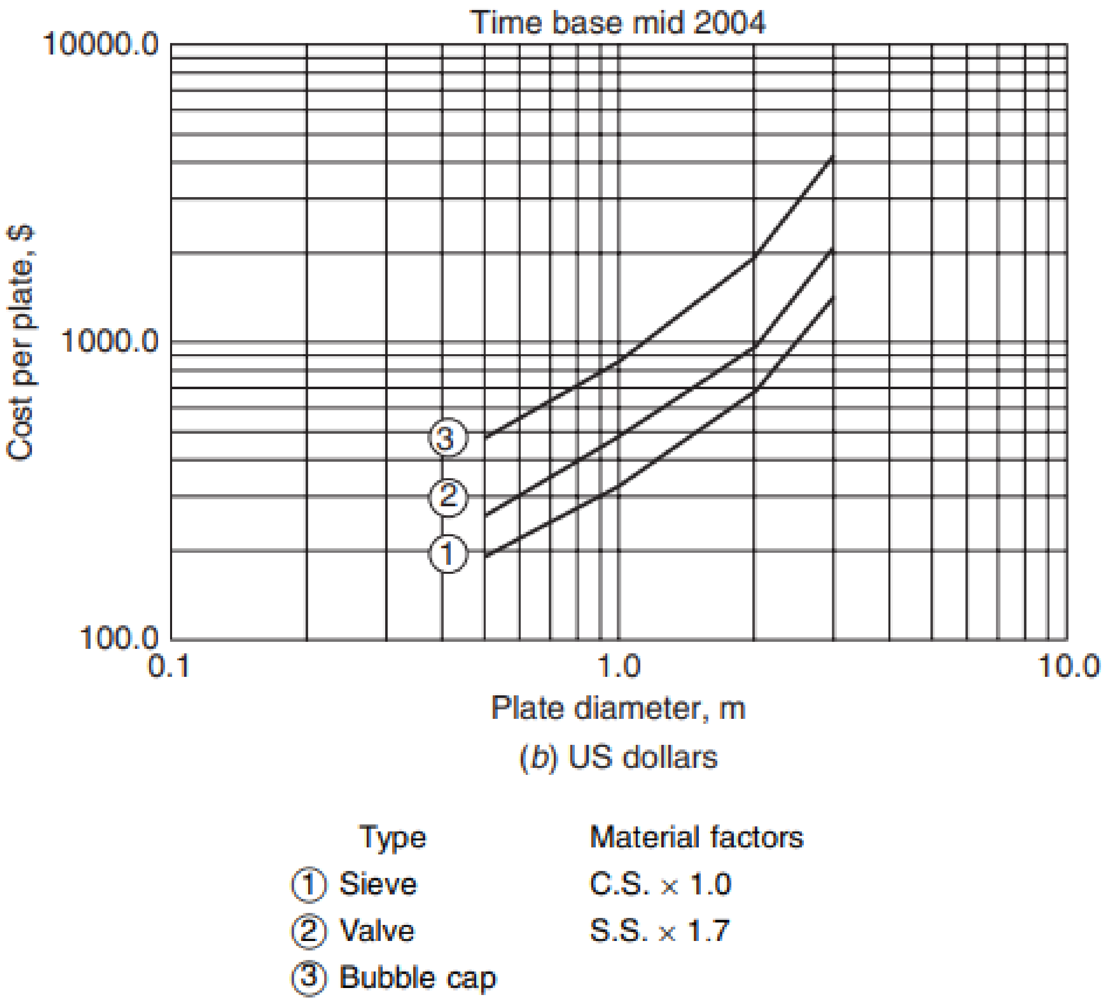


Figure 54: Plate PCE

8.1 VACUUM DISTILLATION COLUMN

Diameter (d) = 1.5 m

Height (h) = 14.4 m

Stages (n) = 14 (Bubble cap)

- Middle two curves' data were interpolated for 1.5 m diameter.
- Since we used stainless steel material because of cost factor, therefore material factor is 2.
- Pressure factor is also 1 for 1-5 bar pressure.
- Material factor is 1.7 for stainless steel plates.

Total cost of Vacuum Distillation Column (2004)

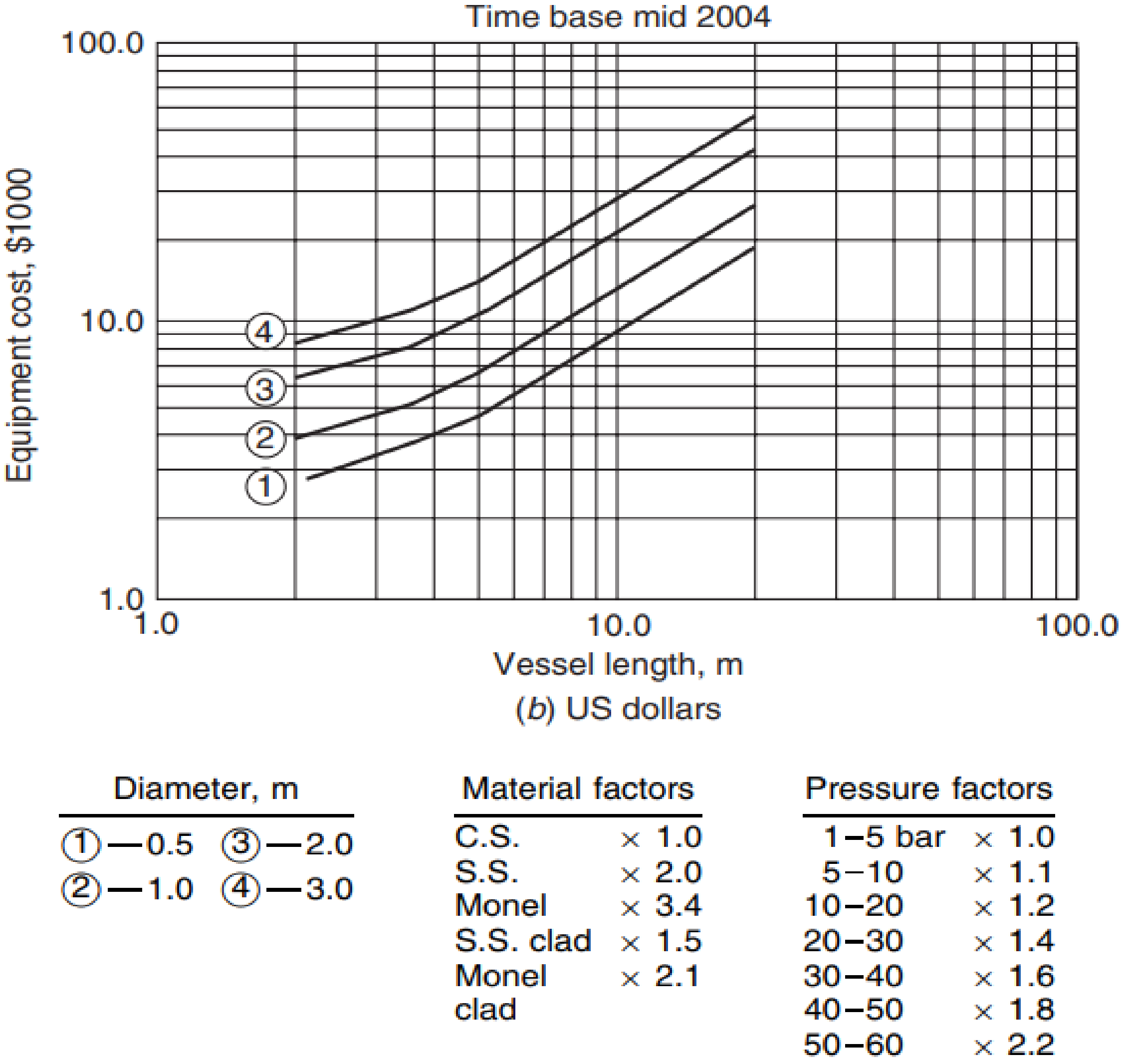
$ 40000*2*1 + 1600*1.7*14 + 120*11072^0.8 + 302000 = $ 617334

Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 979000



Diameter, m

① — 0.5 ③ — 2.0
② — 1.0 ④ — 3.0

Material factors	
C.S.	× 1.0
S.S.	× 2.0
Monel	× 3.4
S.S. clad	× 1.5
Monel clad	× 2.1

Pressure factors	
1–5 bar	× 1.0
5–10	× 1.1
10–20	× 1.2
20–30	× 1.4
30–40	× 1.6
40–50	× 1.8
50–60	× 2.2

Figure 55: Vessel length PCE

8.2 SPLITTER

Diameter (d) = 0.83 m
Length (l) = 2.43 m

- Bottom two curves' data were interpolated for 0.83 m diameter.
- Since we used stainless steel material because of cost factor, therefore material factor is 2.
- Pressure factor is also 1 for 1-5 bar pressure.

Total cost of Splitter (2004) =$ 3600*2*1 = $ 7200

Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 11500

8.3 MIXER

Diameter (d) = 0.94 m
Length (l) = 2.21 m

- Bottom two curves' data were interpolated for 0.94 m diameter.
- Since we used stainless steel material because of cost factor, therefore material factor is 2.
- Pressure factor is 1.2 for 10-20 bar pressure.

Total cost of Mixer (2004) =$ 3800*2*1.2 = $ 9120

Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

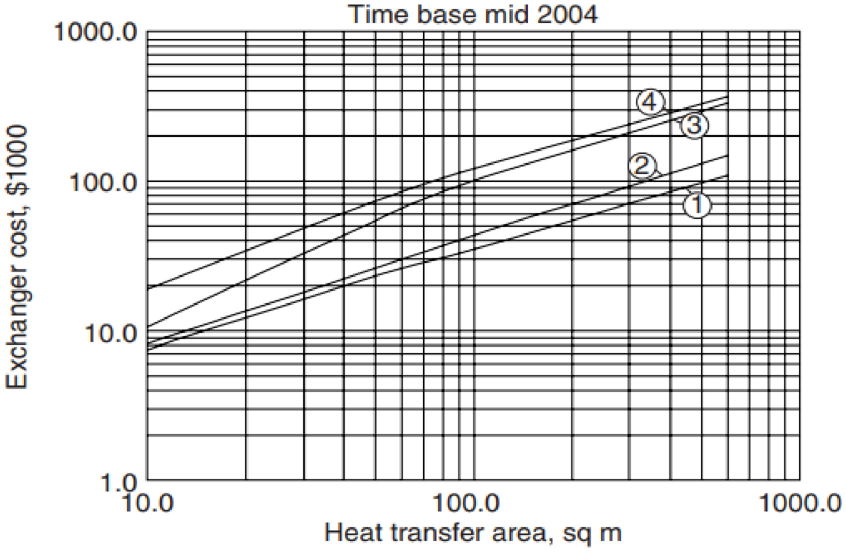
Total cost in 2022 = $ 14500

8.4 PREHEATER

Heat = 24877 kW

Cost = CS^n = $ 540*$24877^{0.77}$ = $ 1310000

8.5 HEAT EXCHANGERS



(b) US dollars

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

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

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

Figure 56: Shell and Tube Heat Exchanger PCE

8.5.1 HEAT EXCHANGER-1

Heat transfer area = 64.3 m^2

- Material is carbon steel for shell and stainless steel for tubes.
- Material factor = 3
- Pressure factor = 1 for 1-10 bar pressure
- Type factor = 0.85 for U-tube

Cost in 2004 =$ 70000*1*0.85 = $ 59500

Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 94500

8.5.2 HEAT EXCHANGER-2

Heat transfer area = 55.7 m^2

- Material is carbon steel for shell and stainless steel for tubes.
- Material factor = 3
- Pressure factor = 1 for 1-10 bar pressure
- Type factor = 0.85 for U-tube

Cost in 2004 =$ 64000*1*0.85 = $ 54400

Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 86300

8.5.3 HEAT EXCHANGER-3

Heat transfer area = 54.6 m^2

- Material is carbon steel for shell and stainless steel for tubes.
- Material factor = 3
- Pressure factor = 1 for 1-10 bar pressure
- Type factor = 0.85 for U-tube

Cost in 2004 =$ 60000*1*0.85 = $ 59500

Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 81000

8.5.4 HEAT EXCHANGER-4

Heat transfer area = 53.3 m^2

- Material is carbon steel for shell and stainless steel for tubes.
- Material factor = 3
- Pressure factor = 1 for 1-10 bar pressure
- Type factor = 0.85 for U-tube

Cost in 2004 =$ 57000*1*0.85 = $ 48500
Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 76900

8.5.5 HEAT EXCHANGER-5

Heat transfer area = 53.9 m^2

- Material is carbon steel for shell and stainless steel for tubes.
- Material factor = 3
- Pressure factor = 1 for 1-10 bar pressure
- Type factor = 0.85 for U-tube

Cost in 2004 =$ 58500*1*0.85 = $ 50000

Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 79300

8.5.6 HEAT EXCHANGER-6

Heat transfer area = 60.4 m^2

- Material is carbon steel for shell and stainless steel for tubes.
- Material factor = 3
- Pressure factor = 1 for 1-10 bar pressure
- Type factor = 0.85 for U-tube

Cost in 2004 =$ 68000*1*0.85 = $ 57800

Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 91700

8.5.7 HEAT EXCHANGER-7

Heat transfer area = 52.8 m^2

- Material is carbon steel for shell and stainless steel for tubes.
- Material factor = 3
- Pressure factor = 1 for 1-10 bar pressure
- Type factor = 0.85 for U-tube

Cost in 2004 =$ 52000*1*0.85 = $ 44200

Index in 2004 = 111
Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 71000

8.6 PUMP-AROUND COOLERS

8.6.1 LVGO PUMP-AROUND COOLER

Heat transfer area = 27.4 m^2

- Material is carbon steel for shell and stainless steel for tubes.
- Material factor = 3
- Pressure factor = 1 for 1-10 bar pressure
- Type factor = 0.85 for U-tube

Cost in 2004 =$ 30000*1*0.85 = $ 25500

Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 40500

8.6.2 HVGO PUMP-AROUND COOLER

Heat transfer area = 26.2 m^2

- Material is carbon steel for shell and stainless steel for tubes.
- Material factor = 3
- Pressure factor = 1 for 1-10 bar pressure
- Type factor = 0.85 for U-tube

Cost in 2004 =$ 27000*1*0.85 = $ 23000

Index in 2004 = 111

Index in 2022 = 176

Cost in 2022 = $\frac{Cost\ in\ 2004\ x\ index\ in\ 2022}{index\ in\ 2004}$

Total cost in 2022 = $ 36500

8.7 PUMPS

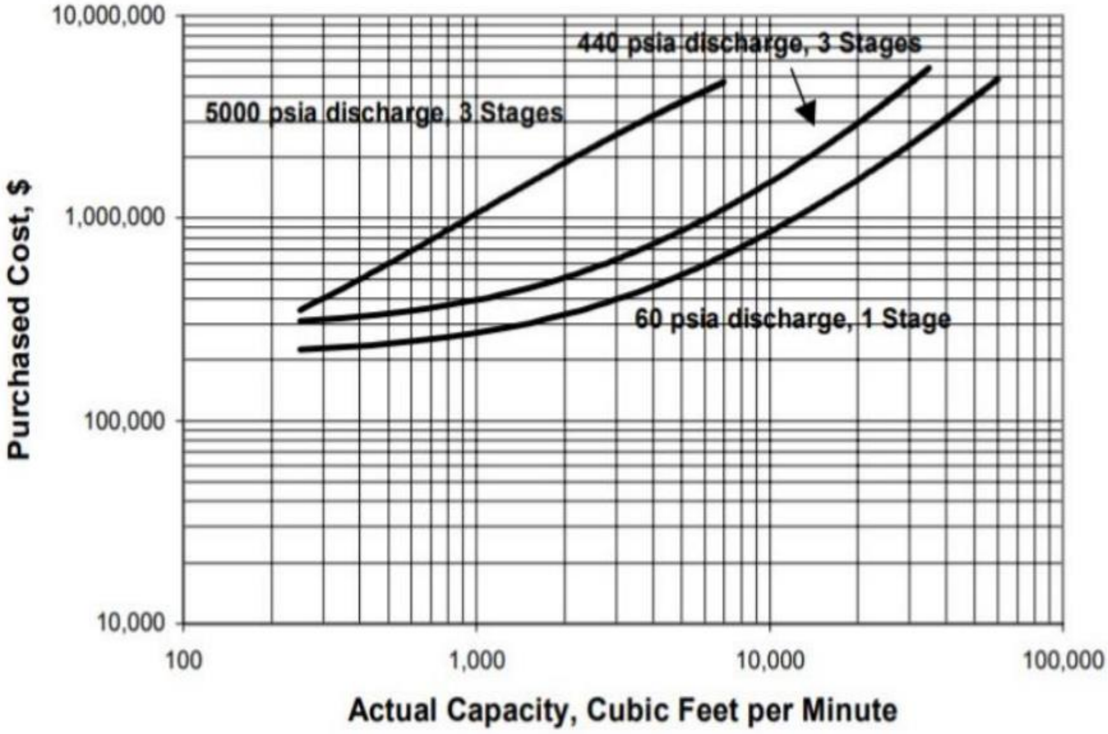


Figure 57: Pumps PCE

ALL PUMPS

Actual Capacity: 1.06 ft^3/min

Less than 60 psi discharge

Single stage Centrifugal pump

Total pump cost = $ 4100

Cost index in 1998 = 5920

Cost index today = 6360

Cost in 2022 = 4100 x 6360/5920

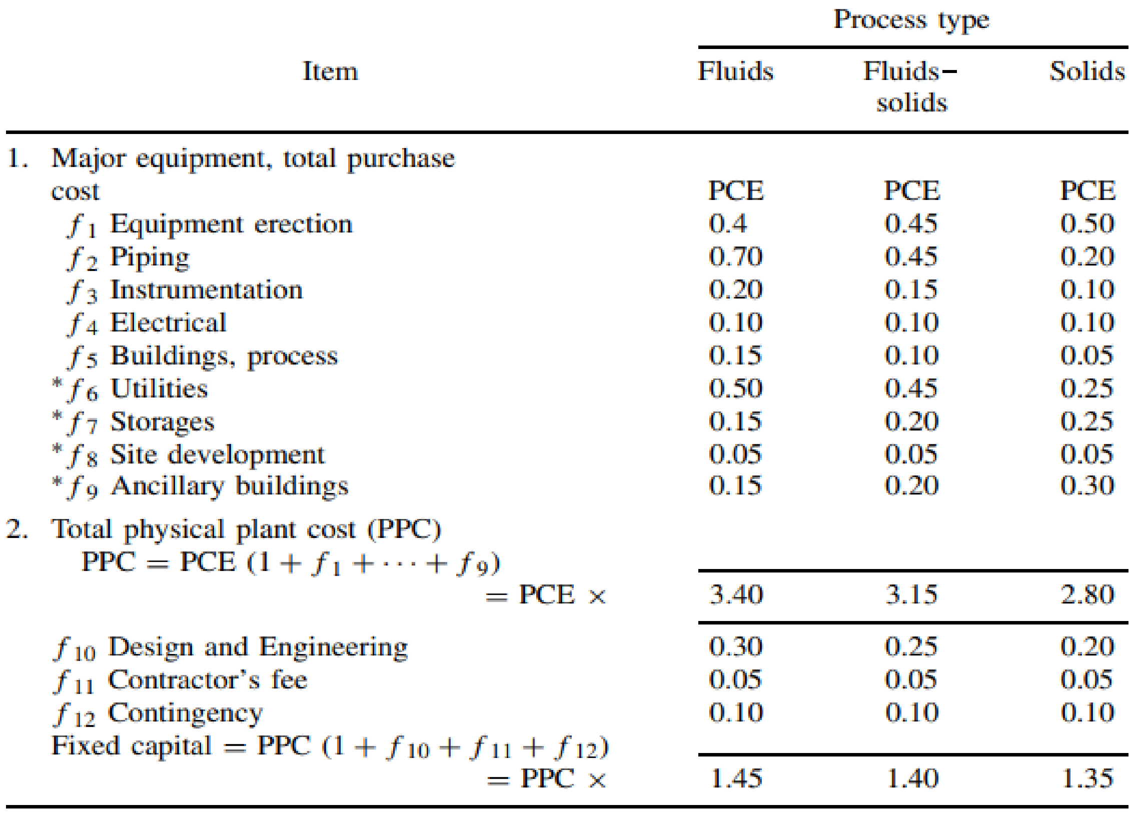
Cost of 1 pump in 2022 = $ 4500

Cost of 6 pumps in 2022 = $ 4500*6 = $ 27000

8.8 STEAM EJECTORS

$90000

8.9 TOTAL COST



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

Figure 58: Fixed capital calculation

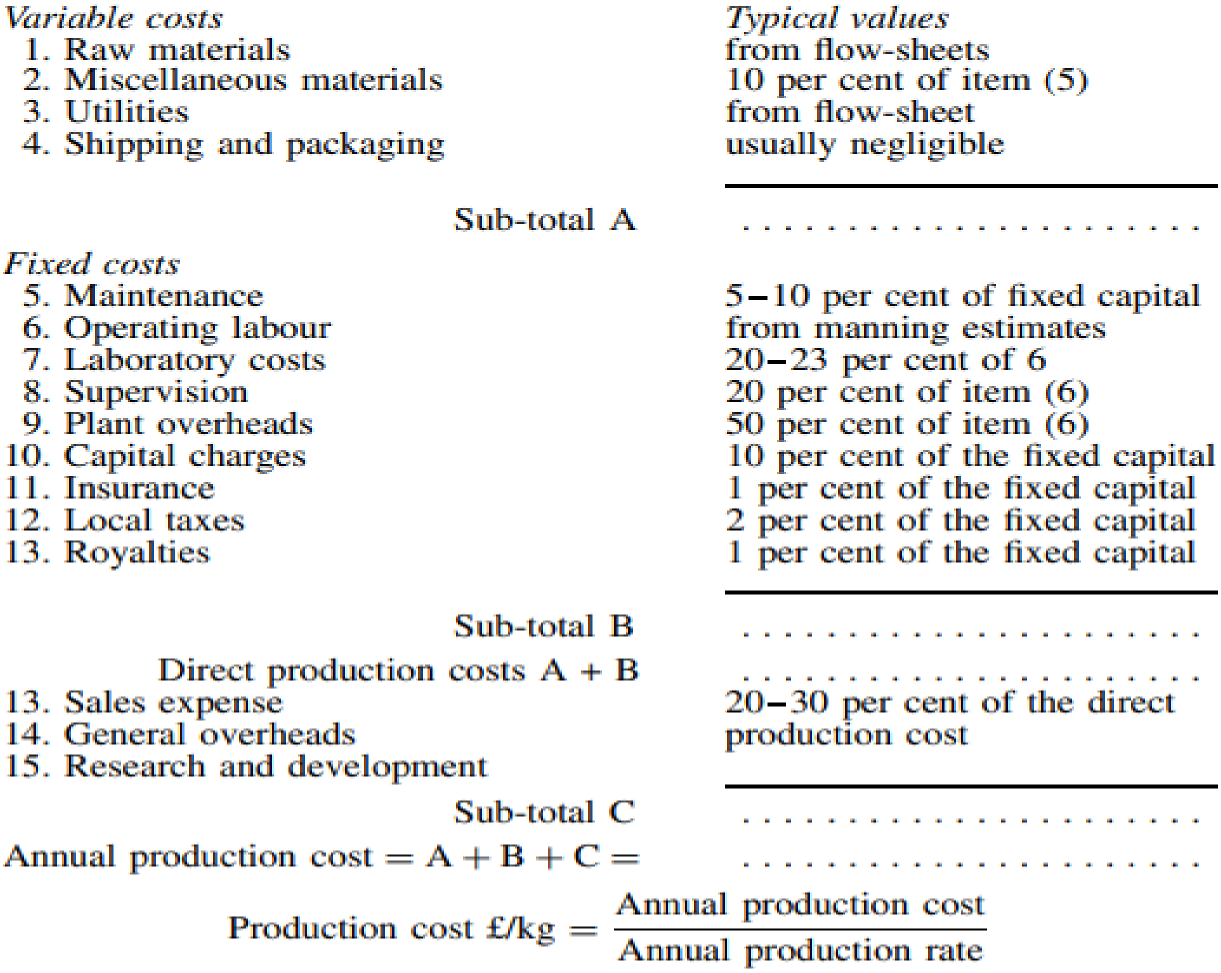
Total PCE = $ 3090500

Physical Plant Cost	
Equipment erection	0.4
Piping	0.7
Instrumentation	0.2
Electrical	0.1
Process	0.15
Storages	0.15
Physical Plant Cost	**$ 8336500**

Table 28: Physical plant cost

Design and Engineering	**0.3**
Contingencies	0.1
Contractor's fees	0.05
Fixed Capital Cost	**$ 12086500**

Table 29: Total investment



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 14. General overheads 15. Research and development	20–30 per cent of the direct production cost
Sub-total C	. . . .
Annual production cost = A + B + C =	. . . .

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

Figure 59: Fixed and variable cost

Fixed Operating Cost	
Maintenance	605000
Operating Cost	854000
Labour	432000
Total	**$ 1895500**

Variable Operating Cost	
Raw Materials	3596000
Miscellaneous	60500
Utilities	103200
Total	**$ 3764500**

Table 30: Fixed and variable cost

Direct Production Costs = $ 5655500

Sales Expense = $ 1135500

Annual Operating Cost = $ 6786500

Annual Profit = Annual Revenue – Annual Operating Cost

Rate of Return (ROR) = Annual Profit/Fixed Capital

Payback Period = 1/ROR [10]

8.10 ECONOMIC ANALYSIS SUMMARY

Purchased Equipment Costs	
Vacuum Distillation Column	979000
Splitter	11500
Mixer	14500
HE-1	94500
HE-2	86300
HE-3	81000
HE-4	76900
HE-5	79300
HE-6	91700
HE-7	71000
LVGO Pump-around Cooler	40500
HVGO Pump-around Cooler	36500
Pumps	27000
Steam Ejectors	90000
Preheater	1310000
Total	**$ 3090500**

Table 31: Purchased equipment cost summary

Summary	
Purchased Equipment Cost (PCE)	3090500
Physical Plant Cost (PPC)	8336500
Fixed Capital	12086500
Fixed Operating Cost	1895500
Variable Operating Cost	3764500

Annual Operating Cost	6786500
Annual Revenue	9996500
Annual Profit	3214500
ROR	0.266
Payback Time	**3.76 years**

Table 32: Economic analysis summary

CHAPTER 9

CONTROL LOOP

9.1 TEMPERATURE CONTROL LOOP

- Control loops are employed to keep process variables within specified range.
- Sensor picks up changes in parameter, sends signal to controller.
- Parameter compared with set point; controller acts accordingly.
- Temperature Control Loop has been installed to maintain stream going to storage.
- Steam flow rate into heat exchanger varied to acquire desired output temperature.

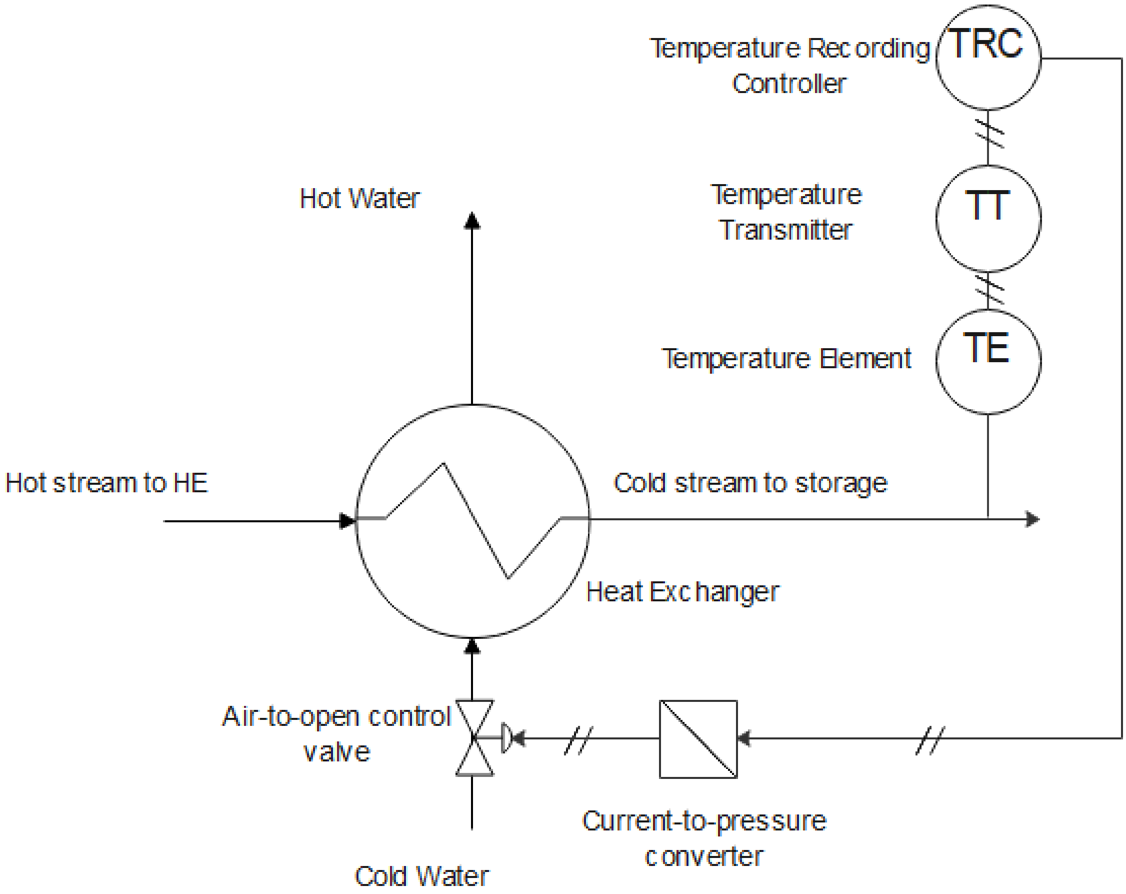


Figure 60: Temperature Control Loop

9.2 FLOW CONTROL LOOP

- A ratio control loop is employed to maintain desired ratio of steam flow rate to atmospheric residue (towards vacuum column) flow rate.
- Set point given to the controller.
- The flow rates are measured via flow sensors.
- Desired ratio achieved by varying flow rate of water.

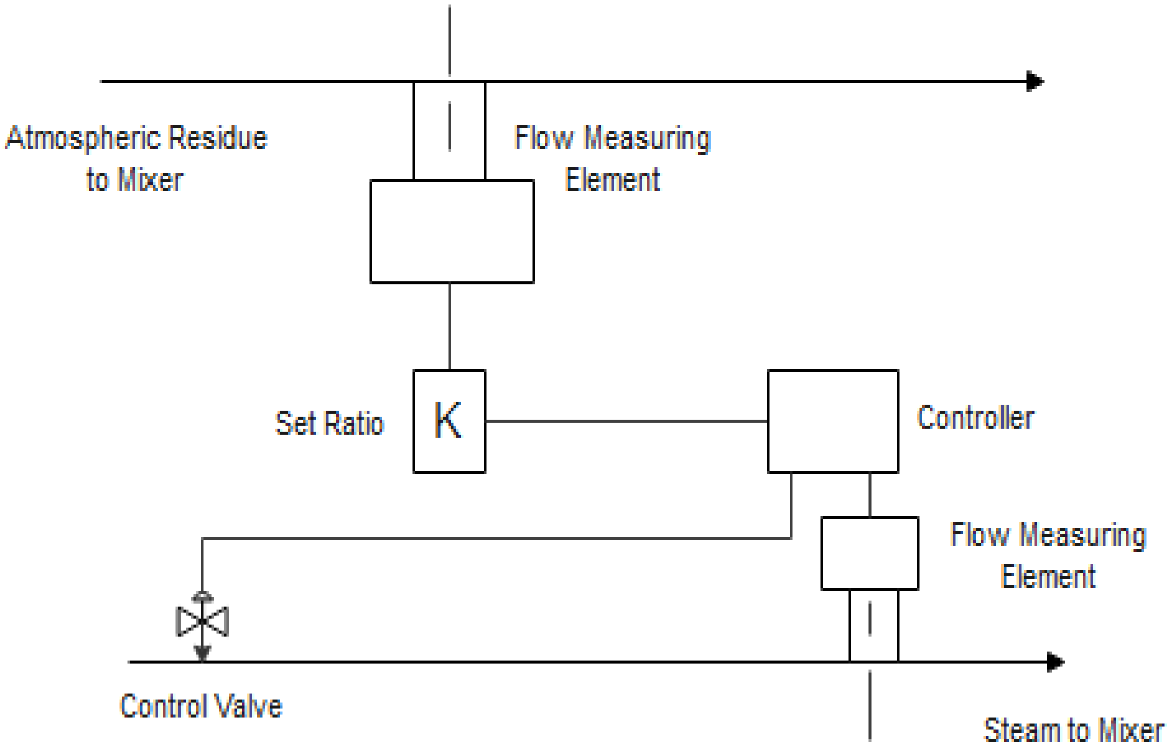


Figure 61: Flow Control Loop

9.3 PRESSURE CONTROL LOOP

- A pressure control loop is installed to monitor and control steam entering in steam ejector.
- Control achieved by detecting deviation in pressure and then varying steam ejector input accordingly.

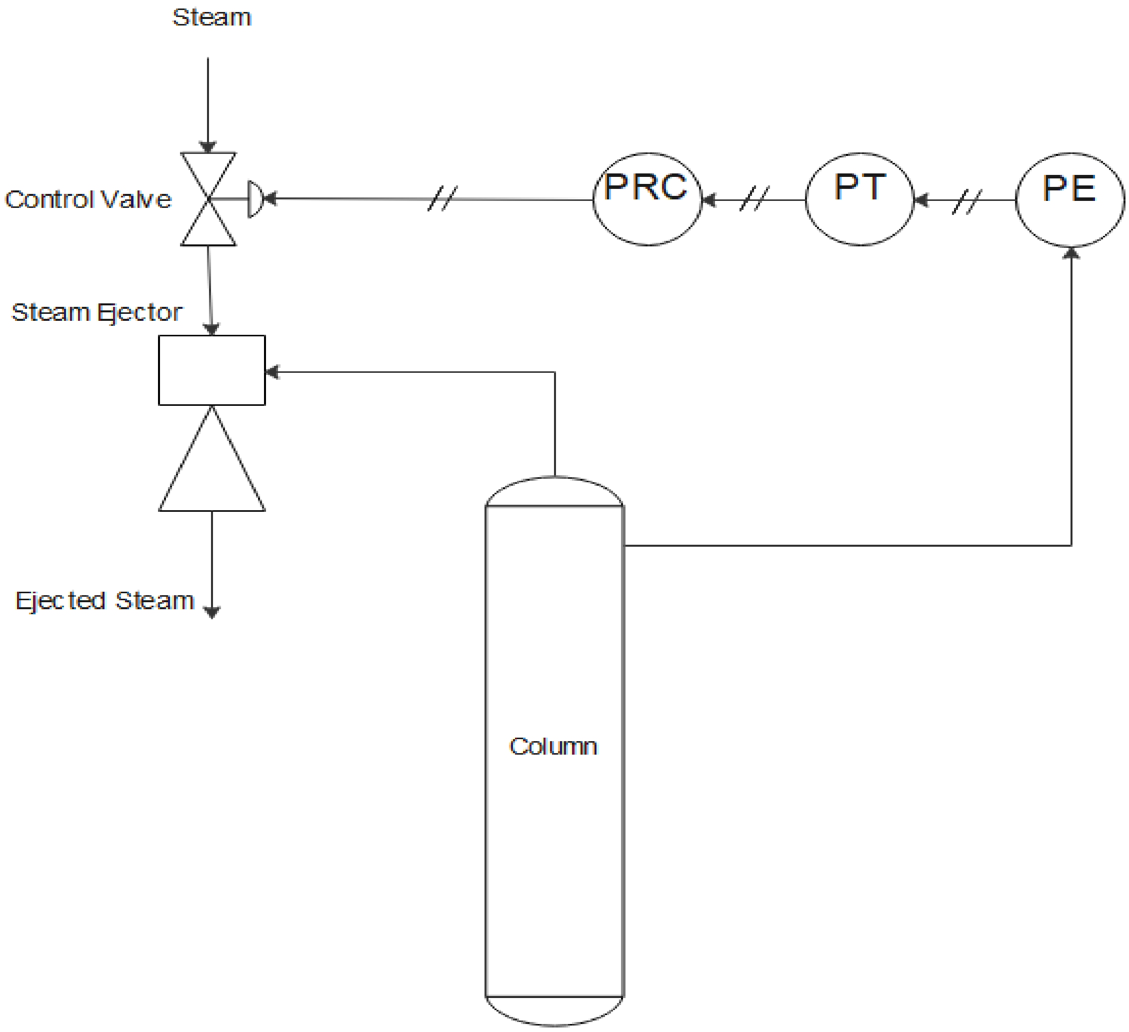


Figure 62: Pressure Control Loop

9.4 LEVEL CONTROL LOOP

- Level control loops are employed for both columns to maintain required liquid holdup.
- Level sensor picks up the deviations, the controller output then varies outlet valve opening. [14]

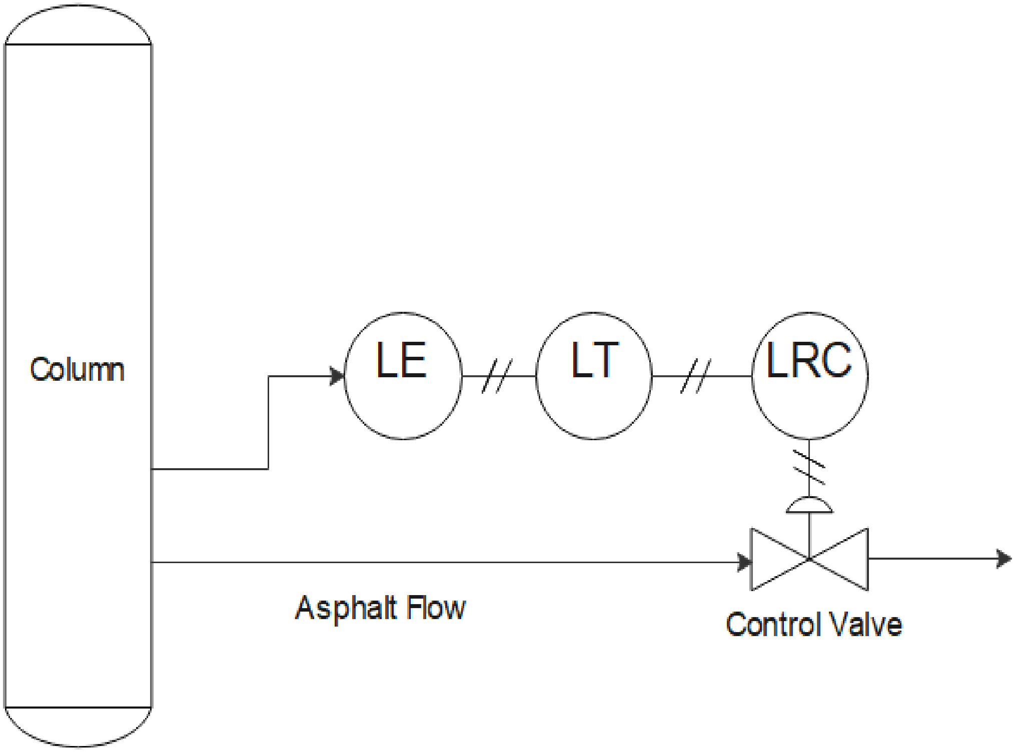


Figure 63: Level Control Loop

CHAPTER 10

HAZOP ANALYIS

A Hazard and Operability study of the proposed is fundamental to its technical feasibility from a safety point of view. Conventionally, such analyses usually involve using specific guidewords to assess the cause, effects, and preventative measures for the process. Examples of said guidewords may be "high temperature", "low pressure" or something else of the sorts. These guidewords vary with the equipment under study within a system. Each component within a system must be considered for a thorough risk assessment. The guidewords are essentially deviations that may potentially be encountered within the equipment and help to imagine several possible scenarios. HAZOP involves:

- Brainstorming what could possibly go wrong in the system. Listing downguidewords.
- A qualitative risk assessment.
- A "bottom-up" approach. Risk assessment relies on the observer's predictive abilities and experience.

KEY WORDS :

- Nodes: Pipe sections or vessels containing where process chemicals maybe present.
- Super nodes: Several nodes considered together as one node for quickerstudies.
- Hazard: Potential source of harm.
- Harm: Actual injury or damage to equipment, personnel, or environment.
- Risk: Probability of harm occurring and its associated level of damage.

METHODOLOGY:

- Identify all major nodes and super nodes.
- Identify why the node was chosen for the HAZOP study.
- Choose a key process parameter for said node.
- Apply all relevant guidewords for chosen parameter. These would result in unwanted deviations within the process. List down the deviations.
- Determine the potential causes of the deviations identified and add themto the table.
- For the deviations brainstormed, identify the outcomes they would leadto. List these down as the consequences.
- Once a node has been thoroughly studied, repeat steps 3 through 6 for the next node. Repeat this step till all chosen nodes have been accounted for. [15]

10.1 PUMPS

Parameters	Guide Word	Deviations	Causes	Consequences	Actions Required
FLOW	NO	No flow of water	Outlet valve closed/blocked, pipeline rupturing, total pump failure	Overheating of pumps, bursting/explosion of pipeline, release of water into the environment	Adjustment of valves on the delivery lines, proper monitoring of pump and connected pipeline.
	LESS	Low flow of water	Partial opening of the downstream valve, improper suction at inlet, pump cavitation	Overheating of pump leading to failure, possible pressure buildup in pipelines	Control system to vary valve opening according to flow requirements, priming of pump to avoid cavitation.
	MORE	More flow of water	Fail open valve at delivery	Flooding of the following vessel	Fail close valve at delivery with strict

		than require	line		control of pump so that it closes. Level indicators on following vessel to identify overflow.

Table 33: PUMPS HAZOP

10.2 PREHEATER

Parameters	Guide Word	Deviations	Causes	Consequences	Actions Required
FLOW	LESS	Oxygen take up	Malfunctioning of Damper Improper burning Fuel pressure too high Tube Rupture	Increased fuel loses Lesser heat transfer	Confirm damper opening in field. Check the draft of the furnace. Call and verify the damper through instrumentation. Check NG pressure on DCS.
	MORE	Oxygen take up	Malfunctioning of Damper Improper burning Fuel flow too low Furnace peep hole opening or any leakage from Furnace shell	Fuel Losses. Furnace temp may decrease	Confirm damper opening in field. May be too high. Adjust air register. Check burner air register opening.

Table 34: PREHEATER HAZOP

10.3 VACUUM DISTILLATION COLUMN

Parameters	Guide Word	Deviations	Causes	Consequences	Actions Required

PRESSURE	LESS	Pressure intop stage lower than require	Improper function of steam ejector. Blockage at top stage inlet	Excessive heating of contents that remain in column (waste of heat), Inefficient separation	Scheduled service and maintenanceof steam ejector. Checking for blockages at exit.
	MORE	Pressure in top stage is higher than required	Blockage at exit Steam ejector speed increased	Inefficient separation, Higher pumparound duty.	Scheduled service and maintenanceof steam ejector.
TEMPERATURE	LESS	Low Temperaturewithin bottom stage	Improper preheatingof feed	Inefficiencies in separation. Flooding within column	Check heat exchangers before bottom stage inlet.
	MORE	High Temperaturewithin bottom stage	Improper preheating of feed	Inefficient separation.	Check heat exchangers before bottom stage inlet.

Table 35: VDC HAZOP

10.4 HEAT EXCHANGER

Parameters	Guide Word	Deviations	Causes	Consequences	Actions Required

FLOW	LESS	Less flow ofoil into Heat exchanger	Pipe blockages	Higher temperature of exiting oil stream since steam flow rate remains unchanged	Temperature sensors at outlet and implementation of control for steam and oil stream flow rates.
	MORE	More than optimal oil flow to heat exchanger	Complete opening of inlet valve upon failure	Less than required temperature of exiting oil stream achieved since same amount of steam is being used to heat a bigger processstream	Temperature sensors at outlet and implementation of control for steam and oil flow rates. High pressure alarms to halt operation if this happens.

Table 36: HEAT EXCHANGER HAZOP

CONCLUSIONS

1. INCREASED PRODUCTION OF ASPHALT:

40% greater asphalt yield is observed after increasing the atmospheric residue splitter ratio (from 0.45 to 0.90) towards vacuum tower, along with a 10% decrease in FFO yield.

The below shown is the graph of comparison of original process and the modified process.

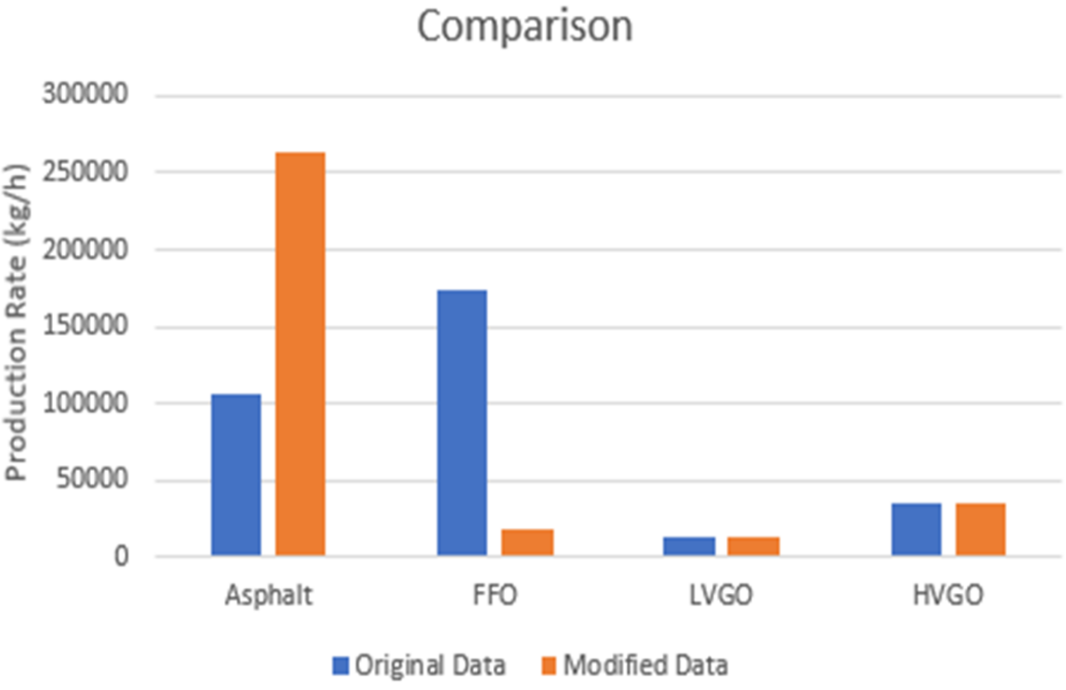


Figure 64: Comparison of product flows

2. EFFICIENT PAYBACK:

Payback after 3 years and approximately 9 months which is nearly 6 months earlier than the original process.

3. GREATER PROFIT:

Modifications offer profit of approximately 3.2 million $.

4. ECONOMICAL, SUSTAINABLE PROCESSING:

Better sustainability in VDC system is observed.

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