PROCESS DESIGN OF DISTILLATION OF CRUDE OIL AND CORROSION CONTROL IN OVERHEAD SYSTEM

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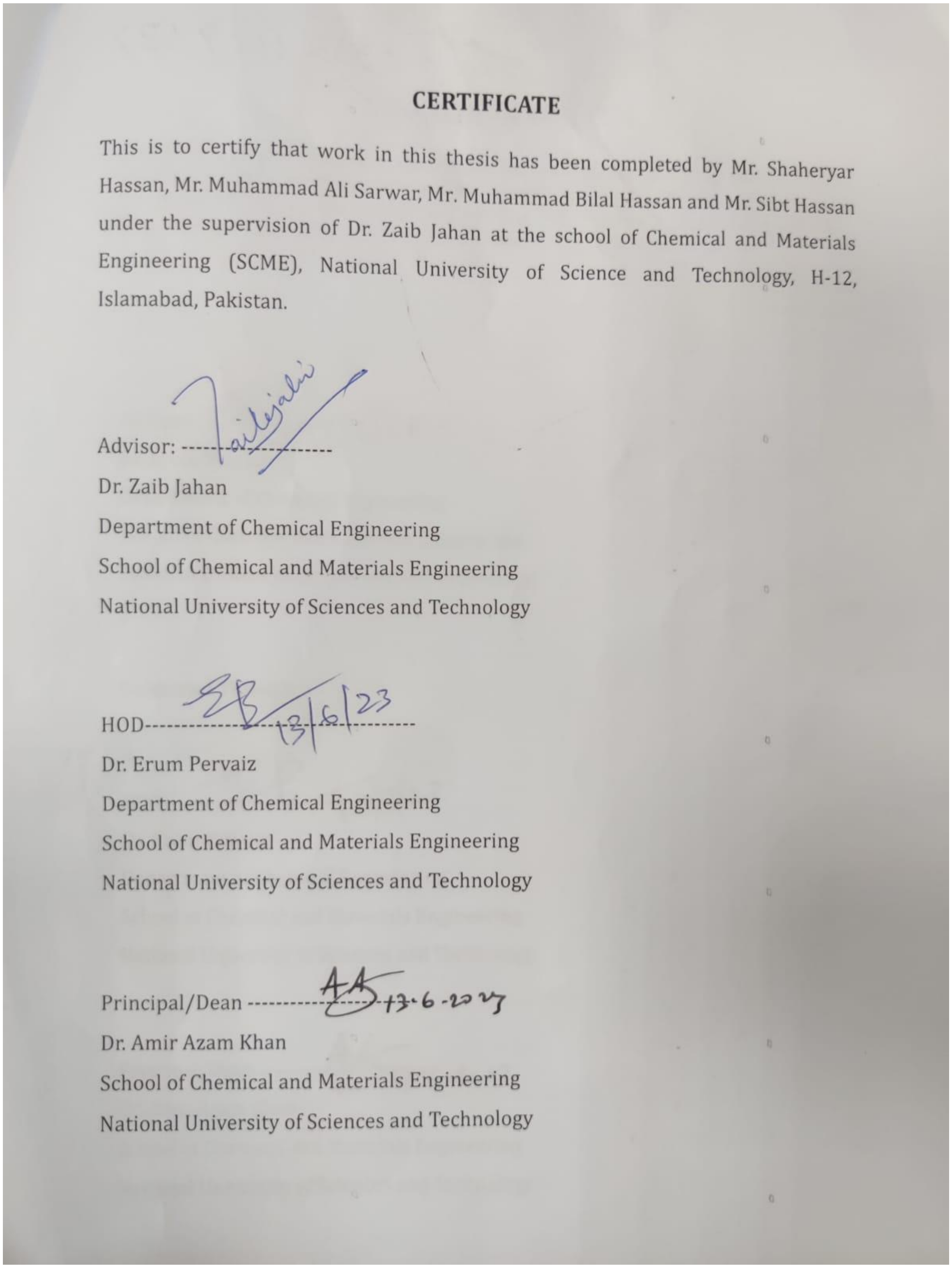

CERTIFICATE

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DEDICATION

This work is dedicated to our parents for always believing in us, our teachers, for broadening our perspective of the world and our tireless hard work.

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We would first like to express our heartfelt gratitude to Allah Almighty for being so merciful. We realize that without His guidance and blessings, it would not have been possible for us to successfully complete the project.

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ABSTRACT

This project assigned by Attock Oil Refinery (ARL) involves the process design of distillation of crude and corrosion removal in the distillation overhead system. Corrosion in the overhead system can cause production loss and unexpected shutdowns, mostly due to leftover salts from the desalter. So, to minimize the effect of corrosion, corrosion neutralizer (methylamine) and corrosion inhibitors are being used to neutralize the pH of the salt and protect the pipelines and 3-phase separator. The project aims to optimize the distillation process to improve product quality and ensure the efficient removal of corrosion in the overhead system to improve production reliability. Corrosion caused significant damage to pipes and equipment of the oil and gas industry. Unexpected shutdowns, loss of production and product quality are greatly affected. This review will cover the types of corrosion, how corrosive salt reacts with metal and its prevention and mitigation method in crude distillation overhead system. This review will also include a case study on crude distillation overhead system, factors responsible for corrosion in crude distillation overhead system and the economic cost of corrosion. The major Problem oil and gas industry facing is corrosion. Each has similar or different methods to reduce the effect of corrosion in distillation overhead system. The methods the company applies depend totally on the composition of crude oil and presence of salts. The report contains several methods in order to mitigate the effect of corrosion in distillation overhead system. Amine addition, corrosion inhibitor, material of selection and use of CORROSOMETER are discussed in detail.

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CHAPTER 1

INTRODUCTION

1.1 Overview

Over millions of years, numerous aquatic organisms such as plants and algae lived in shallow waters, died out, and sank to the bottom of the sea. Eventually, under the changing marine environment and the immense pressures and temperatures of the Earth, these organisms decomposed and turned into the fossil fuels we know today: coal, oil, and natural gas. Most of the oil reserves now lie where the ocean used to be. They are also on the seabed, where huge excavators are set up to bring them to the surface.

The oil itself comes in numerous colours, from black to dark brown, and even darker shades of red, yellow, brown, and even green, due to the different chemical compositions found around the world. Clear or bright colours are observed in compositions low in metals and sulphur.

The main product of petroleum is gasoline, which is consumed in large quantities in daily life. Recently, concerns have been raised that large-scale consumption of fossils will affect the carbon content of the atmosphere, leading to global warming.

Large amounts of oil are found below the surface, most of it in tar pits bubbling to the surface. Crude oil is also present at the bottom of the deepest wells, and development to extract this is underway. Remember, fossil fuels are non-renewable resources that have been developed over millions of years and are destined to eventually run out. Experts believe that humans have not yet reached "peak oil," the period of peak fossil production. This could be achieved as early as 2050. As a result, many industries today are focusing on more and more alternative energy sources.

1.2 Crude Formation

Crude oil forms under a variety of geological conditions that have been observed at the surface over millions of years. First, organisms such as algae, plants, and animals die and then drift toward the deep ocean. as they sink to the bottom of the sea. At the end of its life cycle, it is buried and crushed by the pressure of the sea, along with other layers of sediment and debris of dead organisms. Earth's constant plate activity has dried up the ancient oceans, leaving what is known as sedimentary basins. Beneath these pools, the Earth's mantle compressed organic matter under extreme pressure and temperature and in the absence of oxygen, causing the organic matter to decompose into a waxy substance called kerogen. Further exposure to heat and pressure causes further degradation, known as catagenesis, which converts kerogens into more hydrocarbons. Hydrocarbons themselves consist only of hydrogen and carbon atoms, and their combination, composition and chain length depend on the conditions under which they were formed.

1.3 Classification of Crude Oil

In general, crude oil is primarily composed of hydrocarbons, with approximately 12-13% hydrogen and 85-86% Carbon by weight. Oxygen, nitrogen, sulfur, and small amounts of some metals such as iron, copper, and nickel make up the rest of the composition. A major factor in the molecular arrangement of hydrocarbons depends on the composition of the original sources such as algae and plankton. A secondary effect is the action of heat and pressure on plants and algae. Therefore, when crude oil is extracted from the ground, it is made up of hundreds of different petroleum compounds. Light oils are almost 97% hydrocarbons, whereas heavy oils and asphalts may contain only 50% hydrocarbons. Therefore, it is always desirable to upgrade petroleum feedstocks to produce useful consumer products.

1.3.1 Saturated Chains

Carbon compounds range from the simplest molecule, methane (CH4), which consists of one carbon atom and four hydrogen atoms. Then comes ethane (C2H6) and so on. This extension relation can be used to define a whole class of molecules called paraffins, which are characterized by their chain-like existence.

Formula for CnH2n+2. This class of molecules is called saturated chains. This is because each valence electron on carbon can be accepted by four different hydrogen or carbon molecules, forming only single bonds, and leaving no valence electrons, giving paraffins a chemically very stable structure.

Paraffins exist is multiple structures, they could either be straight chained such as butane in the figure or iso-paraffins (branched chains). In general, crude oil products are naturally occurring normal paraffins that require refiner intervention and processing to convert to the more desirable iso-paraffins. A major concern with natural paraffins stems from their poor engine fuel performance, in contrast to iso-paraffins, which have good engine combustion characteristics. Paraffin, which has a long carbon chain, has waxy properties.

The structure adopted by hydrogen (H) and carbon (C) molecules in four common hydrocarbon compounds. Beyond 4 carbons, the carbon chain can form closed ring structures that are classified as cyclic compounds. Cyclic compounds that are naturally saturated are called naphthenes. In general, naphthenes have poor lubricating properties, but are easier to convert into high quality gasoline than paraffins.

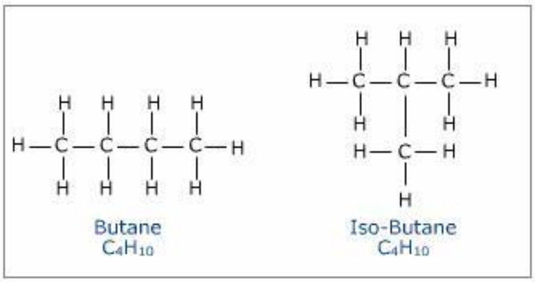


Figure 1.1: Structure of Saturated Hydrocarbons

1.3.2 Unsaturated Chains

Another important class in petroleum refining is compounds composed of unsaturated molecules. In this case, the valence electrons of carbon are not individually bonded to hydrogen or other carbon atoms, but two or three electrons are shared with adjacent carbon atoms to form double or triple bonds. Like saturated chains, unsaturated chains can be branched or cyclic and are called olefins. Crude oil is made up of only a fraction of olefins, most of which are produced in refineries after further processing. Relatively speaking, Olefins are more reactive compounds and can be easily converted into longer chains. An extension of the family of unsaturated compounds, called aromatics, are ring molecules. The most basic aromatic molecule is benzene (C6H6). Because each benzene atom is linked to adjacent carbon atoms by double bonds, benzene is a chemically very active and unstable compound, making it an important building block for the petroleum industry.

Although these unsaturated carbons have excellent combustion properties, they are often unstable, making them difficult to store and transport, and raising environmental concerns.

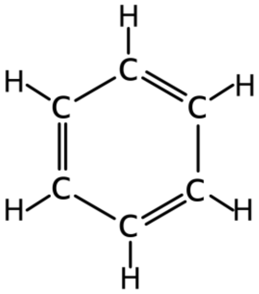


Figure 1. 2: Structure of Unsaturated Hydrocarbons (Benzene)

1.4 Density of Crude Oil

Crude oil density is determined using API gravity. A term coined by the American Petroleum Institute (API), the industry association for the oil and gas industry. This system has standardized a variety of products for the oil and gas market, including pumps, gauges, and drills. There are also numerous units of measurement, the most concerned to us being API Gravity. API Gravity is a reference measurement of the specific gravity of crude oil and water. A value above 10 means that the oil floats and is "light", and a value below 10 means that the oil sinks and is "heavy".

1.5 Crude Oil Reservoirs:

Oil is found in so-called reservoirs near the surface. Deep under the surface, where temperatures and pressures are higher, crude oil is generally found. Due to the high-pressure underground, the oil naturally rises due to the pressure gradient, close to the surface and is blocked by an impermeable layer of rock hundreds of meters below the surface. This crude oil can be trapped in rock structures faulted or cracked by the movement of Earth's plates and masses. Crude oils can also be stratigraphically bound. This means that different rock formations have different permeability and porosity for the movement of crude oil. For example, crude oil can easily pass through sandstone layers, but can become trapped under shale layers. Scientists and experts typically look for typical terrains where crude oil may be trapped beneath the surface. Seismic reflectometry is a widely used technique for locating subsurface crude oil within suspect rock formations. Sound waves are directed toward Earth and reflected by various types of rocks. The returning waves are observed and used to analyze the geological conditions underground to determine if there are oil reserves there.

1.6 Products of Crude Oil

Some of the major products of Crude Oil include Petroleum Gas, Gasoline, Kerosene, Diesel, Lube Oil and Asphalt. Some of these products are discussed here.

1.6.1 Gases

Refineries produce gases like methane, ethane, propane, butane along with hydrogen. Hydrogen produced is generally consumed within the refinery in desulfurization facilities, removing H2S gas from the feed stream and then being parted into elemental Sulphur and Hydrogen. In small quantities, Hydrogen can also be used as part of refinery fuel system. Gases produced in refineries have variable compositions, usually dominant in methane in contrast to other counterpart gases. Methane even though in abundant amount, and similar heating value of that of natural gas, its periodic varies of heating value as plant product makes it unsuitable as consumer gas and is therefore often used within the plant operation. Ethane on the other hand is retrieved from the refinery system and provided to petrochemical system as feedstock. The remaining components, propane and butane as liquified as LPG (Liquid Petroleum Gas) and distributed to consumers for domestic heating purposes and smaller industries.

1.6.2 Gasoline

Gasoline is standardized to meet three major quality requirements: uniform burning pattern, ability to ignite in cold conditions and minimal additives to address environmental concerns.

1.6.3 Diesel

Diesel often powers energy-intensive vehicles such as trains, buses, trucks, and heavy industrial machinery. Diesel engines rely on the compressibility of fuel and air. For gasoline engines, this can be fatal, as the explosion causes knocking and engine damage. However, this process is necessary for diesel engines. In an optimal diesel engine, multiple ignition points are created simultaneously within the compression cylinder, and the fuel flowing into the cylinder is expected to ignite as well. Therefore, unlike gasoline engines, diesel engines tend to favor straight chain hydrocarbons and have their own cetane number compared to blending with hexadecane (cetane). The ideal cetane number is said to be around 50, giving similar burning characteristics to a 50:50 mixture of standard fuels. Slower energy suppliers, such as marine engines and stationary power plants, also tolerate heavier gas oils. Diesel standards were not stringent until the late 1990s. The cetane number is still important, but the sulfur content of the fuel was not an issue. Sulfur levels up to 5,000 ppm were common across the market. However, regulations have now been introduced to limit the sulfur concentration to 10-15ppm.

1.6.4 Kerosene

Although the use of kerosene as a light source has declined, it is still widely used for cooking and heating, and is the primary fuel for aircraft engines. Kerosene is standardized to have a minimum flash point of 49°C when handled and stored in the same manner. For domestic use, kerosene produces a smokeless, odorless fuel. The most important quality standard for gasoline is for jet engines. Temperatures at the high altitudes where jets fly can reach -50°C. Therefore, the propellant must be free of water droplets and remain liquid to prevent entrapment of wax particles. Military jet fuel is even subject to stringent specifications to achieve its specific purpose.

1.6.5 Fuel Oil

Fuel oil, also known as furnace oil, is the residual product of refineries. It is viscous in nature, high in sulfur content and rich in heavy metals. Therefore, its use is limited, and it is often necessary to mix it with low-viscosity light oil for ease of handling. Furnace oil is therefore the only refined product with a lower selling price compared to crude oil. Burning blast furnace oil produces dangerous sulfur trioxide, which requires the industry to install and regulate equipment such as coke ovens and desulfurization plants. Heavy metals also affect their use as they produce ash when burned and lead to burner fouling. Removal of these metals is costly and leads to lower fuel prices on the market.

1.6.6 Lube Oil

Originally, the lubricating properties of petroleum fractions were dependent on the well from which they were extracted. Pennsylvania crude oil showed superior properties due to its high paraffin content. However, the introduction of techniques such as hydrocracking and solvent extraction later expanded the range of feedstocks. The main property of lubricants is viscosity. Requirements range from high-speed spindles for textiles to more viscous materials such as wire ropes and open gears. A wide range of products offer numerous properties of oils, all serving specific purposes, but the automotive industry remains the most important market segment. Viscosity ratings are given by the Society of

Automotive Engineers and range from 5 to 50. Although this rating is commonly used in the United States, the British Petroleum Institute offers virtually the same rating as his SAE.

Ordinary mineral oil provides sufficient lubricity at low temperatures. However, when the oil is diluted, it loses its lubricating properties over a wide temperature range and becomes unusable. To address these concerns, longer chain multigrade polymers were developed that serve the purpose over a wider temperature range. For example, an SAE 10W40 rating means that the viscosity of the lubricant is 10W SAE at -18°C and increases to 40W SAE at 99°C. Therefore, proper lubrication is ensured both in winter and in summer. Antioxidants and detergents are other additives incorporated into lubricants to improve performance and maintain engine cleanliness.

CHAPTER 2

LITERATURE REVIEW

2.1 Abstract

Corrosion caused significant damage to piping and equipment of oil and gas industry. Unexpected shutdowns, loss of production and product quality is greatly effected. This review will cover the types corrosion, how corrosive salt react with metal and its prevention and mitigation method in crude distillation overhead system. This review will also include a case study on crude distillation overhead system, factors responsible for corrosion in crude distillation overhead system and the economic cost of corrosion. The major Problem oil and gas industry facing is corrosion. Each have similar or different methods in order to reduce the effect of corrosion in distillation overhead system. Methods company applies depends totally on the composition of crude oil and presence of salts. Report contain number of methods in order to mitigate the effect of corrosion in distillation overhead system. Amine addition, corrosion inhibitor, material of selection and use of CORROSOMETER are discussed in detail.

2.2 Introduction

Distillation is the application and removal of heat to separate hydrocarbons by their relative volatility or boiling points. This necessary addition of heat normally in the feed stream or at the tower bottoms via a reboiler can also lead to number of unwanted consequences including corrosion. The removal of heat at distillation overhead system is the main area where corrosion is damaging equipment and pipelines along with reduced separation efficiency of the column. Overheads, in general, refer to the pipe work, exchangers, and condensers coming off the top of any distillation, fractionation or separation vessel. This reduced separation efficiency increases the need for column maintenance and unit down time. The Presence of salts are the main factor that is causing all these problems. Composition of Salts in crude oil depend from region to region. There is different kinds of salts present in crude oil such Nacl, Mgcl, Cacl2. **NaCl will not hydrolyse**

and CaCl2 has a very low tendency. At around 140-150°C, MgCl2 can hydrolyse and these chlorides will pass the distillation column to end-up in the overhead system. There these chlorides will attack the metal surface, **resulting in corrosion**. Desalination of crude oil through electric desalter is one of the first step to remove these salts. This is an important step to avoid corrosion in the distillation overhead system. Most of the salts are washed out with the desalter effluent at the desalter. Desalter can't work on 100% efficiency, a small portion of NaCl, CaCl2 and MgCl2 will leave the desalter with the crude oil. So these remaining salts are the main culprit which is why overhead system of distillation is greatly effected. There are two categories of salts found in crude oil: one type is naturally present as inherent components of crude oil, while the other type is intentionally introduced as foreign matter during crude oil extraction, refining and transportation. During the crude oil refining process, such as distillation, these organic chlorides tend to convert into hydrogen chloride (HCl). Subsequently, a small amount of water dissolves this HCl, resulting in the formation of a dilute hydrochloric acid solution. This acidic solution then causes significant corrosion in the top section of the distillation column and the overhead condensing system.

2.3 Different corrosion mechanism in distillation overhead

Corrosion Mechanisms: In the overhead system of the distillation unit, several significant corrosion mechanisms can occur. These mechanisms can be summarized as follows:

1. **HCl Corrosion**: Corrosion arises from the dilution of HCl vapor in liquid water. The initial water droplets formed during condensation can be highly acidic and lead to localized but severe corrosion at very low pH levels. HCl is predominantly generated through the hydrolysis of inorganic chloride salts present naturally in crude oil. Heat-unstable salts like MgCl2 and CaCl2, which hydrolyze at lower temperatures compared to NaCl, contribute significantly to HCl formation. The addition of NaOH to crude oil helps minimize salt hydrolysis.

Reaction:

$$MgCl_2 + H_2O \rightarrow 2HCl + MgO$$

$$CaCl_2 + H_2O \rightarrow 2HCl + CaO$$

2. **Ammonium Chloride (NH_4Cl) Corrosion**: HCl tends to combine with NH_3, forming ammonium chloride (NH_4Cl). NH_3 can originate from hydrocarbon streams blended into crude oil, desalter wash water from other refinery units, or intentional injection into the CDU overhead vapors for pH control. The re-sublimation of solid NH_4Cl salt depends on the partial pressure of both HCl and NH_3 vapors. NH_4Cl salts usually form above the aqueous dew point and remain dry and non-corrosive as long as the process temperature is sufficiently higher than the dew point. Wet NH_4Cl deposits can be highly corrosive and cause fouling, particularly in overhead heat exchangers and air coolers. Preventive measures include reducing NH_3 and HCl levels, water washing to remove salt deposits, minimizing intentionally injected NH_3, and increasing the process temperature in the overhead line.

3. **Amine Hydrochloride Salt Corrosion**: HCl can react with amines to form hydrochloride salts that exhibit similar corrosive behavior to NH_4Cl. Amines can originate from amine solutions injected for pH control, desalter wash water carry-over, upstream facilities where amines are used as H_2S scavengers, or imported residues/synthetic crude feedstocks. These externally introduced amines are often referred to as "tramp amines."

4. **Wet H_2S Damage**: Under wet conditions and sufficient H_2S presence, blistering and cracking can occur, depending on water pH, cyanide content, plate metallurgy, and welding procedures. This study focuses on corrosion due to HCl and salts and does not delve into different cracking mechanisms.

Minimization of Overhead Corrosion: To mitigate the above-mentioned corrosion mechanisms, several common approaches are employed:

1) **Blending:** Corrosive crudes are often blended with less-corrosive ones, based on refinery configuration and crude flexibility.

2) **Crude Oil De-watering:** Minimizing water content in crude oil helps reduce salt dilution. Crude oil storage tanks are drained to achieve this, aiming for a maximum water content of 0.5 wt.% in desalted crude oil.

3) **Crude Oil Desalting:** Mixing crude oil with wash water at elevated temperatures and specific water contents removes inorganic contaminants and solids. Chemical additives, such as demulsifies or pH controllers, are used to enhance the desalting process. The desalter separates the oil and water phases, with most salts and sediments leaving the vessel with the water phase.

4) **Caustic Injection:** Diluted NaOH is added downstream of the desalter to minimize hydrolysis and reduce HCl in the overhead. NaOH converts unstable chloride salts and already formed HCl into NaCl, which is discharged through the bottom of the distillation column.

5) **Injection of Amine**: amines are commonly used to neutralize HCl and control pH in the condensing water phase.

6) **Use of corrosion inhibitor**: Injection of corrosion inhibitors: Corrosion inhibitors, often referred to as "filmers," are injected to reduce corrosion. These inhibitors form a protective film on the internal surfaces, providing some degree of protection against acidic solutions. Organic corrosion inhibitors are commonly used for this purpose.

Corrosometer:

It is another way to reduce the corrosion in distillation overhead. Corrosometer provide regular data about corrosion and from that we can get amount of acidic salt going to overhead. So from that data we can change the efficiency of desalter (desalter electric field).

2.4 Procedure:

Following are our proposed method in order to remove corrosion in overhead system:

1) **Methyl amine addition:**

Methyl amine addition reduces the acidic effect of hcl and convert it into neutral base salt. The reaction is as follows.

$$\mathbf{MgCl2+2H2O{-}{-}{>}2HCl + Mg(OH)2}$$

$$\mathbf{HCl+CH3NH2{-}{-}{>}CH3NH3Cl}$$

Now addition of wash water will remove the precipitate of salts along with it.

2) **Addition of corrosion inhibitor:**

We are using MBT(2-mercaptobenzothiazole) corrosion inhibitor that form a protective film on the internal surfaces, providing some degree of protection against acidic solutions. Organic corrosion inhibitors are commonly used for this purpose.

3) **Caustic addition:**

$$\mathbf{2NaOH + 2HCl \rightarrow 2NaCl + 2H2O}$$

In this reaction, sodium hydroxide (NaOH) reacts with hydrochloric acid (HCl) to form sodium chloride (NaCl) and water (H2O). By consuming the

HCl, the concentration of the acidic species is reduced, which helps mitigate corrosion in the system.

The addition of caustic raises the pH of the system, making it less acidic. This shift towards a higher pH environment further aid in reducing the corrosive effects of acidic compounds Injection of Chemical Agents: NH3 or amines are commonly used to neutralize HCl and control pH in the condensing water phase by

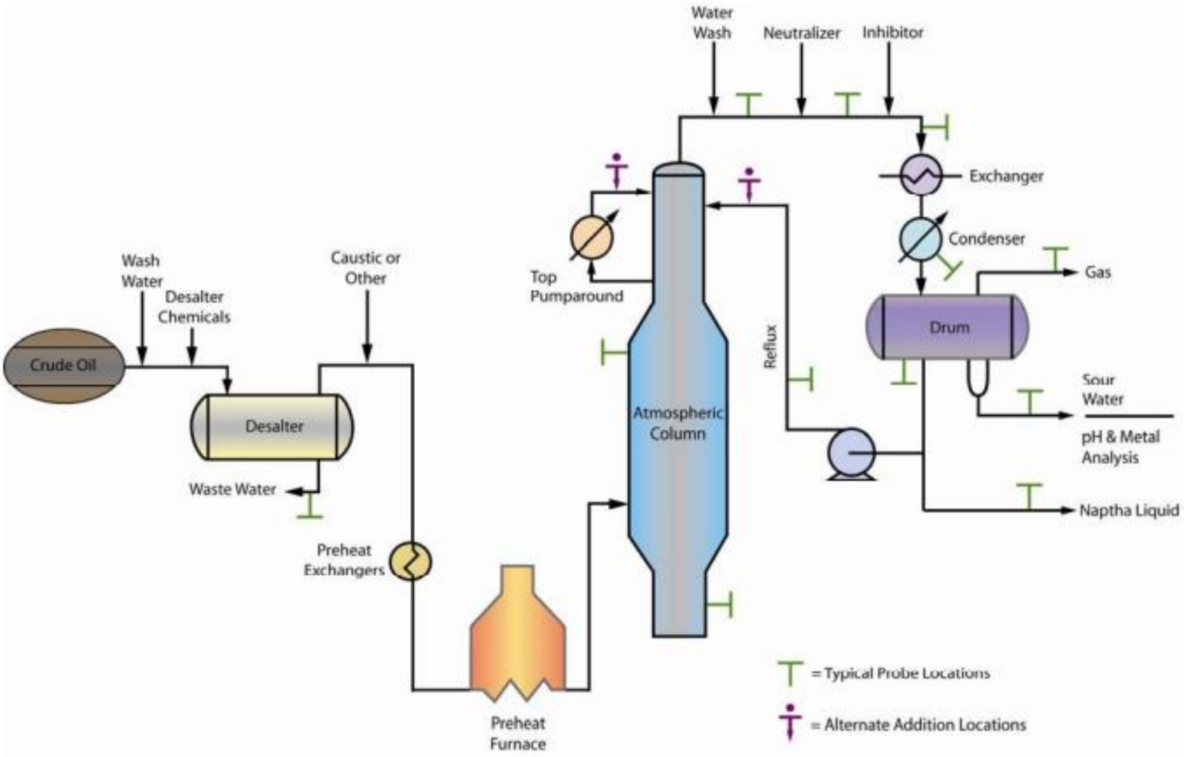


Figure 2.1: addition of various chemicals

CHAPTER 3

PROCESS DESCRIPTION

3.1 Description

Crude Oil contains a variety of components including impurities. These impurities are removed during pre-treatment process. There are different pre-treatment steps, but the most important pre-treatment equipment is Desalter. As the name suggests, the primary purpose of a desalter is to remove salts from the crude oil. Since Desalter is not 100% efficient, some of the salts are carried along the crude oil. These salts are primarily responsible for the Corrosion at the overhead accumulator of Pre-Flash and Atmospheric Towers. Methods to eliminate corrosion at the overhead accumulator have already been discussed in the previous chapter.

The desalted crude oil is then sent to the pre-flash vessel to be separated into liquid and vapor feed streams. The pre-Fash column fractionates light hydrocarbons from crude oil. Light naphtha is overhead stream while heavy naphtha is withdrawn from the bottom. Light naphtha is further treated to yield LPG. Heavy crude is fed into the atmospheric tower. Atmospheric Tower is a multicomponent distillation column that separates crude oil into different fractions. Five different streams including Heavy naphtha, Light Diesel Oil, Heavy Diesel Oil, Kerosene and Residue are separated by the atmospheric tower while off gases escape from the top.

3.2 Desalter

Most desalters used industrially are horizontally oriented and electrostatic in operation. A typical electrostatic desalter has two electrodes with a large potential difference between them. 2 kinds of electric fields are generated as a result: a primary field of around 600 Volts per centimetre between the lower electrode and the water-crude oil interface, and a stronger secondary field (1000 V/cm). While the secondary field is responsible for ionization and the resulting coalescence of the water droplets within the crude oil, the primary field aids in the faster settling of these droplet. Since salts preferentially dissolve in water, settling out and removal of water from crude oil helps achieve more than 90% salt removal. The settled water exits with the salts and moisture previously present within the crude itself.

3.3 Pre-Flash Tower

A pre-flash vessel is employed right after the desalter to split the feed into gaseous and liquid streams. The primary purpose of Pre-Flash tower is to separate light naphtha. This Light Naphtha is then sent to Stabilizer.

This separation is only valid if:

1. There is a large difference between the volatility of the components making up the liquid and vapor streams.

2. The temperature and pressure of the flash are adequate to achieve desired separation.

Desired temperature is achieved using series of heat exchangers. The temperature of the flash separation was 150°C and the pressure was reduced to 1 bar via pressure drop of 10 kPa. Another purpose of pre-flash is to reduce the required heat duty by eliminating the need to heat up the entirety of the feed. Instead, a low pressure and moderate temperature increase allows us to separate the liquid and vapor streams easily and send them to the column separately.

3.4 Atmospheric Tower

The heavy crude from the pre-flash bottom is fed into the atmospheric tower. The stream first enters a direct fired charge heater, where it is heated to 357°C. From the heater, crude enters flash zone of the crude tower. Separation of different components take place here. Heavy Diesel Oil is withdrawn as a side product from tray number 10 and sent to the Heavy Diesel Oil Stripper. Light Diesel Oil and Kerosene are with drawn from tray number 16 and 28 and sent to

LDO and Kerosene Strippers respectively. Residue is the bottom product of Atmospheric tower.

In order to increase overall process efficiency, our products that are being withdrawn from atmospheric tower are being used as hot fluids to heat the crude oil in heat exchangers. As a result, our products cool down while Temperature of crude feed rises accounting for heat efficiency.

3.5 Process Flow Diagram

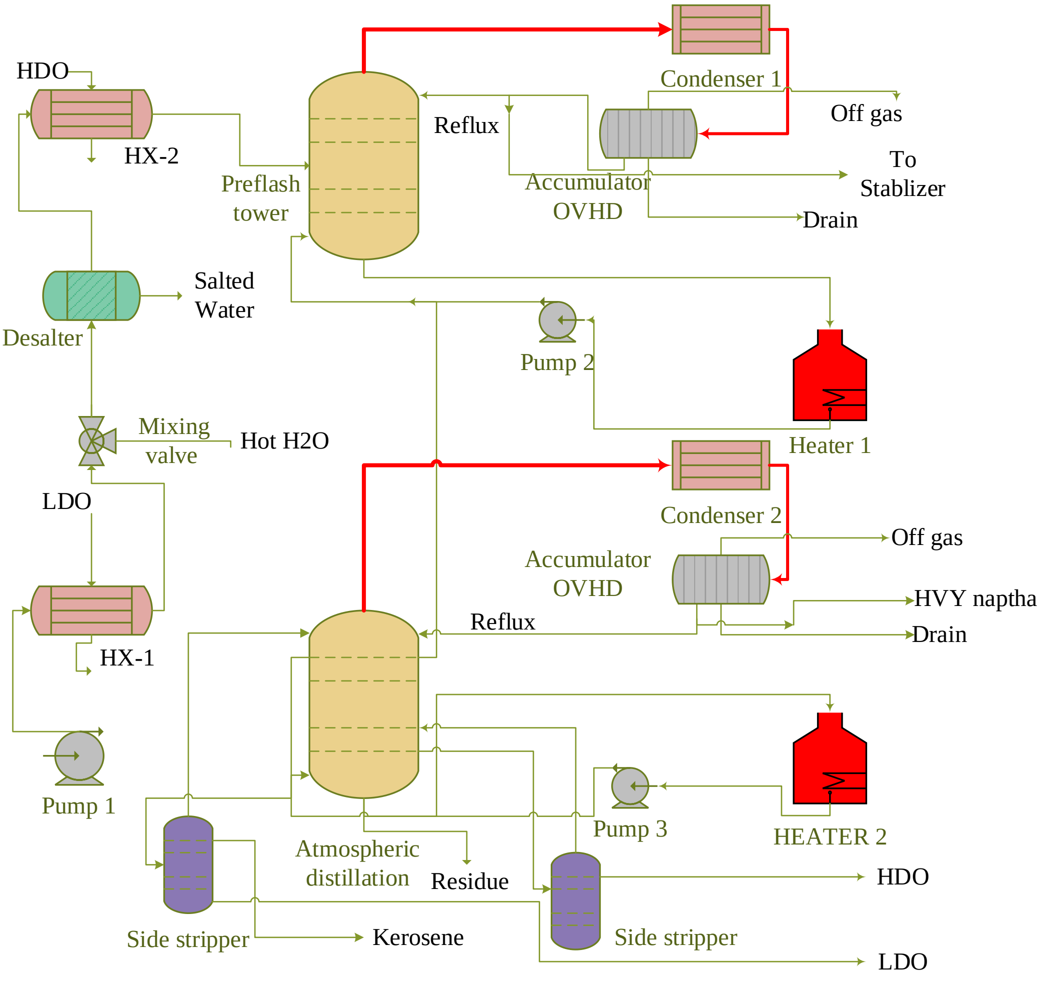


Figure 3.0-1: Process Flow Diagram for Distillation of Crude Oil

CHAPTER 4

MATERIAL BALANCE

4.1 Introduction

The Material Balance is done for a feed rate of 32,000 barrels per day. The law of conservation of mass was applied to all the components of the system.

Input - Output + Generation – Consumption = Accumulation

It is important to note that there is no chemical reaction involved in the distillation of crude oil. All the equipment involved Unit Operations. So, Generation and Consumption terms were eliminated. Since it is a steady state process, mass flow rate entering in the system is equal to the mass flow entering leaving the system. Crude consists of a large number of components, so we used Aspen HYSYS to generate petroleum assay. The Feed shown here is obtained from Petroleum Assay. The total mass flow rate is 171300 kg/hr.

Table 4.1: Feed

Components	Mass Flow rate (kg/hr)
Methane	0.00
Ethane	142.68
Propane	784.39
i-Butane	1146.05
n-Butane	1213.47
H2O	0.00
NBP [1]45*	544.25
NBP [1]71*	677.69
NBP [1]96*	994.29
NBP [1]118*	1096.82
NBP [1]143*	1338.82
NBP [1]169*	2009.38

NBP [1]195*	4391.44
NBP [1]215*	5153.44
NBP [1]243*	3606.69
NBP [1]267*	5448.93
NBP [1]292*	5627.23
NBP [1]317*	5685.96
NBP [1]341*	5634.45
NBP [1]366*	5563.95
NBP [1]391*	5519.62
NBP [1]416*	5489.27
NBP [1]441*	5416.22
NBP [1]465*	5223.25
NBP [1]490*	4931.29
NBP [1]515*	4656.14
NBP [1]540*	4535.91
NBP [1]565*	4642.99
NBP [1]590*	5074.88
NBP [1]614*	5583.53
NBP [1]638*	4688.40
NBP [1]664*	3659.77
NBP [1]689*	3362.97
NBP [1]714*	3489.06
NBP [1]739*	3795.67
NBP [1]763*	3981.51
NBP [1]788*	3933.13
NBP [1]825*	7367.05
NBP [1]872*	6025.83
NBP [1]926*	5604.99
NBP [1]974*	4440.33
NBP [1]1025*	3726.84
NBP [1]1074*	3156.50
NBP [1]1124*	2425.25

NBP [1]1174*	2197.08
NBP [1]1247*	3438.17
NBP [1]1372*	1926.23
NBP [1]1490*	1794.47

For material balance of each equipment, we summed up these hypothetical components in terms of Carbon numbers. This was done using the boiling point data available for each Hydrocarbon. From here on, the results of material balance for each component are shown in tabular form.

4.1.1 Mixing Valve

Mass flow rate entering in Mixing Valve = Mass flow rate leaving the Mixing Valve

Table 4.0-2: Mixing Valve Material Balance

Units (kg/hr)	**IN**		**OUT**
Component	**Crude IN**	**Water IN**	**Mixer OUT**
Methane	0.00	0.00	0.00
Ethane	142.49	0.00	142.49
Propane	783.38	0.00	783.38
Butane	4573.09	0.00	4573.09
C5-C11	61983.21	0.00	61983.21
C12-C17	34648.14	0.00	34648.14
C18-C22	22977.37	0.00	22977.37
C23-C28	17325.99	0.00	17325.99
C29+	28710.33	0.00	28710.33
Water	0.00	12700.59	12700.59
Salts	1000.00	0.00	500.00
Total	171369.00	12700.59	184069.59
Total	**184069.59**		**184069.59**

4.1.2 Desalter

Mass flow rate entering in Desalter = Mass flow rate leaving the Desalter

Table 4.0-3:Desalter Material Balance

Units (kg/hr)	**IN**	**OUT**	
Component	**Desalter IN**	**Desalted Crude Oil**	**Salt & Water**
Ethane	0	0	0
Methane	142.49	142.49	0
Propane	783.38	783.38	0
Butane	4572.75	4572.75	0
C5-C11	61983.21	61983.21	0
C12-C17	34648.14	34648.14	0
C18-C22	22977.37	22977.37	0
C23-C28	17325.99	17325.99	0
C29+	28710.33	28710.33	0
Water	12700.59	380.01	12320.58
Salts	1000	50	950
Total	184844.25	171573.67	13270.58
Total	**184844.25**	**184844.25**	

4.1.3 Pre-Flash Tower

Mass flow rate entering in Pre-Flash = Mass flow rate leaving the Pre-Flash

Table 4.4: Pre-Flash Material Balance

Units (kg/hr)	**IN**	**OUT**	
Component	**Pre-Flash IN**	**Light Naphtha**	**Crude Oil**
Methane	0	0.000	0.000
Ethane	142.49	136.53	5.96

Propane	783.38	730.69	52.69
Butane	4572.75	4151.31	583.74
C5-C11	61983.21	26995.05	34988.16
C12-C17	34648.14	2120.07	32528.08
C18-C22	22977.37	0.00	22825.42
C23-C28	17325.99	0.00	17316.04
C29+	28710.33	0.00	28709.81
Water	379.66	366.50	13.15
Total	171523.32	34500.15	137023.05
Total	**171523.32**	**171523.20**	

4.1.4 Atmospheric Crude Tower

Mass flow rate entering in Column = Mass flow rate leaving the Column

Table 4.5.1: Atmospheric Tower Material Balance (IN)

Components	**Heavy Naphtha**	**Steam**	**Steam 1**	**Light Steam**	**Steam 3**
C1	0	0	0	0	0
C2	5.96	0	0	0	0
C3	52.68	0	0	0	0
C4	583.73	0	0	0	0
C5-C11	34988.1	0	0	0	0
C12-C17	32528.07	0	0	0	0
C18-C22	22825.42	0	0	0	0
C23-C28	17316.03	0	0	0	0
C29+	28709.81	0	0	0	0
Water	13.15	771.1	635	771.1	771.1
Total	137022.95	2948.9	635	771.1	771.1

Table 4.5.2: Atmospheric Tower Material Balance (OUT)

Component	WaterOut	Residue	Heavy Diesel	Light Diesel	Kerosene	Naphtha	OFF Gas
C1	0	0	0	0	0	0	0
C2	0	0	0	0	0	2.17	3.76
C3	0	0	0	0	0.0008	34.18	18.51
C4	0	0	0	0	0.7	517.85	65.56
C5-C11	0	23.6	660.72	14653.42	18638.46	1032.81	23.76
C12-C17	0	749.34	3552.73	27860.36	372.48	0	0
C18-C22	0	739.47	2792.04	19291.83	2.11	0	0
C23-C28	0	575.02	2137.23	14603.78	0	0	0
C29+	0	954.9	3546.23	24208.68	0	0	0
Water	2823.68	2.83	11.63	79.24	37.2	0.53	1.76
Total	2823.68	3045.16	12700.58	100697.31	19050.9508	1587.54	113.35

4.1.5 Light Diesel Stripper

Mass flow rate entering in LD Stripper = Mass flow rate leaving the LD Stripper

Table 4.6: Light Diesel Oil Material Balance

Units (kg/hr)	**IN**		**OUT**	
Components	**Light Steam**	**LD Draw**	**LD Return**	**Light Diesel**
C1	0	0	0	0
C2	0	0	0	0
C3	0	0	0	0
C4	0	0	0	0
C5-C11	0	16525.62	1872.15	14653.41
C12-C17	0	28456.45	596.04	27860.36
C18-C22	0	19331.12	38.84	19292.27
C23-C28	0	14606.27	2.15	14603.89
C29+	0	24208.35	0.029	24208.25
Water	771.1	78.38	770.28	79.2
Total	771.1	103206.19	3279.48	100697.38
Total	**103977.29**		**103977.29**	

4.1.6 Heavy Diesel Stripper

Mass flow rate entering in HD Stripper = Mass flow rate leaving the HD Stripper

Table 4.7: Heavy Diesel Oil Material Balance

Units (kg/hr)	**IN**		**OUT**	
Components	**Steam 3**	**HD Draw**	**HD Return**	**Heavy Diesel**
C1	0	0	0	0
C2	0	0	0	0
C3	0	0	0	0
C4	0	0	0	0
C5-C11	0	1032.5	371.75	660.72
C12-C17	0	3880.99	328.21	3552.73
C18-C22	0	2811.81	19.74	2792.04
C23-C28	0	2138.23	0.98	2137.23
C29+	0	3546.11	0.0395	3546.07
Water	771.1	12.32	771.7	11.63
Total	771.1	13421.96	1492.42	12700.42
Total	**14193.06**		**14193.06**	

4.1.7 _ Kerosene Stripper

Mass flow rate entering in Kero Stripper = Mass flow rate leaving Kero Stripper

Table 4.8: Kerosene Material Balance

Units (kg/hr)	**IN**		**OUT**	
Components	**Steam 1**	**Kero Draw**	**Kero Return**	**Kerosene**
C1	0	0	0	0
C2	0	0.016	0.016	0
C3	0	0.293	0.292	0.0008
C4	0	10	9.27	0.73
C5-C11	0	24607.72	5969.59	18638.13
C12-C17	0	376.45	4.33	372.48
C18-C22	0	2.13	0	2.13
C23-C28	0	0.0097	0	0.0097
C29+	0	0	0	0
Water	635	4.41	602.25	37.2
Total	635	25001.02	6585.74	19050.28
Total	**25636.02**		**25636.02**	

CHAPTER 5

ENERGY BALANCE

5.1 INTRODUCTION

In this chapter, energy balance on main equipment of the crude oil distillation process was carried out. The law of conservation of energy was applied during the calculations that if there is no consumption or generation of energy then the energy of inlet streams will be equal to the energy of outlet streams. Different equations and formulas are used during these calculations.

The energy balance gives us the information about the power required, consumption of fuel and steam generation and it also provides details of energy with product and waste streams.

5.2PUMPS

The two main pumps on which energy balance is carried out are:

1. Crude oil feed pump
2. Heated feed pump to atmospheric tower

The equations required were

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

By ignoring the changings in velocity and losses, and by incorporation pump efficiency, the equation becomes:

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

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

Crude Oil Feed Pump

Table 5.1: Energy Balance for Crude Oil Feed Pump P-001

PUMP P-001	**UNITS**	**IN**	**OUT**
Temperature	C	26.67	26.67
Pressure	kPa	101.30	1732.00
Mass Flow rate	kg/h	172144.00	172144.00
Density	kg/m3	812.00	812.00
DUTY	**KJ/s**	**128.04**	

Heated feed pump to atmospheric tower

Table5.2: Energy Balance for Heated feed pump to atmospheric tower P-002

PUMP 2	**UNITS**	**IN**	**OUT**
Temperature	C	231.10	231.30
Pressure	kPa	411.60	914.90
Mass Flow rate	kg/h	137073.43	137073.43
Density	kg/m3	678.00	678.00
DUTY	**KJ/s**	**37.69**	

5.3 HEAT EXCAHNGER

The energy balance on the heat exchangers are applied to calculate the heat loads and the general equation used for calculation is:

$$\mathbf{Q} = \dot{\mathbf{m}}(\boldsymbol{\Delta}\mathbf{H})$$

$$\mathbf{Q} = \dot{\mathbf{m}}\boldsymbol{Cp}(\boldsymbol{\Delta T})$$

Heat Exchanger 101		**Tube Side Crude**		**Shell Side LD**	
	UNITS	IN	OUT	IN	OUT
Vap Fraction	-	0.00	0.00	0.00	0.00
Temperature	C	26.67	137.20	232.00	64.62
Flow rate	kg/h	172144	172144	100697.3	100697.3
Cp	KJ/Kg-C	1.94	2.44	2.75	2.18
Energy	**KJ/s**	**11527.96**		**11527.73**	

Table5.3: Energy Balance for heat exchanger P-101

5.4 PREFLASH TOWER FURNACE

Furnaces play an important role in crude oil refining process by providing the necessary heat for various thermal processes such as thermal reforming and thermal cracking. The amount of fuel required for the furnace to heat the given amount of feed is calculated and also the amount of air required is found through calculations. The basic formulas used for calculations are:

Table 0-4: Energy Balance for Pre-Flash tower furnace

Atmospheric Tower Furnace			
	UNITS	**Crude IN**	**Crude OUT**
Vap. Fraction		0	0.375
Temperature	C	132.07	231.11
Flow Rate	kg/h	171524	171524
Cp	KJ/Kg-C	2.42	2.74
Latent Heat	KJ/Kg	-	1455
Energy	**KJ/s**	**12161.01**	

$$\mathbf{Q} = \dot{\mathbf{m}}(\boldsymbol{\Delta}\mathbf{H})$$

$$\mathbf{Q} = \dot{\mathbf{m}}\boldsymbol{Cp}(\boldsymbol{\Delta T})$$

The formulas used for the calculation of fuel requirement are:

$$Q=m(HVV)$$

$$m=Q/HVV$$

Table 5.5: Pre-Flash tower furnace fuel requirement

Fuel Requirement		
Fuel Req	kg/hr	793.82
HHV	KJ/Kg	55150.00
Energy	**KJ/s**	**12161.01**

The air required in the furnace for this amount of fuel is:

Table 5.6: Pre-Flash tower furnace air requirement

Combustion Material Balance		
Components	UNITS	
Natural Gas	Kg/hr	793.82
CO2	Kg/hr	2689.5
H20	Kg/hr	2200.5
O2	Kg/hr	2445
N2	Kg/hr	8185
Excess Air	%	20.00
Air Required	**Kg/hr**	10630.28

5.5Pre-Flash Tower

1.1 Pre-flash tower serves the function of separating light naphtha from heavy crude oil and then the heavier products are sent to the atmospheric tower.

1.2 $\mathbf{Q = \dot{m}(\Delta H)}$

1.3 $\mathbf{Q = \dot{m}}\boldsymbol{Cp}(\boldsymbol{\Delta T})$

1.4

Table 5.7: Energy Balance for Pre-flash tower

1.5

Pre-Flash Tower		IN	OUT	
		Crude Oil	Light Naphtha	Pre-Flashed Crude
Vap Fraction		0.375	0.00	1.00
Temperature	C	231.10	113	223
Mass Flow Rate	kg/h	171523.32	34500.37	137022.95
Cp	KJ/Kg-C	2.70	2.60	3.18
Latent Heat	KJ/Kg	1455	707.4	1372
Energy	KJ/s	26513.22	1619.60	24893.62
TOTAL	KJ/s	26513.22	26513.22	

5.6Air Cooler

This air cooler is used to lower the temperature of the top product of pre-flash tower which is the light naphtha. The formulas used for energy balance calculations are:

$$\mathbf{Q = \dot{m}(\Delta H)}$$

$$\mathbf{Q = \dot{m}\boldsymbol{Cp}(\boldsymbol{\Delta T})}$$

Table 5.8: Energy Balance for Air Cooler

		Crude		**Air**	
Air Cooler	UNITS	IN	OUT	IN	OUT
Vap Fraction		1	0.00	0.00	0.00
Temperature	C	113	65	25	32.2
Flow rate	kg/h	34450	34450	1481099	1481099
Cp	KJ/Kg-C	2.47	2.28	1.00	1.03
Latent Heat	KJ/Kg	-	719.9	**-**	
Energy	**KJ/s**	**2962.2**		**2962.2**	

5.7ATMOSPHERIC TOWER FURNACE

Furnaces play an important role in crude oil refining process by providing the necessary heat for various thermal processes such as thermal reforming and thermal cracking. The amount of fuel required for the furnace to heat the given amount of feed is calculated and also the amount of air required is found through calculations. The basic formulas used for calculations are:

$$\mathbf{Q = \dot{m}(\Delta H)}$$

$$\mathbf{Q = \dot{m}}\boldsymbol{Cp(\Delta T)}$$

Table 5.9. Energy Balance for Atmospheric tower furnace

Atmospheric Tower Furnace			
	UNITS	**Crude IN**	**Crude OUT**
Vap. Fraction		0	0.67
Temperature	C	228.40	354.44
Flow Rate	kg/h	137024.49	137024.49
Cp	KJ/Kg-C	2.77	3.03
Latent Heat	KJ/Kg	-	1417
Energy	**KJ/s**	**13903.44**	

The formulas used for the calculation of fuel requirement are:

$$Q=m(HVV)$$

$$m=Q/HVV$$

Table 5.10: Atmospheric tower furnace fuel requirement

Fuel Requirement		
Fuel Req	kg/hr	907.56
HHV	KJ/Kg	55150.00
Energy	**KJ/s**	**13903.44**

The air required in the furnace for this amount of fuel is:

Table 5.11: Atmospheric tower furnace air requirement

Combustion Material Balance		
Components	UNITS	
Natural Gas	Kg/hr	907.56
CO2	Kg/hr	3074.50
H20	Kg/hr	2515.50
O2	Kg/hr	2795.19
N2	Kg/hr	9357.19
Excess Air	%	20.00
Air Required	**Kg/hr**	**12153.41**

5.8 ATMOSPHERIC TOWER

Atmospheric tower is the main distillation tower in crude oil refining process. Atmospheric tower servers to separate heavy crude into various products like kerosene, light diesel oil, heavy diesel oil and residue.

$$\mathbf{Q} = \dot{\mathbf{m}}(\boldsymbol{\Delta}\mathbf{H})$$

$$\mathbf{Q} = \dot{\mathbf{m}}\boldsymbol{Cp}(\boldsymbol{\Delta T})$$

Table 5.12: Energy Balance for Atmospheric Tower IN

IN						
		Crude IN	**Kerosene Stripper Steam**	**LD Stripper Steam**	**HD Stripper Steam**	**DT Steam**
Vap. Fraction		0.67	1	1	1	1
Temperature	C	354.40	257.20	260.00	260.00	260.00
Pressure	kPa	287.50	287.50	446.10	446.10	446.10
Flow Rate	Kg/hr	137023	635.00	771.10	771.10	771.10
Cp	KJ/Kg-C	3.02	2.00	2.01	2.01	2.01
Latent Heat	KJ/Kg	1416.2	2189	2143	2143	2143
Energy	kJ/s	40101	2190.56	2680.56	2680.56	2680.5
Total	**KJ/s**	**50331.12**				

Table 5.13: Energy Balance for Atmospheric Tower OUT

OUT									
	UNITS	**Off Gases**	**Condenser**	**Water out**	**Heavy Naphtha**	**Kero**	**Light Diesel**	**Heavy Diesel**	**Residue**
Vap. Fraction		1		0	0	0	0	0	0
Temperature	C	45.42		45.42	45.42	129.60	232.00	214.80	207.50
Pressure	kPa	211.6		211.60	211.60	269.10	290.30	300.90	315.10
Flow Rate	Kg/hr	113.4		2827	1588	19050	100700	12700	3045
Cp	KJ/Kg-C	1.76		4.32	2.31	2.52	2.76	2.68	2.64
Latent Heat	KJ/Kg	534.7		2217	467.1	499.6	1374	1460	1512
Energy	kJ/s	70.75	8011.4	8437	1003	7009.7	19724.6	4138.90	1505
Total	KJ/s	**49827.32**							

CHAPTER 6

EQUIPMENT DESIGN

6.1 Equipment Design

This section includes the design of major components within the proposed process. The design of four key components is part of this chapter.

1. Desalter
2. Heat Exchanger
3. Pre-Flash Tower
4. Atmospheric Crude Tower

6.1.1 Desalter

The design of a desalter is very crucial because it is the main equipment to remove salts from the crude oil that can lead to corrosion in the overhead section of the distillation column.

Table 6.1 : Desalter conditions

Desalter Conditions	
Temperature (°C)	133.4
Pressure (bar)	11.7
Oil density ρ_g (kg/m^3)	725.4
Water density ρ_l (kg/m^3)	935.5
Superficial velocity (m/s)	1.49

So first we calculate the superficial velocity of oil-water mixture entering in the desalter:

$$Superficial\,Velocity = u_t = \frac{d^2 g(\rho_{water} - \rho_{oil})}{18\,\mu_{oil}}$$

$$\boldsymbol{u_t = 1.49\ \frac{m}{s}}$$

By using superficial velocity, the drag coefficient is calculated:

$$C_D \rho_{water}.\frac{\pi(d_m)^2}{4}.\ \frac{(u_t)^2}{2} = \ |\rho_{water} - \rho_{oil}|.\,g.\frac{\pi(d_m)^3}{6}$$

$$\boldsymbol{\frac{C_D}{d_m} = 0.6}$$

Desalter Design Equation:

$$3d^2 = 421\left(\frac{TZQ_g}{P}\right)\sqrt{\frac{\rho_{oil}}{\rho_{water} - \rho_{oil}}.\frac{C_D}{d_m}}$$

Now by using the desalter design equation, the length and diameter of the desalter is calculated:

$$\mathbf{Diameter\ of\ Desalter = d = 4.56\ ft = 1.51\ meters}$$

$$\mathbf{Length\ of\ Desalter = L = 3d = 4.53\ meters}$$

6.1.2 Heat Exchanger

In the crude oil refining process, heat exchangers are crucial components that facilitate the transfer of thermal energy between different process streams. Designing a heat exchanger that utilizes kerosene as the heating medium to heat the crude oil feed offers several advantages, including efficient heat transfer and enhanced energy utilization. This section focuses on the design considerations and principles involved in such a heat exchanger.

The tube side data is :

Table 6.2: Tube side data

Tube Side	
Crude oil	
Inlet Temp(t_1) ^{0}C	27.1
Exit Temp(t_2) ^{0}C	56.1
Mass Flow Rate(W) (kg/hr)	71882
Viscosity (cP)	1.79
Specific gravity	0.81
Thermal Conductivity (W/m.K)	0.151
C_p (kJ/kg.^{0}C)	1.95
Fouling Factor (m^2. ^{0}C/ W)	5.3 x 10-4

The shell side data is:

Table 6. 3: Shell side data

Shell Side	
Hot Kerosene Product	
Inlet Temp(t_1) ^{0}C	165
Exit Temp(t_2) ^{0}C	55.03
Mass Flow Rate(W) (kg/hr)	16585
Viscosity (cP)	1.64
Specific gravity	0.79

Thermal Conductivity (W/m.K)	0.145
Cp (kJ/kg. ^{0}C)	2.1
Fouling Factor (m2. ^{0}C/ W)	1.8 x 10-4

We used the kern's method to design the heat exchanger:

True Log Mean Temperature Difference:

$$LMTD = \frac{110-28}{\ln(\frac{110}{28})}$$

$$\mathbf{LMTD = 59.92}℃$$

Using values of R and S to find Temperature Difference Factor from figure:

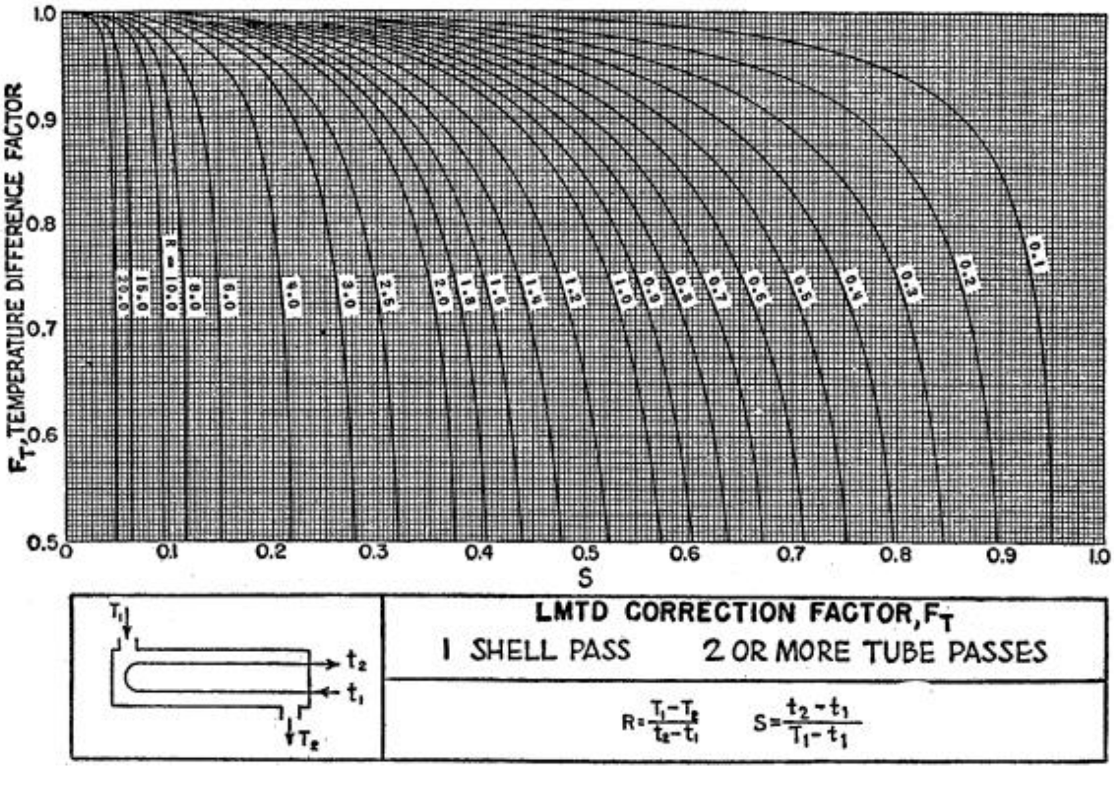


Figure 6.1: LMTD Correction Factor

From figure 6.4:

$$F_T = 0.99$$

$$\text{True LMTD} = (0.98)(59.92) = 59.72℃$$

Assuming the Design Coefficient for further calculations based on literature data for crude oil and kerosene:

$$U_D = 260\mathrm{W/m^2k}$$

Total heat transfer Area:

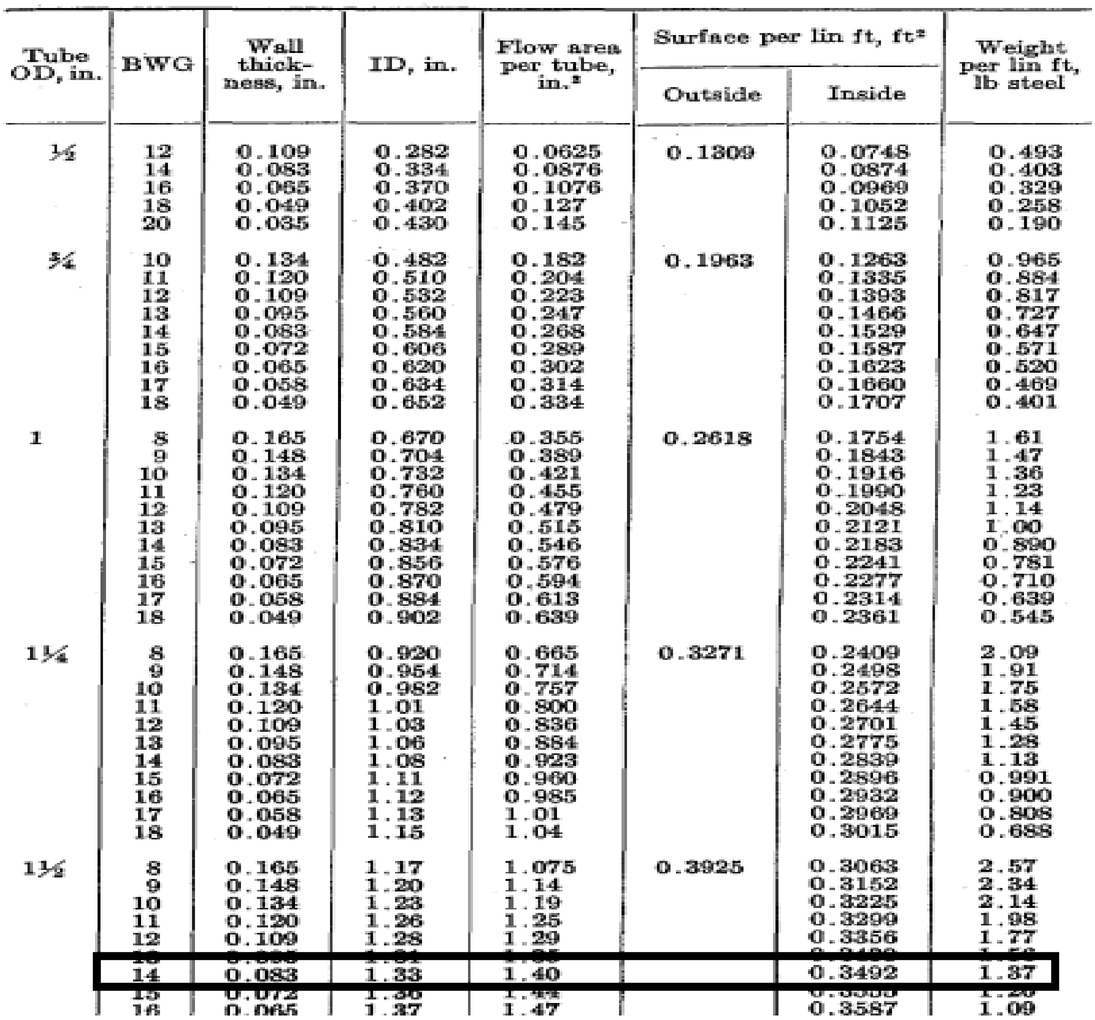
$$\text{Area} = \frac{Q}{U_D * \Delta t}$$

$$\text{Area} = \frac{1062256}{(260)(58.72)}$$

$$\mathbf{Area = 68.18m^2}$$

Tube Section Calculations:

Assuming Tube sizing:



Tube OD, in.	BWG	Wall thick-ness, in.	ID, in.	Flow area per tube, in.^{2}	Surface per lin ft, ft^2		Weight per lin ft, lb steel
					Outside	Inside	
½	12	0.109	0.282	0.0625	0.1309	0.0748	0.493
	14	0.083	0.334	0.0876		0.0874	0.403
	16	0.065	0.370	0.1076		0.0969	0.329
	18	0.049	0.402	0.127		0.1052	0.258
	20	0.035	0.430	0.145		0.1125	0.190
¾	10	0.134	0.482	0.182	0.1963	0.1263	0.965
	11	0.120	0.510	0.204		0.1335	0.884
	12	0.109	0.532	0.223		0.1393	0.817
	13	0.095	0.560	0.247		0.1466	0.727
	14	0.083	0.584	0.268		0.1529	0.647
	15	0.072	0.606	0.289		0.1587	0.571
	16	0.065	0.620	0.302		0.1623	0.520
	17	0.058	0.634	0.314		0.1660	0.469
	18	0.049	0.652	0.334		0.1707	0.401
1	8	0.165	0.670	0.355	0.2618	0.1754	1.61
	9	0.148	0.704	0.389		0.1843	1.47
	10	0.134	0.732	0.421		0.1916	1.36
	11	0.120	0.760	0.455		0.1990	1.23
	12	0.109	0.782	0.479		0.2048	1.14
	13	0.095	0.810	0.515		0.2121	1.00
	14	0.083	0.834	0.546		0.2183	0.890
	15	0.072	0.856	0.576		0.2241	0.781
	16	0.065	0.870	0.594		0.2277	0.710
	17	0.058	0.884	0.613		0.2314	0.639
	18	0.049	0.902	0.639		0.2361	0.545
1¼	8	0.165	0.920	0.665	0.3271	0.2409	2.09
	9	0.148	0.954	0.714		0.2498	1.91
	10	0.134	0.982	0.757		0.2572	1.75
	11	0.120	1.01	0.800		0.2644	1.58
	12	0.109	1.03	0.836		0.2701	1.45
	13	0.095	1.06	0.884		0.2775	1.28
	14	0.083	1.08	0.923		0.2839	1.13
	15	0.072	1.11	0.960		0.2896	0.991
	16	0.065	1.12	0.985		0.2932	0.900
	17	0.058	1.13	1.01		0.2969	0.808
	18	0.049	1.15	1.04		0.3015	0.688
1½	8	0.165	1.17	1.075	0.3925	0.3063	2.57
	9	0.148	1.20	1.14		0.3152	2.34
	10	0.134	1.23	1.19		0.3225	2.14
	11	0.120	1.26	1.25		0.3299	1.98
	12	0.109	1.28	1.29		0.3356	1.77
	[illegible]	[illegible]	[illegible]	[illegible]		[illegible]	[illegible]
	14	0.083	1.33	1.40		0.3492	1.37
	[illegible]	0.072	[illegible]	[illegible]		[illegible]	[illegible]
	16	0.065	1.37	1.47		0.3587	1.09

Figure 6.2: Heat Exchanger and Condenser Tube Data

We took the BWG 14 Tube:

Tube outer diameter = 0.038 m

Wall Thickness = 0.002 m

Tube Internal Diameter = 0.03 m

Flow area per tube = 0.04 m^2

Flow area:

$$\text{Flow Area} = \frac{N_t a_t}{144n} = 0.206\text{m}^2$$

Mass flowrate:

From the material balance, total mass flowrate is 71882 kg/hr

Mass velocity:

$$G_t = \frac{w}{a_t} = 3.47 * 10^5 \frac{\text{kg}}{\text{hr.m}^2}$$

Number of tubes:

$$N_t = \text{Mass}\frac{\text{flowrate}}{\text{Mass velocity} * \text{Flow area per tube}} = 232$$

Length of a single tube:

$$\text{Total Length of tubes} = \frac{\text{Area}}{\pi * \text{OD}} = \frac{68.1}{(\pi)(0.0381)}$$

$$\text{Total Length of tubes} = 569.9\text{m}$$

$$\text{Length of single tube} = \frac{569.9}{232} = 2.46\text{ m}$$

Reynolds Number:

$$\text{Re}_t = \frac{DG_t}{\mu} = 1817.51$$

Determining Prandtl Number:

$$Pr = \frac{\mu \times \text{Cp}}{\text{K}} = 0.0231$$

Determining Chiltern-Colburn Factor (j_H)

Using our calculated Reynolds Number and figure 6 we calculate the j_H factor.

$$j_H = 61$$

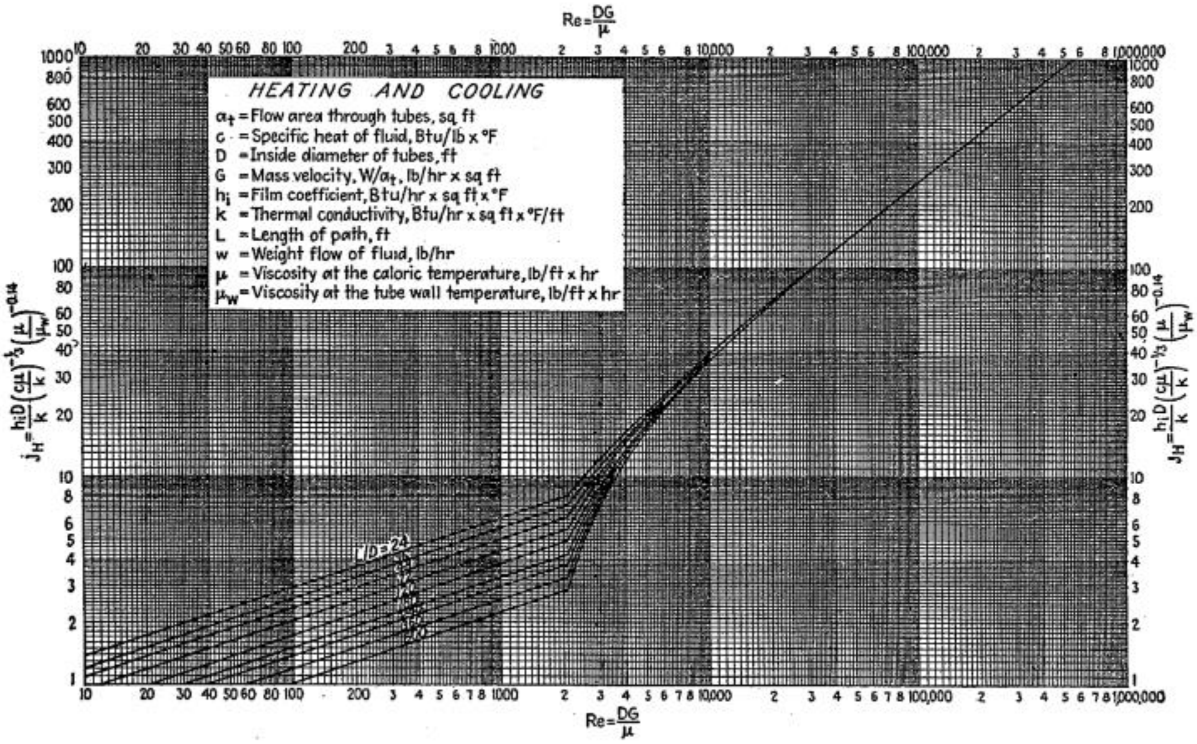


Figure 6.3: jH vs Re curve for Tube Side

Calculating hi:

$$h_i = j_H\left(\frac{k}{D}\right) * \left(\frac{c\mu}{k}\right)^{\frac{1}{3}} * \varphi_t$$

$$hi = 781.14\ W/m^2k$$

$$h_{io} = h_i * \frac{ID}{OD}$$

$$h_{io} = (781.14)\left(\frac{0.337}{0.0381}\right) = 1457.89\ W/m^2k$$

Friction factor for kerosene:

$$f = 0.00018$$

Pressure Drop:

$$\Delta P_t = \frac{f G_t^2 L n}{5.22 * 10^{10} D_S \varphi_t} = 2.3\, psi$$

Shell Side calculations:

Assumed the Internal Diameter of the shell using relevant data from literature:

$$\mathrm{ID} = 1.12\mathrm{m}$$

Determining Pitch:

$$\mathrm{P_T} = (1.25 * 0.0381)$$

$$\mathrm{P_T} = 0.0476\ \mathrm{m}$$

Determining Clearance:

$$\mathrm{C} = \mathrm{P_T} - \mathrm{OD(Tube)}$$

$$\mathrm{C} = (1.25 * 0.0381) - (0.0381) = 0.0095\ \mathrm{m}$$

Baffles Calculations:

$$\text{Baffle spacing} = 0.536\ \mathrm{m}$$

$$\text{Baffle thickness} = (\text{tube thickness})(2.5) = (0.0021)(2.5) = 0.0052\mathrm{m}$$

$$\text{Baffle Height} = (0.75)(\text{Internal shell diameter}) = (0.75)(1.12) = 0.843\mathrm{m}$$

Number of Baffles:

$$\mathrm{B} = \frac{\text{length of the tube}}{(\text{Baffle spacing})(\text{Baffle thickness})} = 4$$

Flow area:

$$\mathrm{a_s} = \frac{\mathrm{ID} * \mathrm{CB}}{\mathrm{P_T}} = 0224\mathrm{m}^2$$

Mass Flowrate:

Mass flowrate (W) from the material balance is 16585 kg/hr

Mass Velocity:

$$G_s = \frac{W}{a_s} = 73918.8\ \frac{kg}{hr.m^2}$$

Equivalent Diameter:

$$d_e = 0.0269\ m$$

The Equivalent Diameter was found using figure 4:

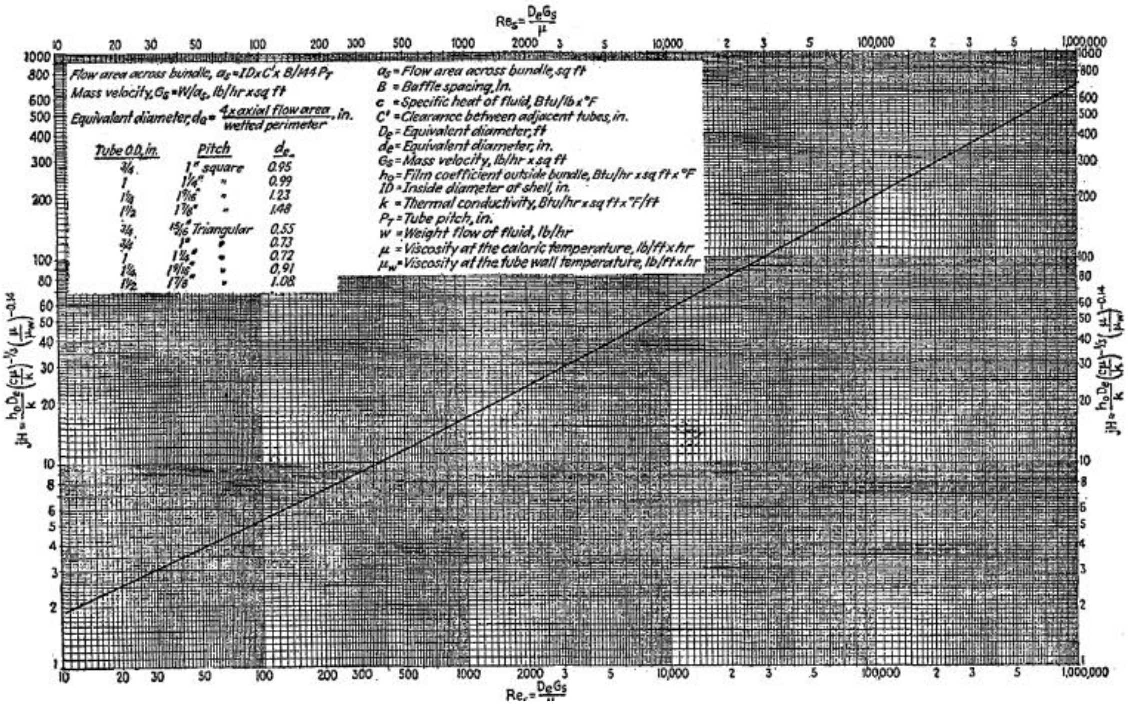


Figure 6.4: jH vs Re curve for Shell Side

Reynolds Number:

$$\mathrm{Re_s} = \frac{\mathrm{D_e G_s}}{\mu} = 337$$

Determining Prandtl Number:

$$Pr = \frac{\mu \times \mathrm{Cp}}{\mathrm{K}} = 0.0236$$

Determining Chiltern-Colburn Factor ($\mathrm{j_H}$):

Using the Reynolds number, we were able to find the $\mathrm{j_H}$.

$$\mathrm{j_H} = 36$$

Calculating the $\mathrm{h_o}$ factor:

$$\mathrm{h_o} = \mathrm{j_H} \left(\frac{\mathrm{k}}{\mathrm{D_e}}\right) \left(\frac{\mathrm{c}\mu}{\mathrm{k}}\right)^{\frac{1}{3}} = 556.69\ \mathrm{W/m^2k}$$

Friction factor for shell side:

$$f = 0.00013$$

Pressure Drop:

$$\Delta P_s = \frac{f G_s^2 D_e (N+1)}{5.22 * 10^{10} D_e s \varphi_t} = 9.6\, psi$$

Design Overall Coefficient:

$$U_C = \frac{h_{io} h_o}{h_{io} + h_o} = 308.59\ \ W/m^2 k$$

Dirt Factor:

$$R_d = \frac{U_C - U_D}{U_C U_D} = 4.7 * 10^{-3}$$

Overall Summary of the Design Specifications:

Table 6.4: Heat Exchanger Design Specifications

Type	Shell and tube
LMTD (^{0}C)	59.92
Heat Transfer Area (m^2)	68.18
Heat Duty (KJ/hr)	3824122.4
No of tubes	232 tubes
Length of a tube (m)	2.46
Tube Outer Diameter (mm)	38.1
Tube Inner Diameter (mm)	33.7
Tube Pattern	Triangular Pitch
Tube Pitch (mm)	46.6
Passes	1
Shell Diameter (m)	1.1242
No of Baffles	4
Baffle Spacing (m)	0.536
Baffle Thickness (m)	0.0052
Baffle height (m)	0.843
Tube side Pressure Drop (psi)	2.3
Shell Side Pressure Drop (psi)	9.6

6.1.3 Pre-Flash Tower

Pre-Flash Tower is a distillation column that is designed to separate Light Naphtha and Heavy Crude. Light Naphtha obtained from pre-flash tower is further treated to yield Liquified Petroleum Gas (LPG). It works on the principle of volatility difference. While Designing the Pre-Flash Tower, first of all Light key and Heavy key components were decided. After selection of Light and Heavy key components, compositions of top and bottom streams were determined. Bubble Point and Dew Point for the column were calculated using Antonine's equation. Relative Volatility for each component was calculated with reference to Heavy key component. Fenske's equation was used to calculate minimum number of stages while Underwood's equation was used to calculate Minimum Reflux Ratio. O'Connell relation was used to calculate Actual number of stages. Furthermore, Height and Diameter of Column were calculated.

The column top temperature was obtained by carrying out dew point calculations at several temperatures and then choosing the one for which the following relation was satisfied:

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

For our system, this relation was satisfied at 63°C.

Similarly, the column bottom temperature was obtained by carrying out bubble point calculations at several temperatures and then choosing the one for which the following relation was satisfied:

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

For our system, this relation was satisfied at 281.5 °C.

The minimum number of plates were calculated using Fenske's equation.

$$\mathrm{N_m} = \frac{\log\left(\frac{x_D}{x_B}\right)\left(\frac{y_B}{y_D}\right)}{\log \alpha_{LK}}$$

Minimum number of plates came out to be 9.63.

The minimum reflux ratio was calculated using the Underwood's relation.

$$\sum \frac{\alpha_i x_{if}}{\alpha_i - \theta} = 1 - q$$

$$\sum \frac{\alpha_i x_{id}}{\alpha_i - \theta} = R_m + 1$$

The minimum reflux ratio came out to be 0.71.

O'Connell relation was used to calculate column efficiency and actual number of plates.

$$E_o = 51 - 32.5 * \log(\mu_a \alpha_a)$$

Actual number of plates were equal to 30.

The Diameter of Column was calculated using the following relation.

$$D_c = \sqrt{\frac{4V_w}{\pi \boldsymbol{\rho}_v u_v}}$$

In the above relation u_v was calculated by:

$$u_v = (-0.171 l_t^2 + 0.27 l_t - 0.047)((\frac{\boldsymbol{\rho}_L - \boldsymbol{\rho}_v}{\boldsymbol{\rho}_v})\ \hat{}\,0.5)$$

Table 6.5: Design specification sheet for Pre-Flash Tower

Specification Sheet	
Theoretical Stages	17
Actual Stages	30
Plate type	Valve
Reflux Ratio	0.99
Column Diameter (m)	1.5
Column Height (m)	23.5

6.1.4 Atmospheric Crude Tower

Atmospheric Tower is a distillation column that separates crude oil into its fractions. To design Atmospheric Tower, an approach like the one discussed for Pre-Flash was used. To begin with, Light key and Heavy key components were decided. After selection of Light and Heavy key components, compositions of top and bottom streams were determined. Bubble Point and Dew Point for the column were calculated using Antonine's equation. Relative Volatility for each component was calculated with reference to Heavy key component. Fenske's equation was used to calculate minimum number of stages while Underwood's equation was used to calculate Minimum Reflux Ratio. O'Connell relation was used to calculate Actual number of stages. Furthermore, Height and Diameter of Column were calculated.

The column top temperature was obtained by carrying out dew point calculations at several temperatures and then choosing the one for which the following relation was satisfied:

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

For our system, this relation was satisfied at 71.2 °C.

Similarly, the column bottom temperature was obtained by carrying out bubble point calculations at several temperatures and then choosing the one for which the following relation was satisfied:

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

For our system, this relation was satisfied at 345.8 °C.

The minimum number of plates were calculated using Fenske's equation.

$$\mathrm{N_m} = \frac{\log\left(\frac{\mathrm{x_D}}{\mathrm{x_B}}\right)\left(\frac{\mathrm{y_B}}{\mathrm{y_D}}\right)}{\log \alpha_{\mathrm{LK}}}$$

Minimum number of plates came out to be 12.96.

The minimum reflux ratio was calculated using the Underwood's relation.

$$\sum \frac{\alpha_\mathrm{i} \mathrm{x_{if}}}{\alpha_\mathrm{i} - \theta} = 1 - \mathrm{q}$$

$$\sum \frac{\alpha_\mathrm{i} \mathrm{x_{id}}}{\alpha_\mathrm{i} - \theta} = \mathrm{R_m} + 1$$

The minimum reflux ratio came out to be 1.5.

O'Connell relation was used to calculate column efficiency and actual number of plates.

$$E_o = 51 - 32.5 * \log(\mu_a \alpha_a)$$

Actual number of plates were equal to 40.

The Diameter of Column was calculated using the following relation.

$$D_c = \sqrt{\frac{4V_w}{\pi \boldsymbol{\rho}_v u_v}}$$

In the above relation u_v was calculated by:

$$u_v = (-0.171 l_t^2 + 0.27 l_t - 0.047)((\frac{\boldsymbol{\rho}_L - \boldsymbol{\rho}_v}{\boldsymbol{\rho}_v})\ \hat{}0.5)$$

Table 6.6: Design specification sheet for Atmospheric Tower

Specification Sheet	
Theoretical Stages	24
Actual Stages	40
Plate type	Valve
Reflux Ratio	1.5
Column Diameter (m)	3.3
Column Height (m)	30

CHAPTER 7

SIMULATION

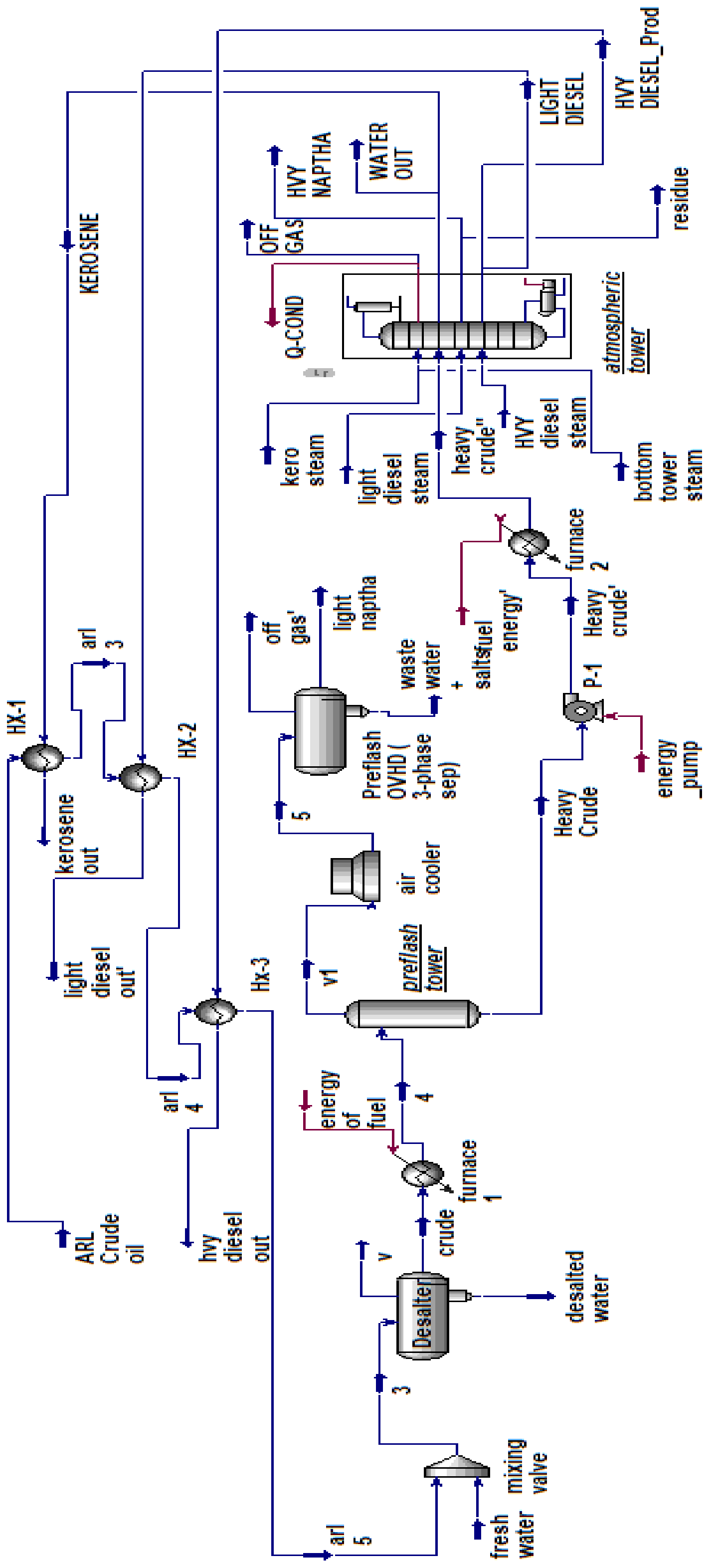
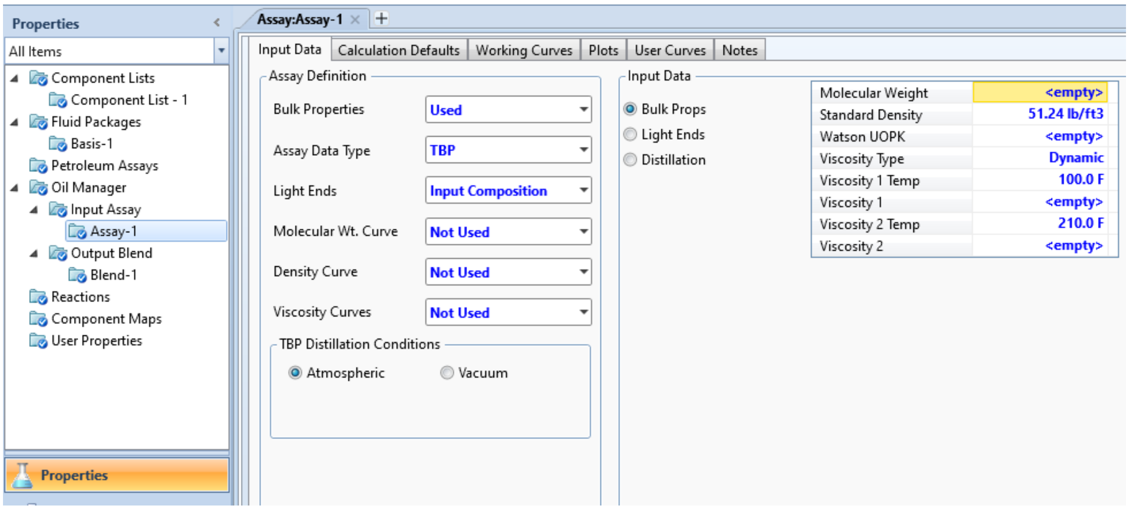


Figure 7.1: simulation of whole plant

7.1 Petroleum Essay input:

To input petroleum essay we following steps on Aspen Hysys were performed:

i. Open Aspen Hysys V11.
ii. Click on Properties tab at left bottom corner.
iii. Click on component list and add all the lighter fractions one by one.
iv. Add oxygen, nitrogen, Sulphur, methyl amine, HCl components also.
v. Now go to the oil manager, click on input assay and generate "assay-1".
vi. On input taskbar, select assay data type as "TBP".
vii. Write the properties of crude oil such standard density in "bulk props" , composition of light ends in "light ends" and assay percentage in "distillation" checkbox.



Assay Definition

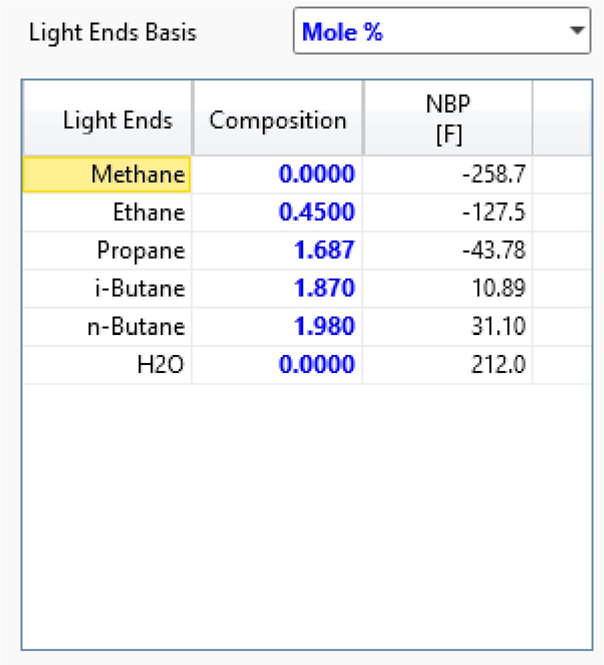
Bulk Properties	Used
Assay Data Type	TBP
Light Ends	Input Composition
Molecular Wt. Curve	Not Used
Density Curve	Not Used
Viscosity Curves	Not Used

TBP Distillation Conditions: ● Atmospheric ○ Vacuum

Input Data: ● Bulk Props ○ Light Ends ○ Distillation

Molecular Weight	<empty>
Standard Density	51.24 lb/ft3
Watson UOPK	<empty>
Viscosity Type	Dynamic
Viscosity 1 Temp	100.0 F
Viscosity 1	<empty>
Viscosity 2 Temp	210.0 F
Viscosity 2	<empty>

Figure7.2: assay generation 1



Light Ends Basis: Mole %

Light Ends	Composition	NBP [F]
Methane	0.0000	-258.7
Ethane	0.4500	-127.5
Propane	1.687	-43.78
i-Butane	1.870	10.89
n-Butane	1.980	31.10
H2O	0.0000	212.0

Figure 7.3: composition of light end

viii. Now click on calculate to make petroleum assay according to properties of crude oil.

ix. Go to output blend beneath input assay in oil manager and add “assay-1” that we just generated.

7.1.1 Following properties we get:

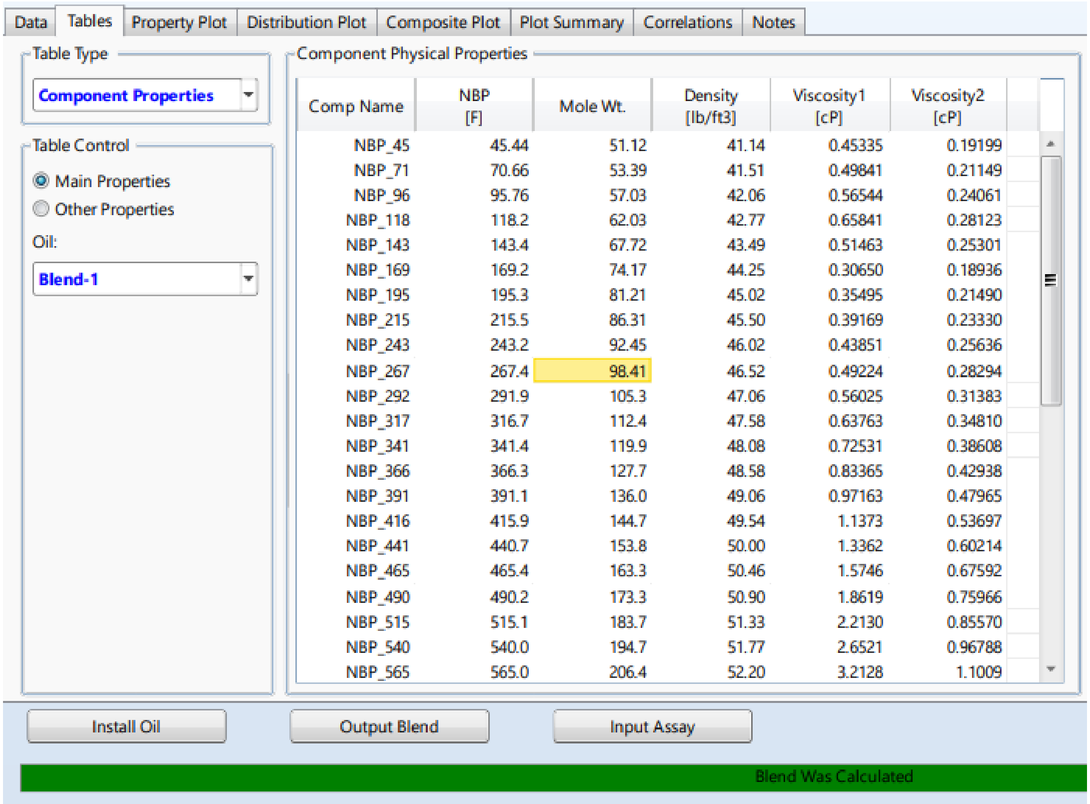


Table Type: Component Properties

Table Control

● Main Properties
○ Other Properties

Oil: Blend-1

Component Physical Properties

Comp Name	NBP [F]	Mole Wt.	Density [lb/ft3]	Viscosity1 [cP]	Viscosity2 [cP]
NBP_45	45.44	51.12	41.14	0.45335	0.19199
NBP_71	70.66	53.39	41.51	0.49841	0.21149
NBP_96	95.76	57.03	42.06	0.56544	0.24061
NBP_118	118.2	62.03	42.77	0.65841	0.28123
NBP_143	143.4	67.72	43.49	0.51463	0.25301
NBP_169	169.2	74.17	44.25	0.30650	0.18936
NBP_195	195.3	81.21	45.02	0.35495	0.21490
NBP_215	215.5	86.31	45.50	0.39169	0.23330
NBP_243	243.2	92.45	46.02	0.43851	0.25636
NBP_267	267.4	98.41	46.52	0.49224	0.28294
NBP_292	291.9	105.3	47.06	0.56025	0.31383
NBP_317	316.7	112.4	47.58	0.63763	0.34810
NBP_341	341.4	119.9	48.08	0.72531	0.38608
NBP_366	366.3	127.7	48.58	0.83365	0.42938
NBP_391	391.1	136.0	49.06	0.97163	0.47965
NBP_416	415.9	144.7	49.54	1.1373	0.53697
NBP_441	440.7	153.8	50.00	1.3362	0.60214
NBP_465	465.4	163.3	50.46	1.5746	0.67592
NBP_490	490.2	173.3	50.90	1.8619	0.75966
NBP_515	515.1	183.7	51.33	2.2130	0.85570
NBP_540	540.0	194.7	51.77	2.6521	0.96788
NBP_565	565.0	206.4	52.20	3.2128	1.1009

Blend Was Calculated

Figure 7.4: composition and properties off crude oil component.1

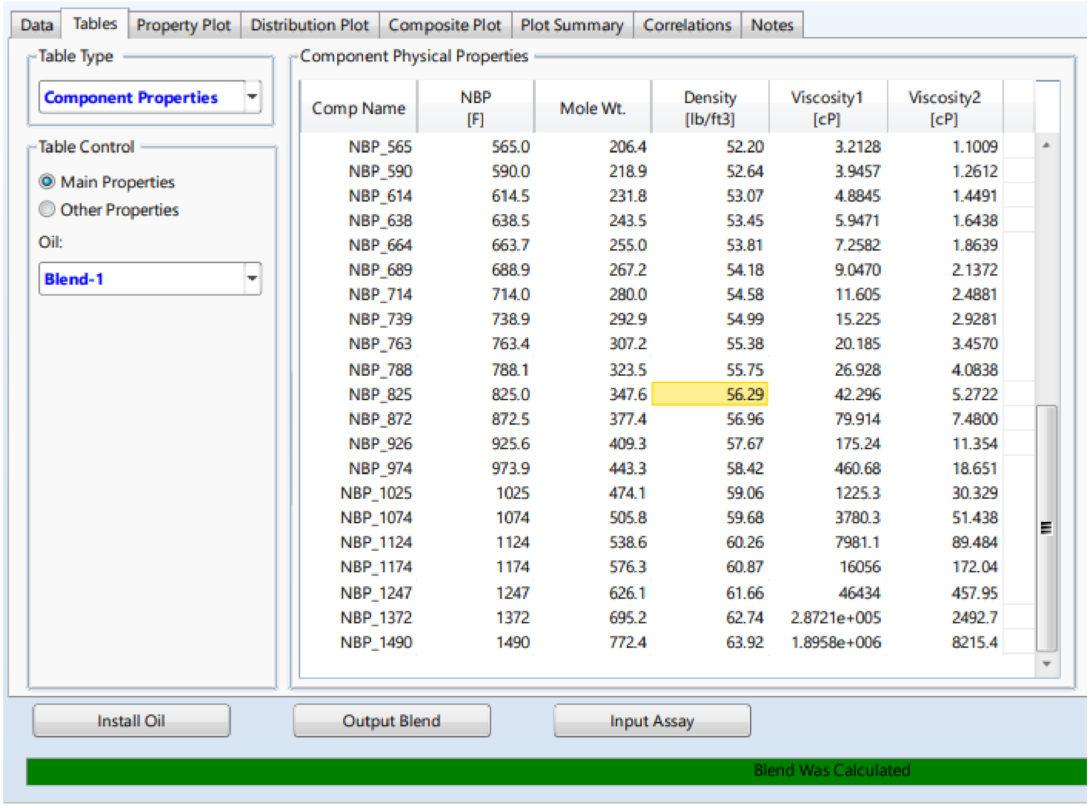


Table Type: Component Properties

Table Control

● Main Properties

○ Other Properties

Oil: Blend-1

Component Physical Properties

Comp Name	NBP [F]	Mole Wt.	Density [lb/ft3]	Viscosity1 [cP]	Viscosity2 [cP]
NBP_565	565.0	206.4	52.20	3.2128	1.1009
NBP_590	590.0	218.9	52.64	3.9457	1.2612
NBP_614	614.5	231.8	53.07	4.8845	1.4491
NBP_638	638.5	243.5	53.45	5.9471	1.6438
NBP_664	663.7	255.0	53.81	7.2582	1.8639
NBP_689	688.9	267.2	54.18	9.0470	2.1372
NBP_714	714.0	280.0	54.58	11.605	2.4881
NBP_739	738.9	292.9	54.99	15.225	2.9281
NBP_763	763.4	307.2	55.38	20.185	3.4570
NBP_788	788.1	323.5	55.75	26.928	4.0838
NBP_825	825.0	347.6	56.29	42.296	5.2722
NBP_872	872.5	377.4	56.96	79.914	7.4800
NBP_926	925.6	409.3	57.67	175.24	11.354
NBP_974	973.9	443.3	58.42	460.68	18.651
NBP_1025	1025	474.1	59.06	1225.3	30.329
NBP_1074	1074	505.8	59.68	3780.3	51.438
NBP_1124	1124	538.6	60.26	7981.1	89.484
NBP_1174	1174	576.3	60.87	16056	172.04
NBP_1247	1247	626.1	61.66	46434	457.95
NBP_1372	1372	695.2	62.74	2.8721e+005	2492.7
NBP_1490	1490	772.4	63.92	1.8958e+006	8215.4

Blend Was Calculated

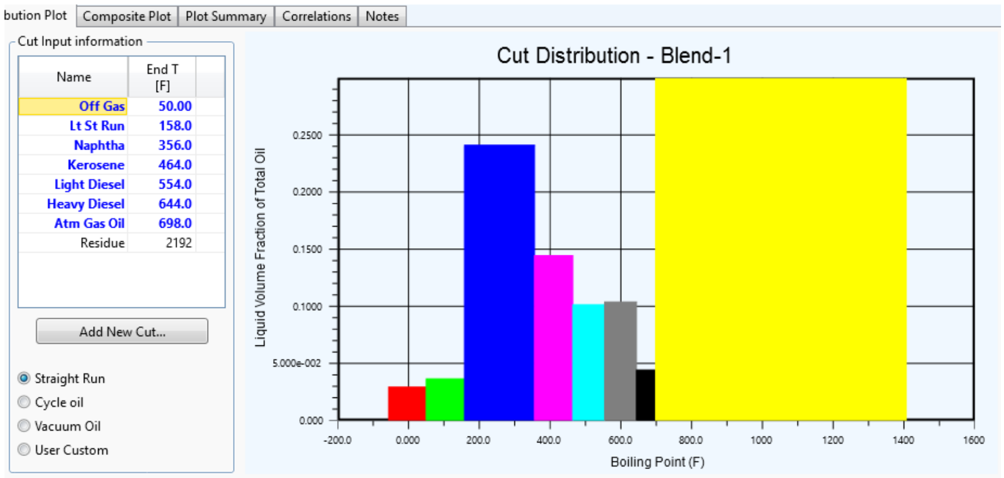
Figure 7.4: composition and properties off crude oil component 2

7.1.2 Product analysis:

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

True-Boiling-Curve:

7.1.3 Distribution plot:

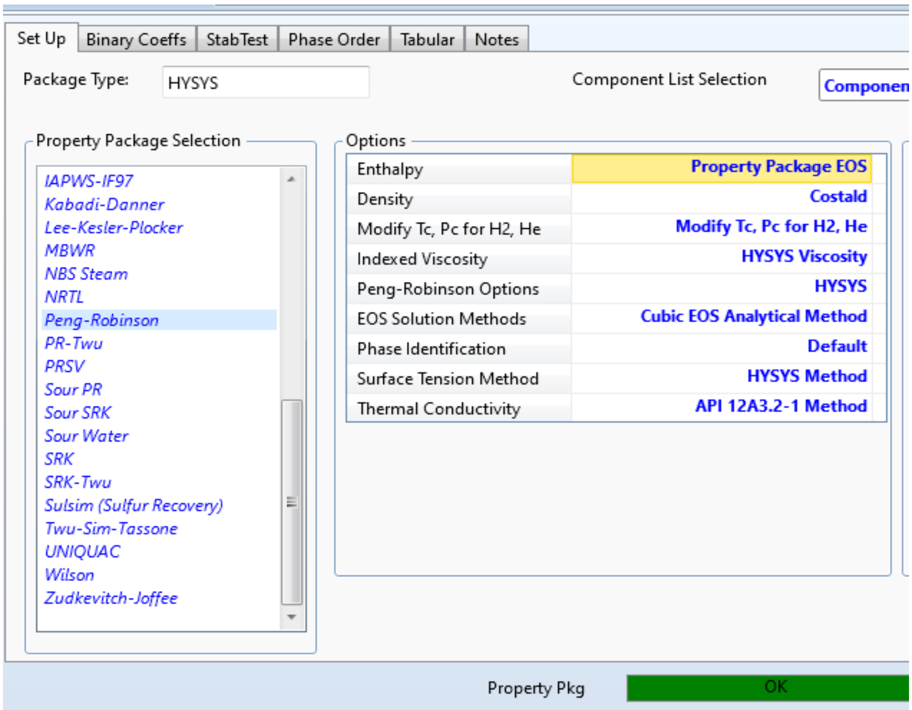


Name	End T [F]
Off Gas	50.00
Lt St Run	158.0
Naphtha	356.0
Kerosene	464.0
Light Diesel	554.0
Heavy Diesel	644.0
Atm Gas Oil	698.0
Residue	2192

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

7.1.4 Fluid Package:

The Peng Robinson fluid package was utilized to calculate Enthalpy and the Cubic equation of state analytical method was employed for property calculations, along with default settings for other parameters.



Package Type: HYSYS

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 7.5: fluid package

Figure illustrates the Fluid Package, which consists of a set of equations that assist the program in determining relevant fluid properties based on specified conditions. Since a process involves transitioning between states, the simulation requires underlying relationships to determine the values of system properties. The choice of fluid package depends on the composition of the components involved.

In the case of Hydrocarbon systems, where the majority of components are nonpolar and the process involves distillation, the Peng Robinson fluid package is sufficient.

7.1.5 Atmospheric Tower:

The atmospheric tower is a crucial component of the refining process, and it features three side strippers for the pump around kerosene, light diesel, and heavy diesel. These side strippers help separate impurities and optimize the quality of the respective products. At the top of the atmospheric tower, the condenser ensures efficient heat exchange and facilitates the separation of the desired products. It produces two main streams: the reflux and the product stream, which consists of heavy naphtha and off gas. The reflux ratio, controlling the amount of liquid sent back to the tower, can be adjusted by regulating the condenser duty. This allows for precise control over the refining process and ensures the production of high-quality products.

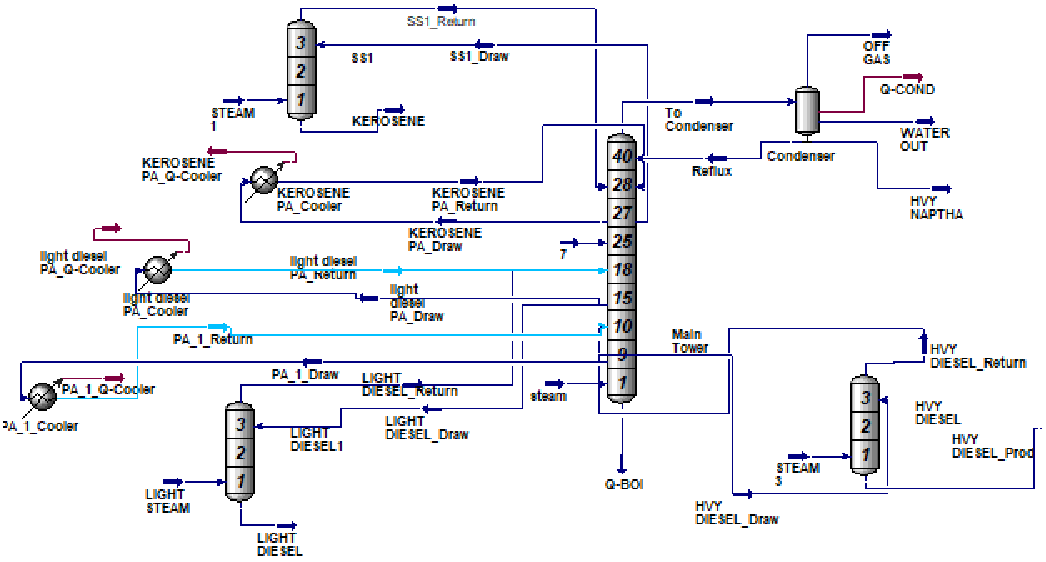
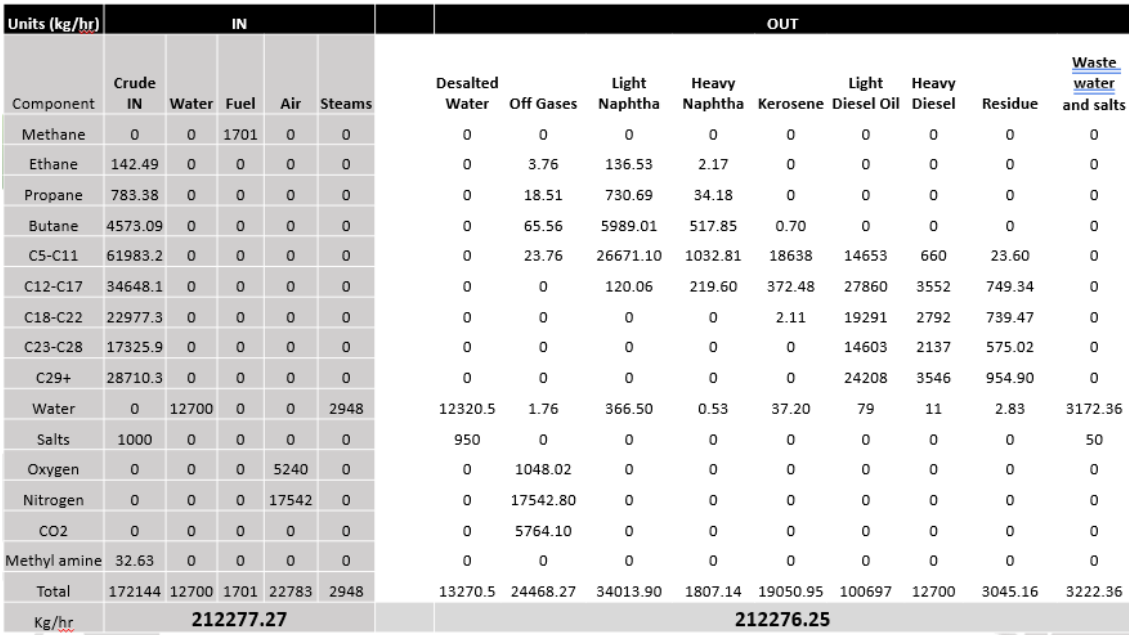


Figure7.6: sub flowsheet of atmospheric tower and stripper

7.1.6 Overall Material balance:

Table 7.1: Overall mass balance from hysys



Units (kg/hr)	IN					OUT								
Component	Crude IN	Water	Fuel	Air	Steams	Desalted Water	Off Gases	Light Naphtha	Heavy Naphtha	Kerosene	Light Diesel Oil	Heavy Diesel	Residue	Waste water and salts
Methane	0	0	1701	0	0	0	0	0	0	0	0	0	0	0
Ethane	142.49	0	0	0	0	0	3.76	136.53	2.17	0	0	0	0	0
Propane	783.38	0	0	0	0	0	18.51	730.69	34.18	0	0	0	0	0
Butane	4573.09	0	0	0	0	0	65.56	5989.01	517.85	0.70	0	0	0	0
C5-C11	61983.2	0	0	0	0	0	23.76	26671.10	1032.81	18638	14653	660	23.60	0
C12-C17	34648.1	0	0	0	0	0	0	120.06	219.60	372.48	27860	3552	749.34	0
C18-C22	22977.3	0	0	0	0	0	0	0	0	2.11	19291	2792	739.47	0
C23-C28	17325.9	0	0	0	0	0	0	0	0	0	14603	2137	575.02	0
C29+	28710.3	0	0	0	0	0	0	0	0	0	24208	3546	954.90	0
Water	0	12700	0	0	2948	12320.5	1.76	366.50	0.53	37.20	79	11	2.83	3172.36
Salts	1000	0	0	0	0	950	0	0	0	0	0	0	0	50
Oxygen	0	0	0	5240	0	0	1048.02	0	0	0	0	0	0	0
Nitrogen	0	0	0	17542	0	0	17542.80	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	5764.10	0	0	0	0	0	0	0
Methyl amine	32.63	0	0	0	0	0	0	0	0	0	0	0	0	0
Total	172144	12700	1701	22783	2948	13270.5	24468.27	34013.90	1807.14	19050.95	100697	12700	3045.16	3222.36
Kg/hr	**212277.27**					**212276.25**								

7.1.7 Overall energy balance:

Table 7.2: Overall energy balance from hysys

Energy In (KJ/s)		Energy Out (KJ/s)	
Pumps	165.73	Products	32371
Wash Water	21732	Water Out	18011
Furnaces	26070	Condensers	8001
Strippers Steams	10631	Off Gases	70.75
Total	58598.73	**Total**	58453.75

CHAPTER 8

ECONOMIC ANALYSIS

8.1 Cost Analysis

The primary objective of any chemical processing plant is to maximize profitability and generate revenues. Chemical manufacturers strive to minimize the cost per unit of the product manufactured while ensuring that both the quality and production capacity remain uncompromised. Prior to initiating any business venture or commencing production on an industrial scale, a comprehensive economic analysis is conducted. The purpose of this analysis is to ascertain the economic feasibility of the venture. It encompasses evaluating the entire plant or specific components that require replacement or optimization to identify superior alternatives. One pivotal parameter in this analysis is cost analysis, which provides investors with vital financial information for assessing various alternatives and making well-informed decisions.

In our case, we performed a thorough economic analysis of the entire plant to assess its potential for successful operation. To accomplish this, we employed the methodology proposed by Coulson and Richardson, which is widely recognized in the field. This rigorous evaluation allowed us to gauge the plant's viability, identifying areas where improvements could be made to enhance cost-effectiveness and ultimately boost profitability. By scrutinizing all aspects of the operation, from raw material procurement to product distribution, we obtained a comprehensive understanding of the plant's financial outlook.

Economic analysis is an essential tool for chemical processing plants, enabling them to strike a balance between cost reduction and maintaining high-quality output and production capacity. By leveraging such analyses, businesses can streamline operations, minimize expenses, and maximize profitability, ultimately positioning themselves for long-term success in the industry.

8.1.1 Desalter

Operating pressure = 12.7 bar

Volume = 9.6 m^3

Material = Carbon Steel

The following figure was used to calculate the cost of the desalter. From the figure by using the volume we find the index and then from the following formula we calculate the cost of the Desalter.

Purchase cost of Desalter(C_e)= C S^n

$$= 1750\ (9.6)^{0.6}$$

$$= \$\ 6798$$

Where, Ce = Purchase cost of equipment

C = Cost constant

S = Characteristics size parameter

n = Index

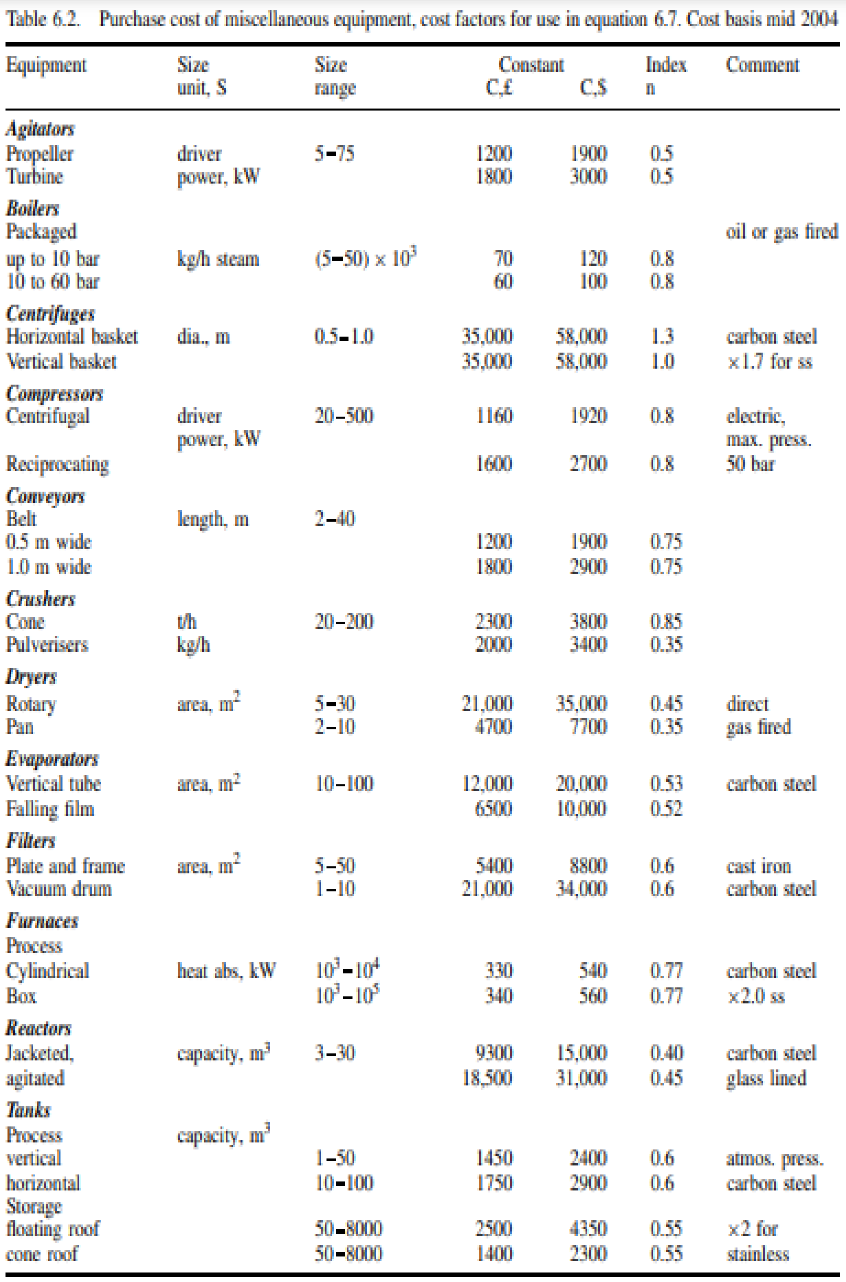


Table 6.2. Purchase cost of miscellaneous equipment, cost factors for use in equation 6.7. Cost basis mid 2004

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. 50 bar
Reciprocating			1600	2700	0.8	
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 8.1: Cost of Desalter

8.1.2 Heat Exchanger

- Shell and tube Heat Exchanger (u-tube)
- Heat Transfer area = 68.18 m^2
- Number of tubes = 232

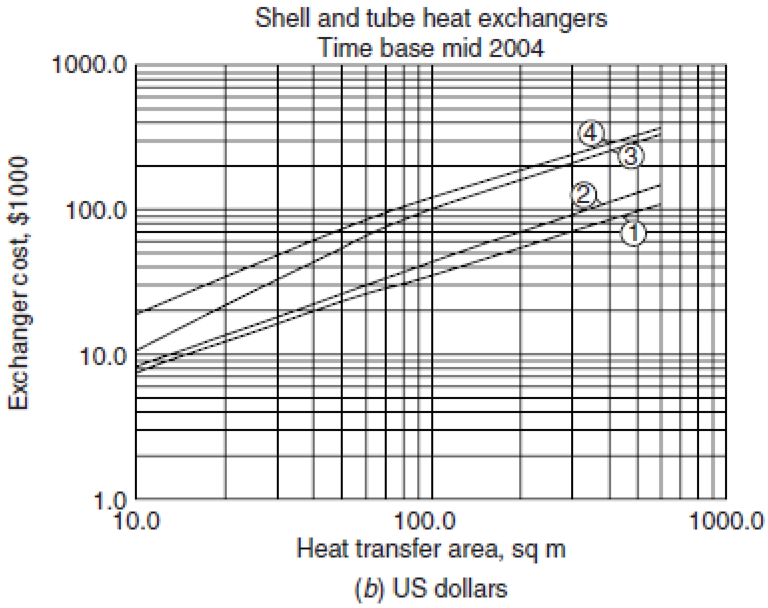
- Pressure = 4 bar
- Material of tubes = Stainless Steel
- Material of Shell = Carbon Steel

From the graph using heat transfer area, we find the cost of heat exchanger. Then we imply different factors to find the cost of heat exchanger.

Cost of HE = Bare Cost*Pressure Factor*Type Factor

= $ 30,000 (0.8)(1)

= $ 24000



Materials: Shell	Materials: 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

Figure8.2: Heat Exchanger cost graph

8.1.3Pre-Flash Tower

- Vessel Diameter = 1.52 m
- Height of vessel = 23.47 m
- Pressure of vessel =4. 116 bars
- Material of vessel = Carbon Steel

- Plate diameter = 1.52 m
- Plate type = Valve
- Number of trays = 32
- Material of Trays = Stainless Steel

For the cost of pre-flash tower, we use two graphs. From the one graph we find the cost of the vessel. And from the other graph we find the cost of the one plate from which we find the cost of all plates.

Cost of Vessel = Bare Cost*Pressure Factor*Type Factor

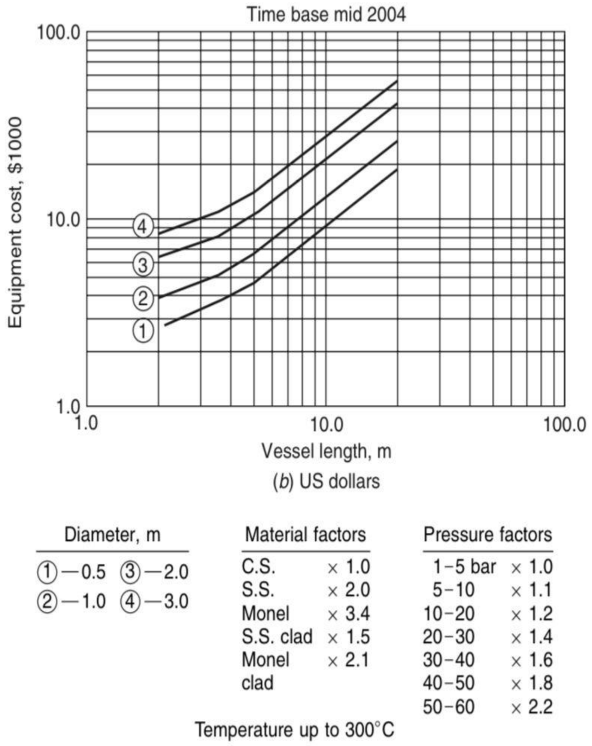
= $ 7000 (1)(1)

= $ 7000

Cost of Plates = 32(Bare Cost*Material Factor)

= $ 21760

Total Cost of Stripper = $ 28760



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

Temperature up to 300°C

Figure 6.6*a*, *b*. Horizontal pressure vessels. Time base mid-2004.
Purchase cost = (bare cost from figure) × Material factor × Pressure factor

Figure8.3:Vessel cost graph

8.1.4 Atmospheric Tower

- Vessel Diameter = 3 m
- Height of vessel = 28 m
- Pressure of vessel =2. 116 bars
- Material of vessel = Carbon Steel
- Plate diameter = 3m
- Plate type = Valve
- Number of trays = 40
- Material of Trays = Stainless Steel

For the cost of pre-flash tower, we use two graphs. From the one graph we find the cost of the vessel. And from the other graph we find the cost of the one plate from which we find the cost of all plates.

Cost of Vessel = Bare Cost*Pressure Factor*Type Factor

= $ 30000 (1)(1)

= $ 30000

Cost of Plates = 40(Bare Cost*Material Factor)

= $ 74800

Total Cost of Stripper = $ 104800

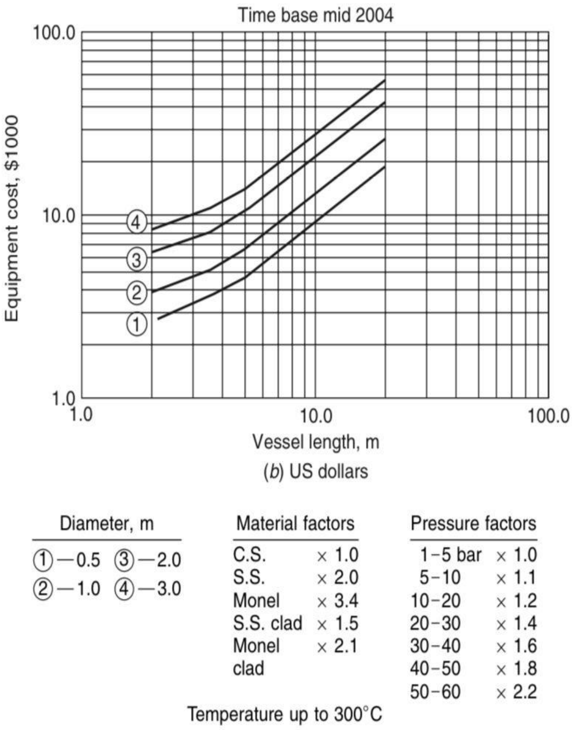


Figure 6.6*a*, *b*. Horizontal pressure vessels. Time base mid-2004.
Purchase cost = (bare cost from figure) × Material factor × Pressure factor

Figure8.4: Vessel cost graph

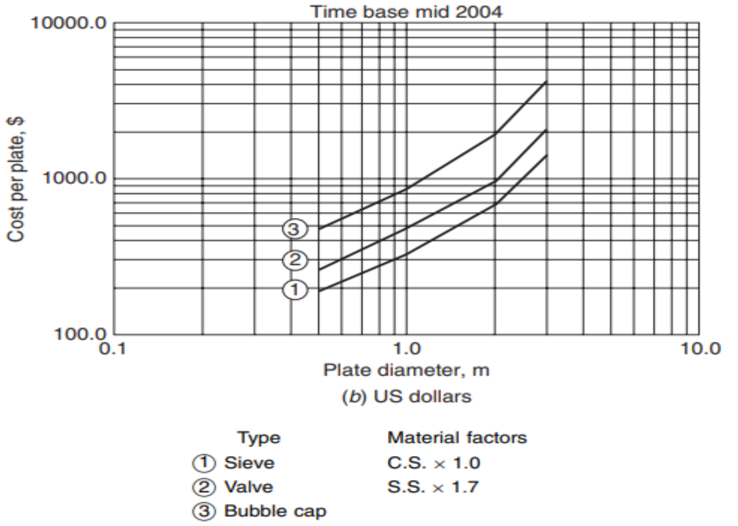


Figure 6.7*a*, *b*. Column plates. Time base mid-2004 (for column costs see Figure 6.4)
Installed cost = (cost from figure) × Material factor

Figure8.5:Plates cost graph

8.1.5 Kero Stripper

- Vessel Diameter = 0.76 m
- Height of vessel = 9.75 m
- Pressure of vessel =1 bar
- Material of vessel = Carbon Steel
- Plate diameter = 0.76 m
- Plate type = Valve
- Number of trays = 3
- Material of Trays = Stainless Steel

For the cost of pre-flash tower, we use two graphs. From the one graph we find the cost of the vessel. And from the other graph we find the cost of the one plate from which we find the cost of all plates.

Cost of Vessel = Bare Cost*Pressure Factor*Type Factor

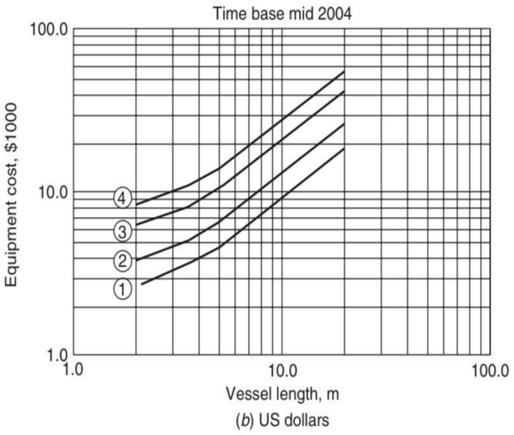
= $ 3500 (1)(1)

= $ 3500

Cost of Plates = 3(Bare Cost*Material Factor)

= $ 1071

Total Cost of Stripper = $ 4571



Diameter, m

①—0.5 ③—2.0
②—1.0 ④—3.0

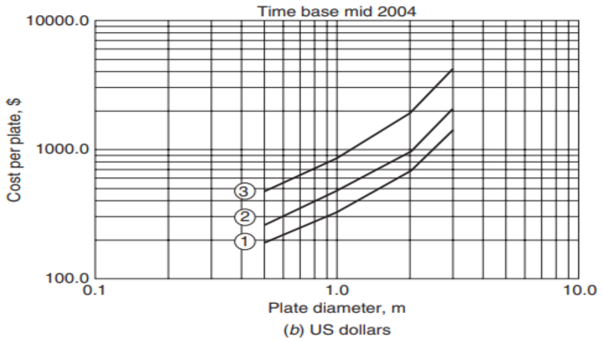
Material factors	
C.S.	× 1.0
S.S.	× 2.0
Monel	× 3.4
S.S. clad	× 1.5
Monel clad	× 2.1

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

Temperature up to 300°C

Figure 6.6*a*, *b*. Horizontal pressure vessels. Time base mid-2004.
Purchase cost = (bare cost from figure) × Material factor × Pressure factor

Figure8.6 :Vessel cost graph



Type	Material factors
① Sieve	C.S. × 1.0
② Valve	S.S. × 1.7
③ Bubble cap	

Figure 6.7*a*, *b*. Column plates. Time base mid-2004 (for column costs see Figure 6.4)
Installed cost = (cost from figure) × Material factor

Figure8.7:Plates cost graph

8.1.6 LDO Stripper

- Vessel Diameter = 0.76 m
- Height of vessel = 4.87 m
- Pressure of vessel =1 bar
- Material of vessel = Carbon Steel
- Plate diameter = 0.76 m
- Plate type = Valve
- Number of trays = 3
- Material of Trays = Stainless Steel

For the cost of pre-flash tower, we use two graphs. From the one graph we find the cost of the vessel. And from the other graph we find the cost of the one plate from which we find the cost of all plates.

Cost of Vessel = Bare Cost*Pressure Factor*Type Factor

= $ 2900 (1)(1)

= $ 2900

Cost of Plates = 3(Bare Cost*Material Factor)

= $ 1071

Total Cost of Stripper = $ 3971

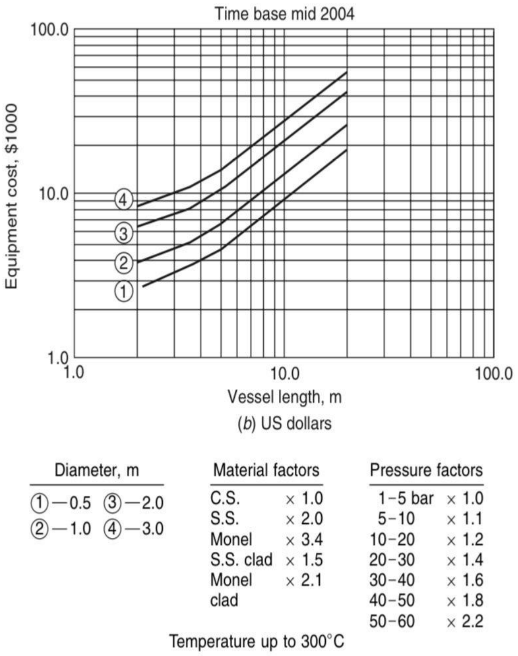


Figure 6.6*a*, *b*. Horizontal pressure vessels. Time base mid-2004.
Purchase cost = (bare cost from figure) × Material factor × Pressure factor

Figure 8.8: Vessel cost graph

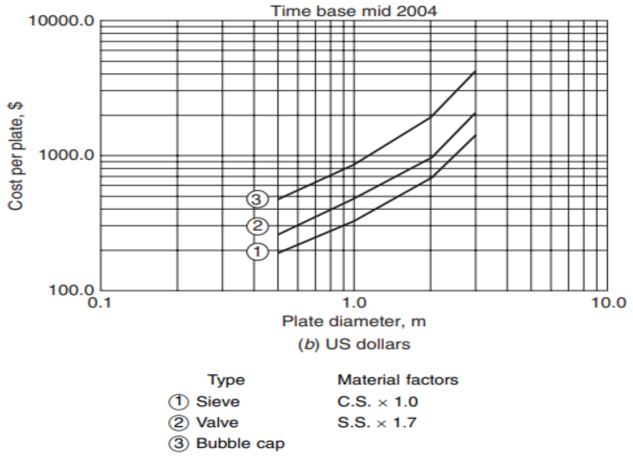


Figure 6.7*a*, *b*. Column plates. Time base mid-2004 (for column costs see Figure 6.4)
Installed cost = (cost from figure) × Material factor

Figure8.9: Plates cost graph

8.1.7 HDO Stripper

- Vessel Diameter = 0.76 m
- Height of vessel = 9.75 m
- Pressure of vessel =1 bar

- Material of vessel = Carbon Steel
- Plate diameter = 0.76 m
- Plate type = Valve
- Number of trays = 3
- Material of Trays = Stainless Steel

For the cost of pre-flash tower, we use two graphs. From the one graph we find the cost of the vessel. And from the other graph we find the cost of the one plate from which we find the cost of all plates.

Cost of Vessel = Bare Cost*Pressure Factor*Type Factor

= $ 3500 (1)(1)

= $ 3500

Cost of Plates = 3(Bare Cost*Material Factor)

= $ 1071

Total Cost of Stripper = $ 4571

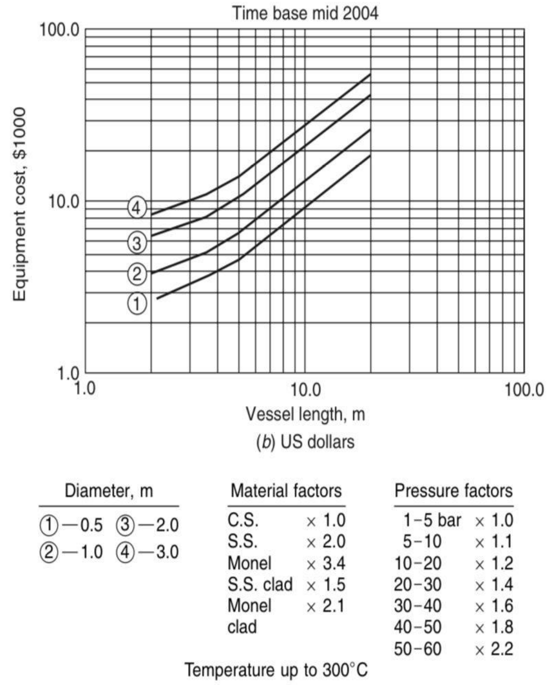


Figure 6.6*a*, *b*. Horizontal pressure vessels. Time base mid-2004.
Purchase cost = (bare cost from figure) × Material factor × Pressure factor

Figure 8.10:Vessel cost graph.

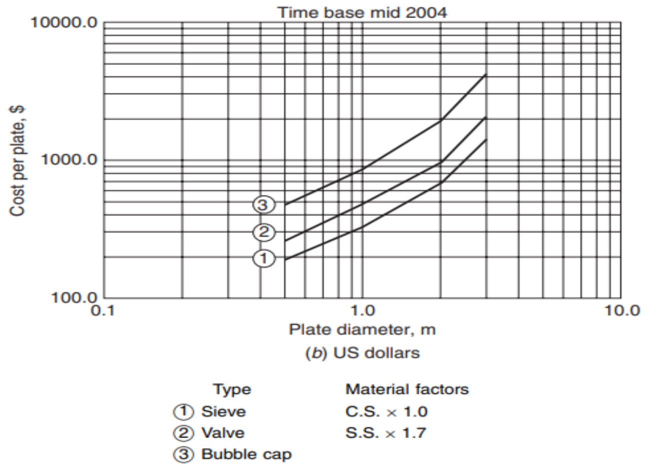


Figure 6.7*a*, *b*. Column plates. Time base mid-2004 (for column costs see Figure 6.4) Installed cost = (cost from figure) × Material factor

Figure8.11:Plates cost graph

8.1.8 Pre-Flash Furnace

- Heat Absorption = 12700 KW
- Index = 0.77
- Material = Carbon Steel

From the chart by using the heat absorption value we find the index which we put in the formula to find the cost of the furnace.

Purchase cost of Furnace(C_e)= C S^n

$$= 330\ (12700)^{0.77}$$

$$= \$\ 4676917$$

Where, Ce = Purchase cost of equipment

C = Cost constant

S = Characteristics size parameter

n = Index

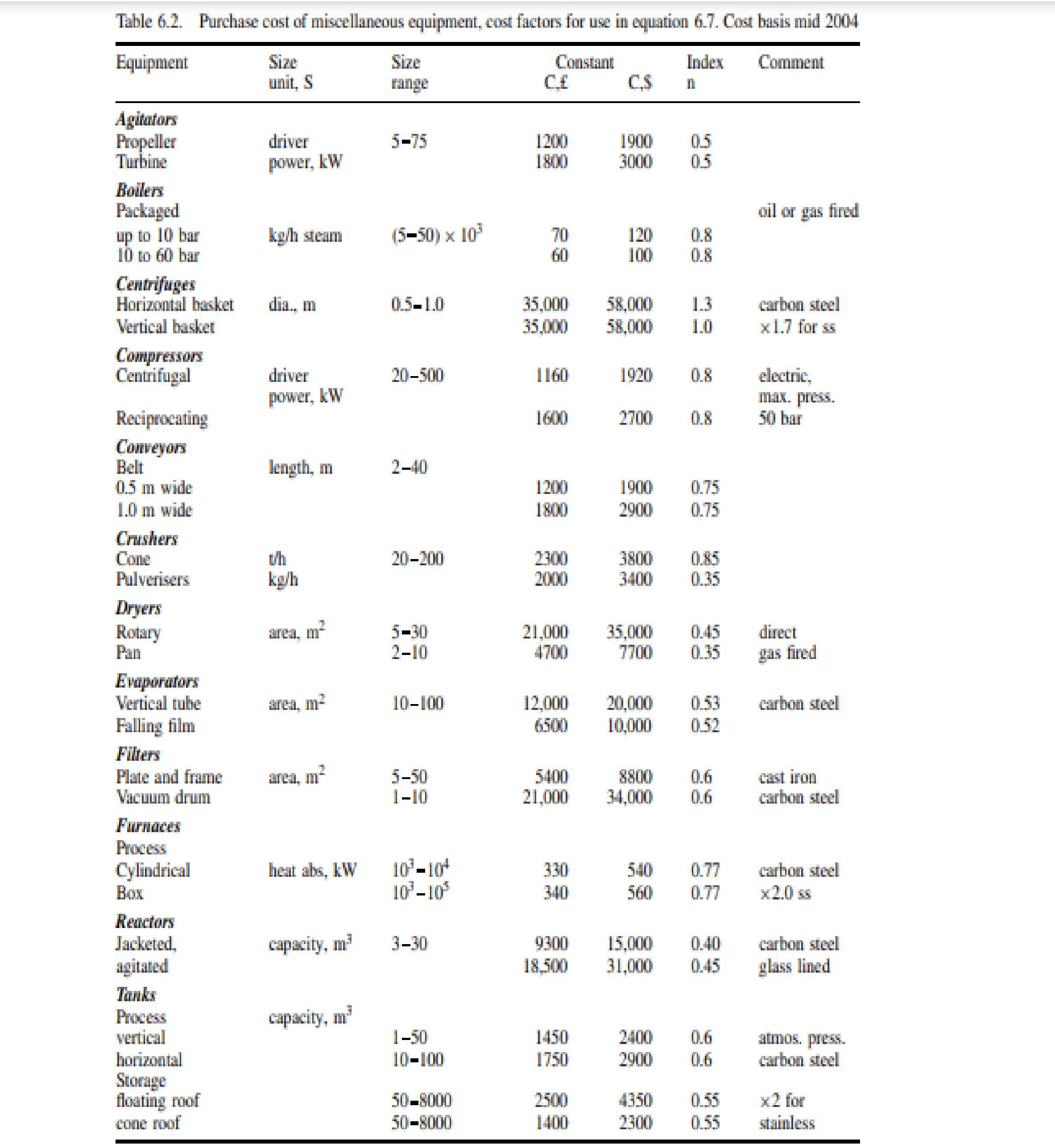


Table 6.2. Purchase cost of miscellaneous equipment, cost factors for use in equation 6.7. Cost basis mid 2004

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 8. 12:Cost of furnace

8.1.9 Atmospheric Furnace

- Heat Absorption = 13900 KW
- Index = 0.77

- Material = Carbon Steel

From the chart by using the heat absorption value we find the index which we put in the formula to find the cost of the furnace.

Purchase cost of Furnace(C_e)= C S^n

= 330 $(13900)^{0.77}$

= $ 511252

Where, Ce = Purchase cost of equipment

C = Cost constant

S = Characteristics size parameter

n = Index

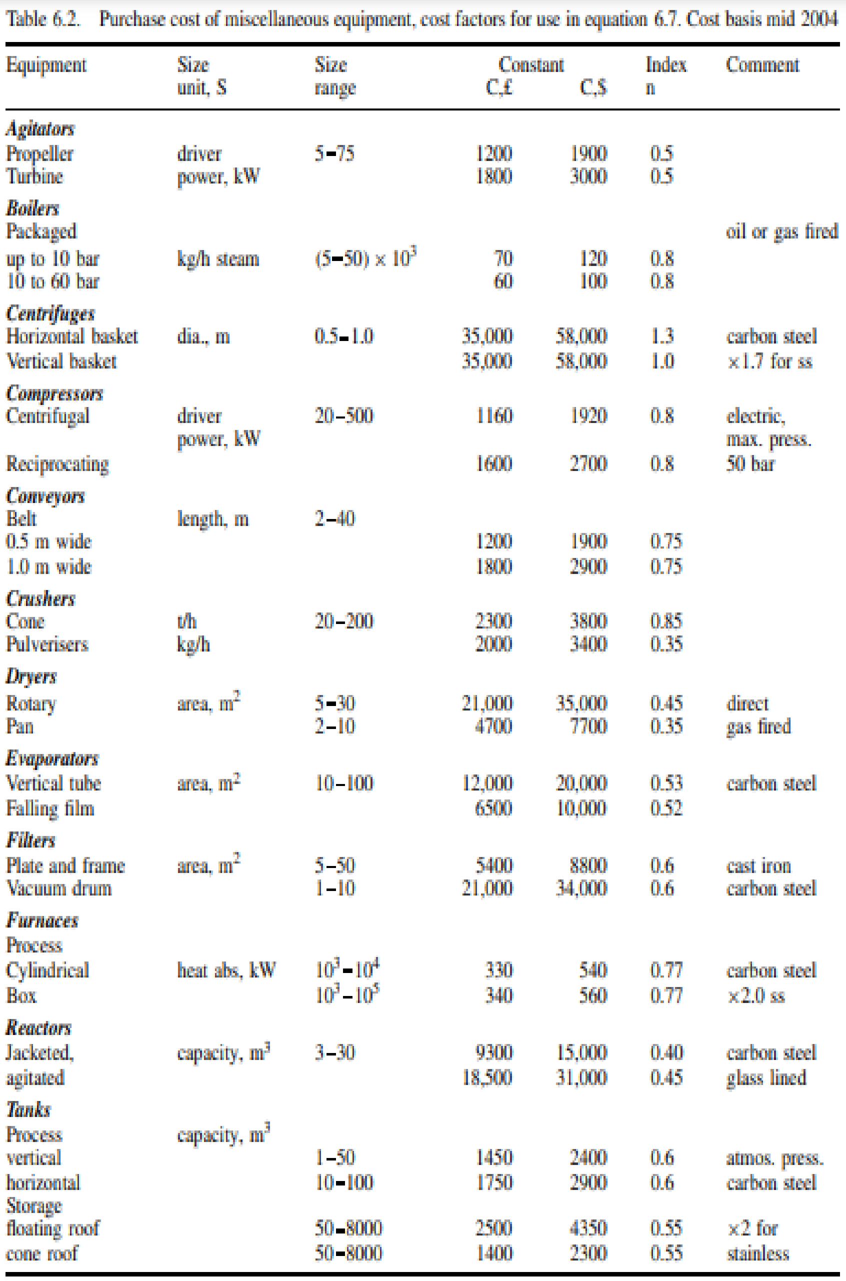


Table 6.2. Purchase cost of miscellaneous equipment, cost factors for use in equation 6.7. Cost basis mid 2004

Equipment	Size unit, S	Size range	Constant		Index	Comment
			C,£	C,$	n	
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 8.13:Cost of furnace

8.1.10Pre-Flash Overhead

- Operating pressure = 0.68 bar
- Volume = 6.27 m^3
- Material = Carbon Steel

From the chart by using the Volume we find the index which we put in the formula to find the cost of the overhead.

Purchase cost of Overhead (C_e)= C S^n

= 1750 $(6.27)^{0.6}$

= $ 5265

Where, Ce = Purchase cost of equipment

C = Cost constant

S = Characteristics size parameter

n = Index

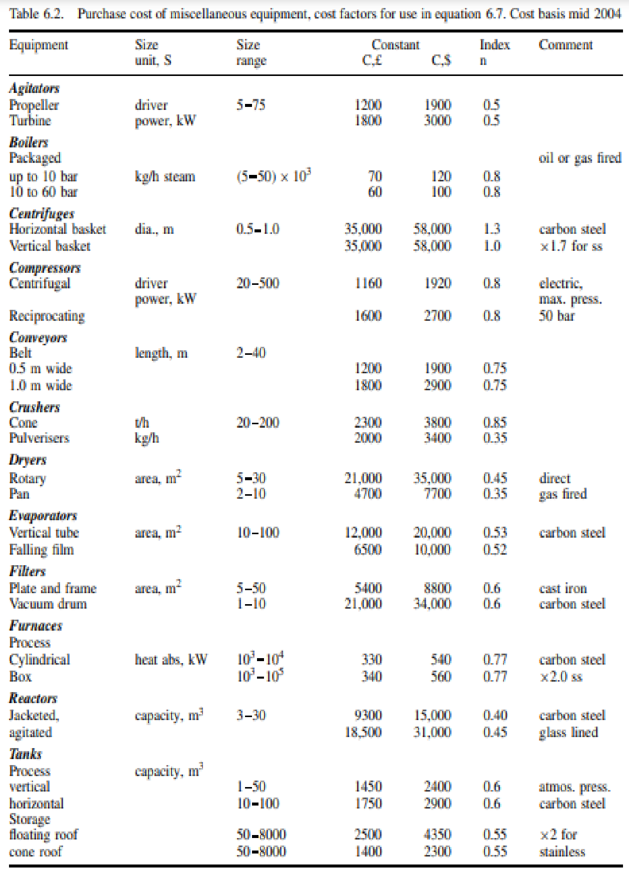


Table 6.2. Purchase cost of miscellaneous equipment, cost factors for use in equation 6.7. Cost basis mid 2004

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

Figure8.14:Cost of OverHead

8.1.11 Atmospheric Overhead

- Operating pressure = 10.68 bar
- Volume = 0.48 m^3
- Material = Carbon Steel

From the chart by using the Volume we find the index which we put in the formula to find the cost of the overhead.

Purchase cost of Overhead (C_e)= C S^n

= 1750 $(0.48)^{0.6}$

= $ 1126

Where, Ce = Purchase cost of equipment

C = Cost constant

S = Characteristics size parameter

n = Index

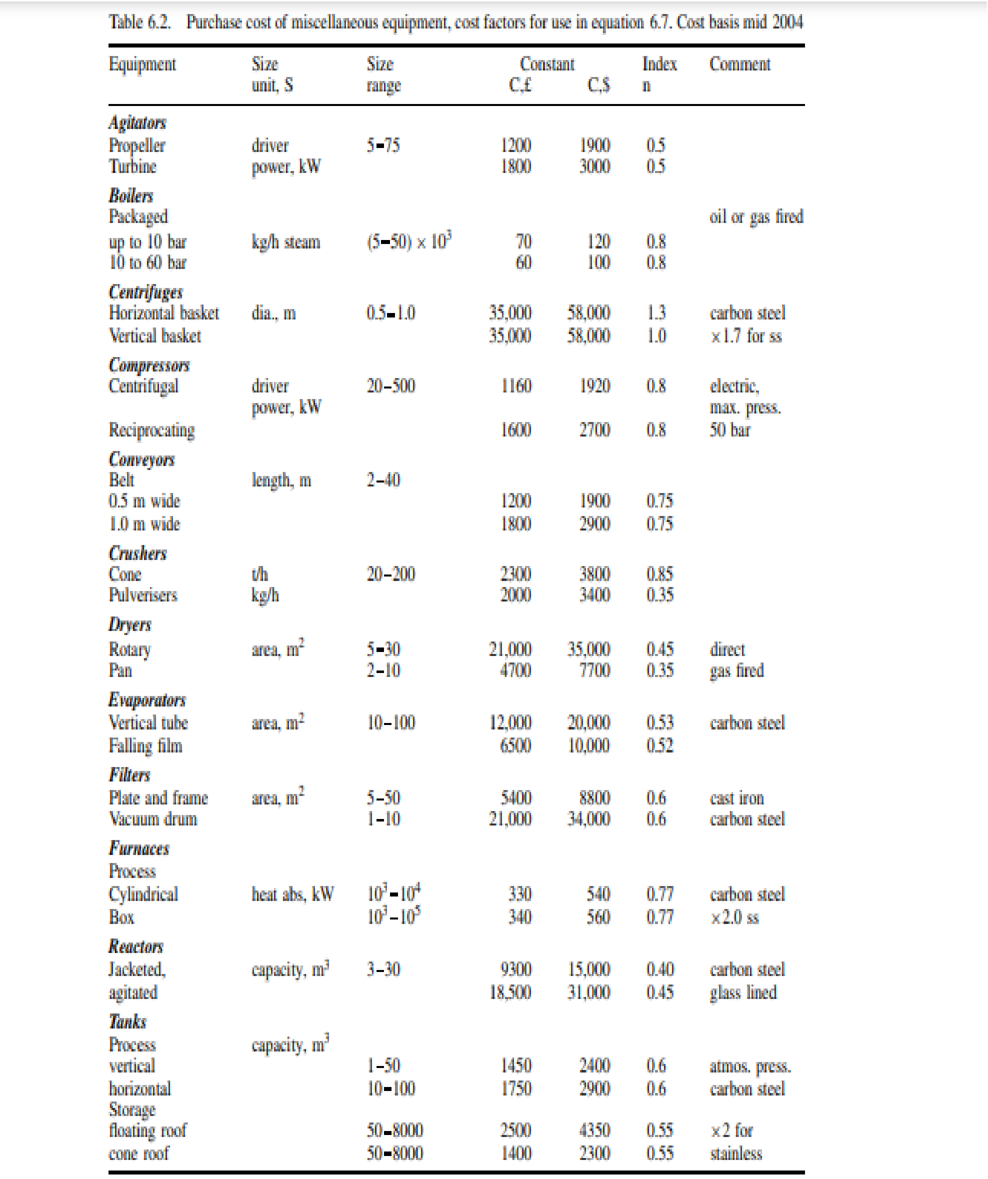


Table 6.2. Purchase cost of miscellaneous equipment, cost factors for use in equation 6.7. Cost basis mid 2004

Equipment	Size unit, S	Size range	Constant		Index n	Comment
			C,£	C,$		
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

Figure8.15 :Cost of OverHead

8.1.12 Purchased Cost of Equipment

Cost of 8 Heaters = $192000

Cost of Desalter =$ 6798

Cost of Pre-Flash Furnace =$ 476916

Cost of Atmospheric Furnace = $ 511252

Cost of Pre-Flash Tower =$ 28760

Cost of Atmospheric Tower =$ 104800

Cost of LDO Stripper =$ 4571

Cost of HDO Stripper =$ 3971

Cost of Kero Stripper =$ 4571

Cost of Pre-Flash OVHD=$ 5265

Cost of Atmospheric OVHD=$ 1126

Total Purchased cost of Equipment (PCE) = $1340031

8.2 Fixed capital

Total plant physical cost = 1340031 (0.4+0.7+0.2+0.1) = $5896136

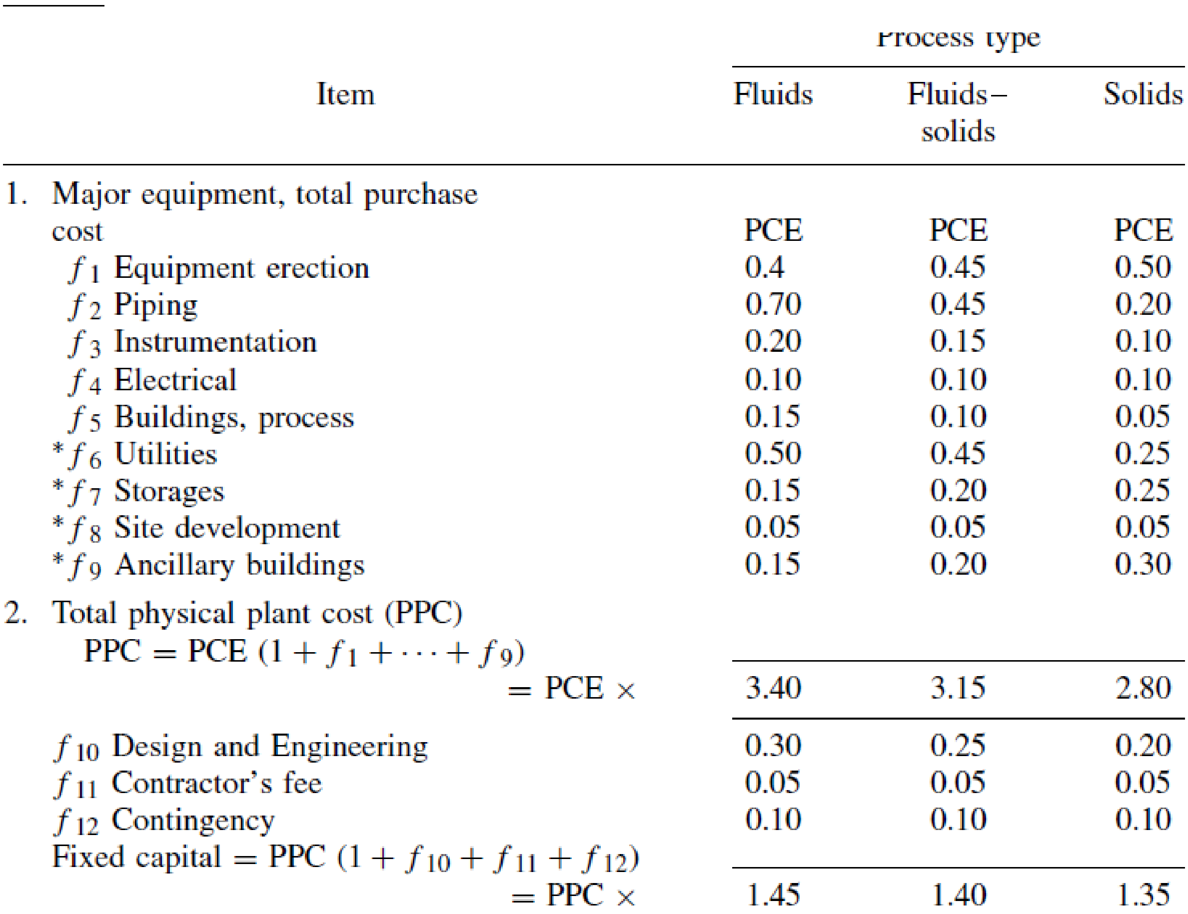
Fixed capital cost = 5896136 (0.3+0.05+0.1) = $8549397

Working capital cost = 8549397*0.05 = $427469

Total cost of raw materials = $444800000

Total capital investment = 8549397 + 427469 + 444800000 = $453776866

Table 8.0-1: Factors for estimation of project fixed capital 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

Annual operating cost

Plant attainment = 0.95

Operating time = 0.95*365 = 347 days/year

8.3 Variable cost

1. Raw materials

Annual Cost of Crude oil according to 32000 barrel/day crude oil = $ 444800000

1. Miscellaneous

   Cost = $59845

2. Utilities

   Cost = $12582.9

Total variable cost =$ 444872727.9

8.4 Fixed operating cost

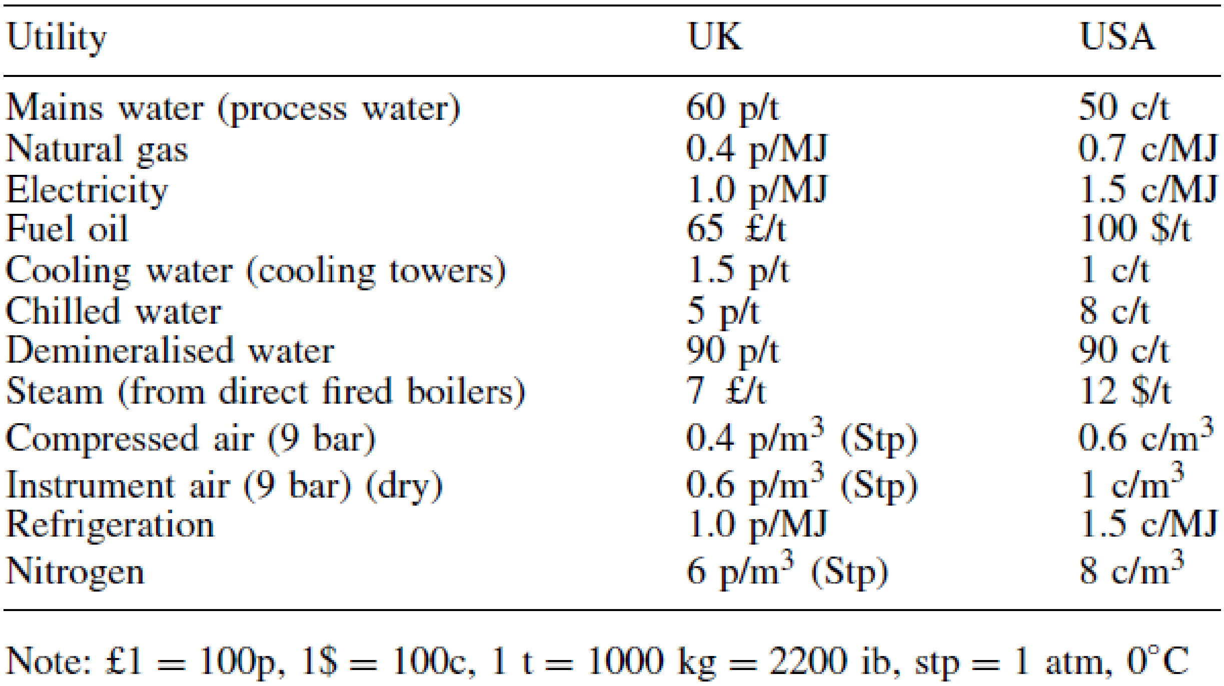
Total cost after applying the factors listed in the below table = $ 1867569

Annual operating cost = $ 558384999

Cost Inflation: Cost in year 2022 = Cost in year 2004 $(\frac{CPE\ index\ 2022}{CPE\ index\ 2004})$

= $ 558384999$(\frac{317}{109})^{*}$

Table 8. 0-2:Utility cost.

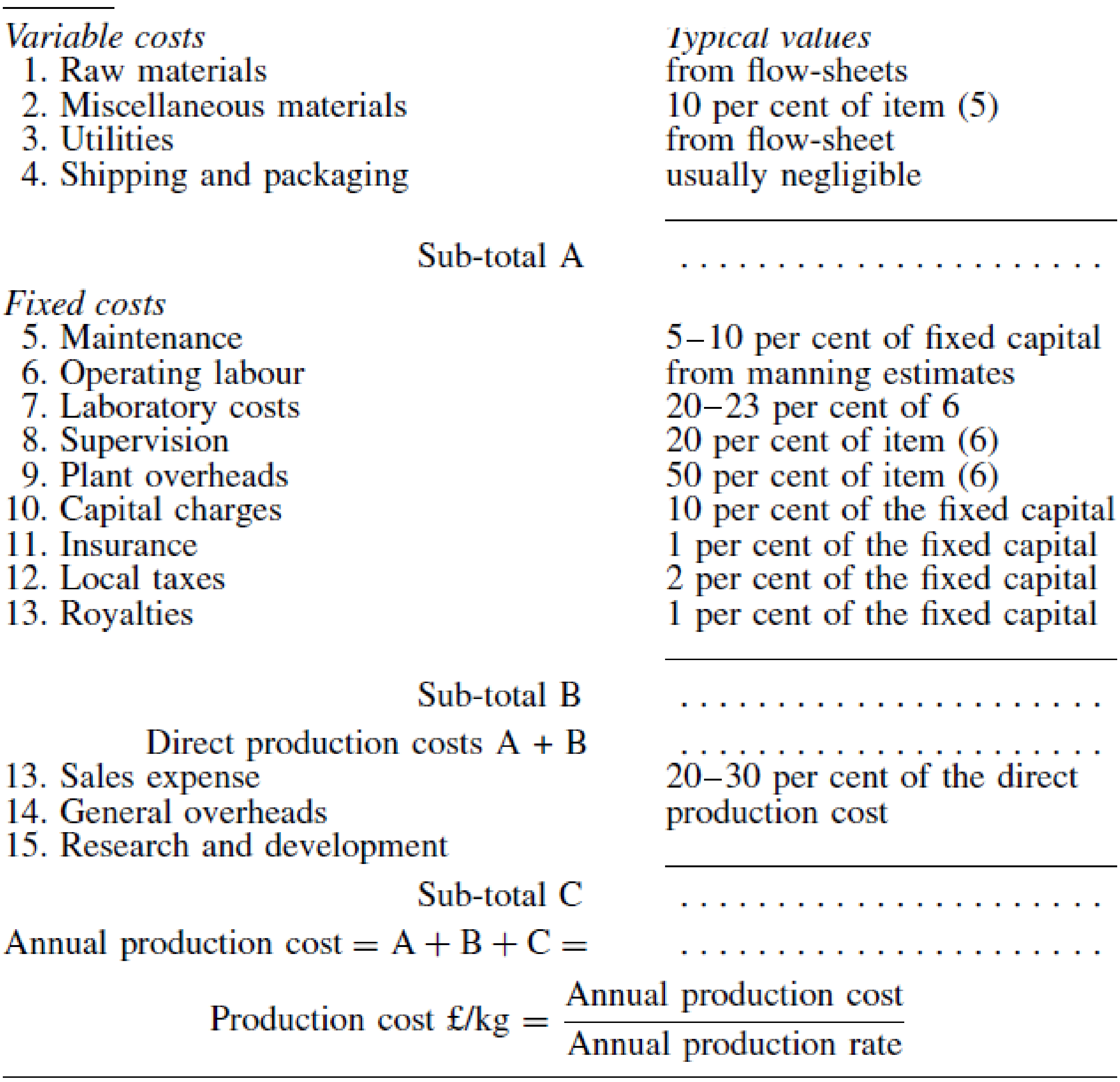


Utility	UK	USA
Mains water (process water)	60 p/t	50 c/t
Natural gas	0.4 p/MJ	0.7 c/MJ
Electricity	1.0 p/MJ	1.5 c/MJ
Fuel oil	65 £/t	100 \$/t
Cooling water (cooling towers)	1.5 p/t	1 c/t
Chilled water	5 p/t	8 c/t
Demineralised water	90 p/t	90 c/t
Steam (from direct fired boilers)	7 £/t	12 \$/t
Compressed air (9 bar)	0.4 p/m^3 (Stp)	0.6 c/m^3
Instrument air (9 bar) (dry)	0.6 p/m^3 (Stp)	1 c/m^3
Refrigeration	1.0 p/MJ	1.5 c/MJ
Nitrogen	6 p/m^3 (Stp)	8 c/m^3

Note: £1 = 100p, 1$ = 100c, 1 t = 1000 kg = 2200 ib, stp = 1 atm, 0°C

= $ 1623927015

Table 8.0-3:Factors for estimation of annual operating cost.



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 =	. . . .

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

8.5 Rate of return

Total sale price = $ 1140589068

Total investment = $453776866

Payback Period= $\frac{Total\ investment}{Profit}$

Payback period = 2.1 years

8.6 Summary for Economic Analysis

Table 8. 4 :Summary for economic analysis.

Fixed capital	$8549397
Total capital investment	$453776866
Annual operating cost	$558384999
Annual revenue	$1140589068
Payback Period	2.1 years

CHAPTER 9

CONTROL LOOPS

9.1 Control Loops

Control loops play a crucial role in the functioning of process simulation software like Aspen HYSYS. These loops are responsible for regulating and maintaining optimal process conditions, ensuring efficient and safe operations. Aspen HYSYS provides a comprehensive set of tools and features to design and implement control loops effectively.

At the heart of control loops in Aspen HYSYS lies the concept of a process variable, which represents a parameter that needs to be controlled, such as temperature, pressure, or flow rate. The software allows users to define these variables and set target values. Control loops continuously monitor the process variable and make adjustments to bring it closer to the desired setpoint.

Aspen HYSYS offers various types of control algorithms, including proportional-integral-derivative (PID) controllers, to implement control strategies. These algorithms use mathematical calculations to compute the required adjustments and generate control signals to actuate valves, pumps, or other process equipment.

The software also enables users to visualize control loops through graphical representations, such as loop diagrams and trend plots. This helps in understanding the behavior of the process variable and the effectiveness of the control loop.

Moreover, Aspen HYSYS allows for tuning control loops by adjusting parameters like gain, integral time, and derivative time. This tuning process ensures optimal performance and stability of the control system.

9.1.1 Temperature Control Loops

1. Control loops employed to keep process variables within specified range.

2. Sensor picks up changes in parameter, sends signal to controller.

3. Parameter compared with set point; controller acts accordingly.

4. A Temperature Control Loop has been installed to maintain Stream entering Pre-Flash at 150°C.

5. Steam flow rate into heat exchanger varied to acquire desired output temperature.

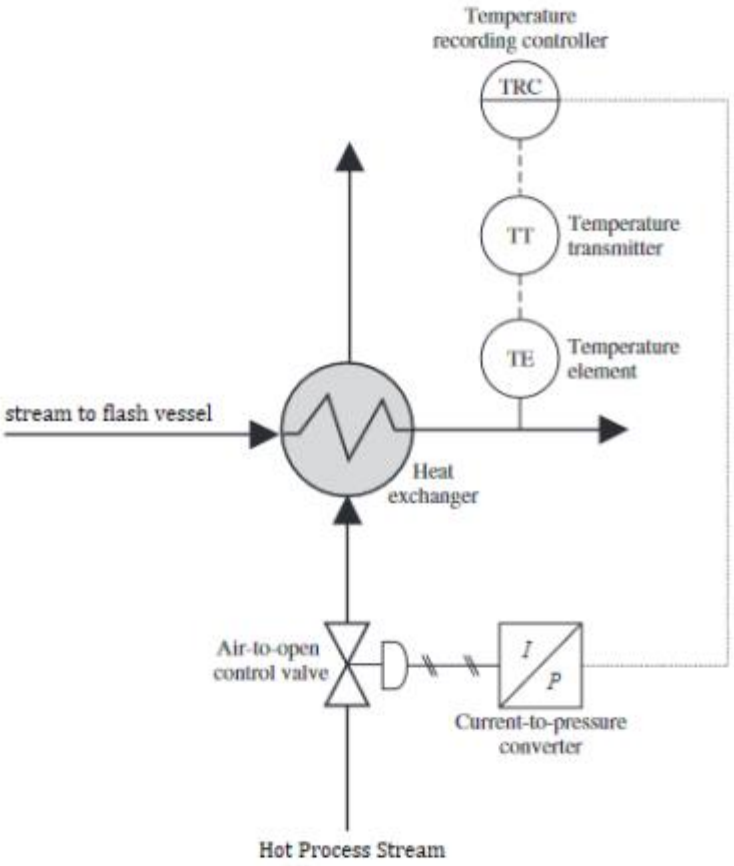


Figure 9.1: Temperature Control Loop

9.1.2 Flow Control Loop

1. A ratio control loop is employed to maintain desired ratio of water flow rate to crude oil feed flow rate.

2. Set point given to the controller.

3. The flow rates are measured via flow sensors.

4. Desired ratio achieved by varying flow rate of water.

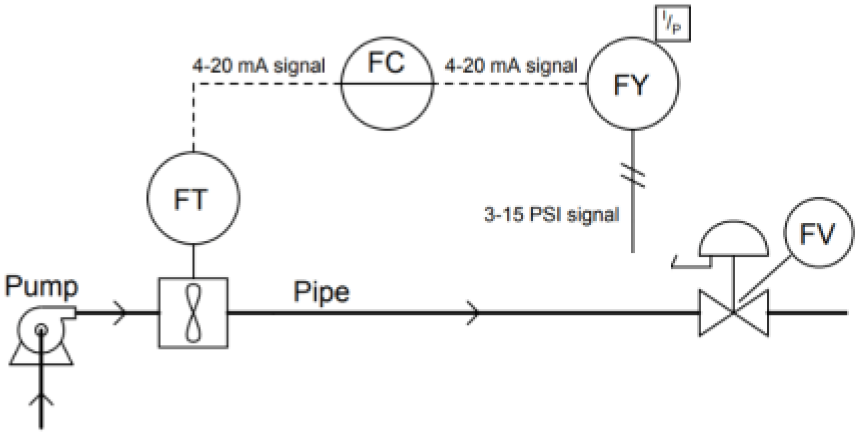


Figure 9.2: Flow Control Loop

9.1.3 Pressure Control Loop

1. A pressure control loop is installed to monitor and control the exit temperature of compressor.

2. Control achieved by detecting deviation in pressure and then varying compressor energy input accordingly.

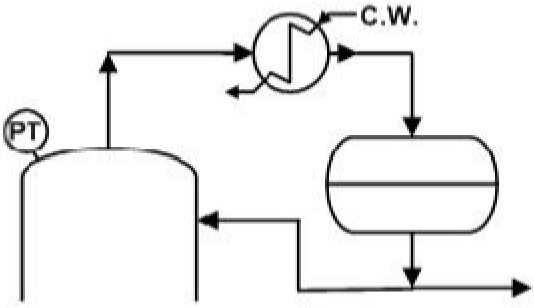


Figure 9.3: Pressure Control Loop

9.1.4 Level Control Loop

1. Level control loops employed for both columns to maintain the required liquid holdup.

2. Level sensor picks up the deviations, the controller output then varies outlet valve opening.

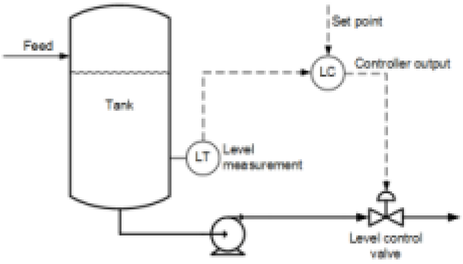


Figure 9.4: Level Control Loop

CHAPTER 10

HAZOP ANALYSIS

10.1 Hazop analysis:

Conducting a Hazard and Operability (HAZOP) study is crucial for assessing the technical feasibility and safety aspects of a proposed project. Traditionally, this type of analysis involves employing specific guidewords to evaluate potential causes, effects, and preventive measures related to the process. Guidewords, such as "high temperature" or "low pressure," are specific to the equipment being studied within the system. Each component of the system must be thoroughly evaluated to ensure a comprehensive risk assessment. The HAZOP process involves several key steps:

1. Brainstorming and identifying potential deviations and guidewords applicable to the system.

2. Conducting a qualitative risk assessment to determine the severity of each deviation.

3. Employing a "bottom-up" approach that relies on the experience and predictive abilities of the observer.

In the HAZOP study, various terms and concepts are utilized:

1. Nodes: These refer to pipe sections or vessels where process chemicals are present.
2. Super nodes: Multiple nodes considered together as one for more efficient analysis.
3. Hazard: A potential source of harm.
4. Harm: Actual injury or damage to equipment, personnel, or the environment.
5. Risk: The probability of harm occurring and the associated level of damage.

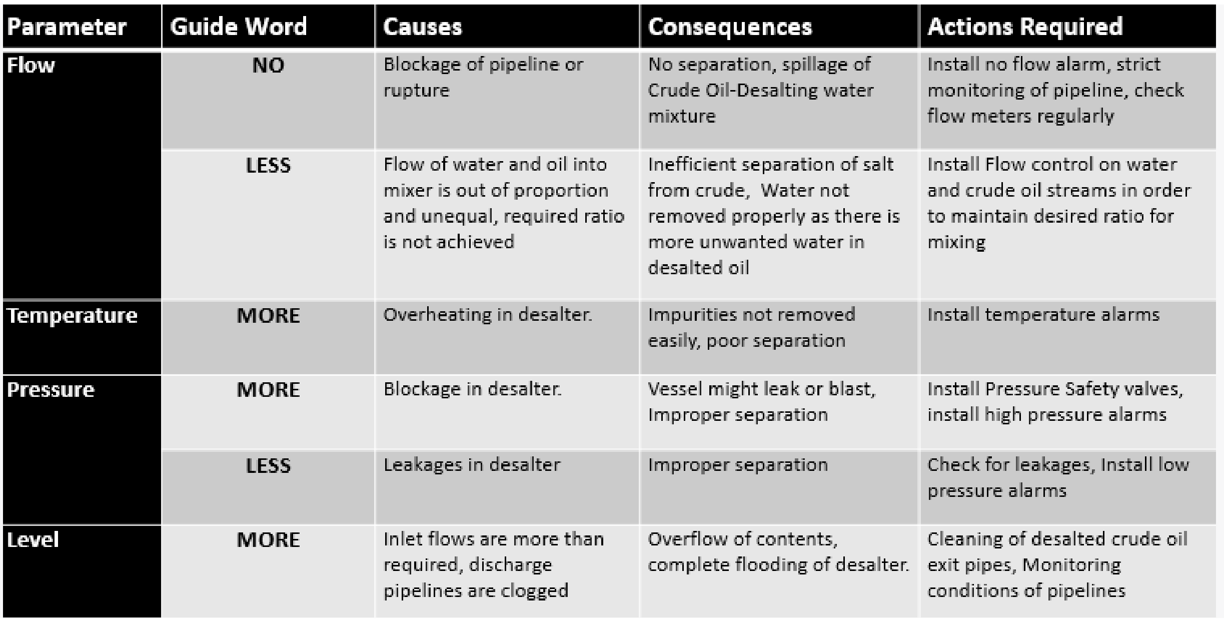
The methodology for conducting a HAZOP study involves the following steps:

1. Identify all major nodes and super nodes within the system.
2. Justify why each node has been selected for the HAZOP study.
3. Choose a key process parameter for each node.
4. Apply relevant guidewords to the chosen parameter, identifying potential deviations in the process and documenting them.
5. Determine the potential causes for the identified deviations and add them to the analysis table.
6. For each brainstormed deviation, identify the consequences it would lead to and document them.
7. Repeat steps 3 through 6 for the remaining nodes until all chosen nodes have been thoroughly studied.

Following this systematic methodology ensures a comprehensive evaluation of the system, enabling the identification of potential hazards and the implementation of appropriate preventive measures

10.1.1 Desalter:

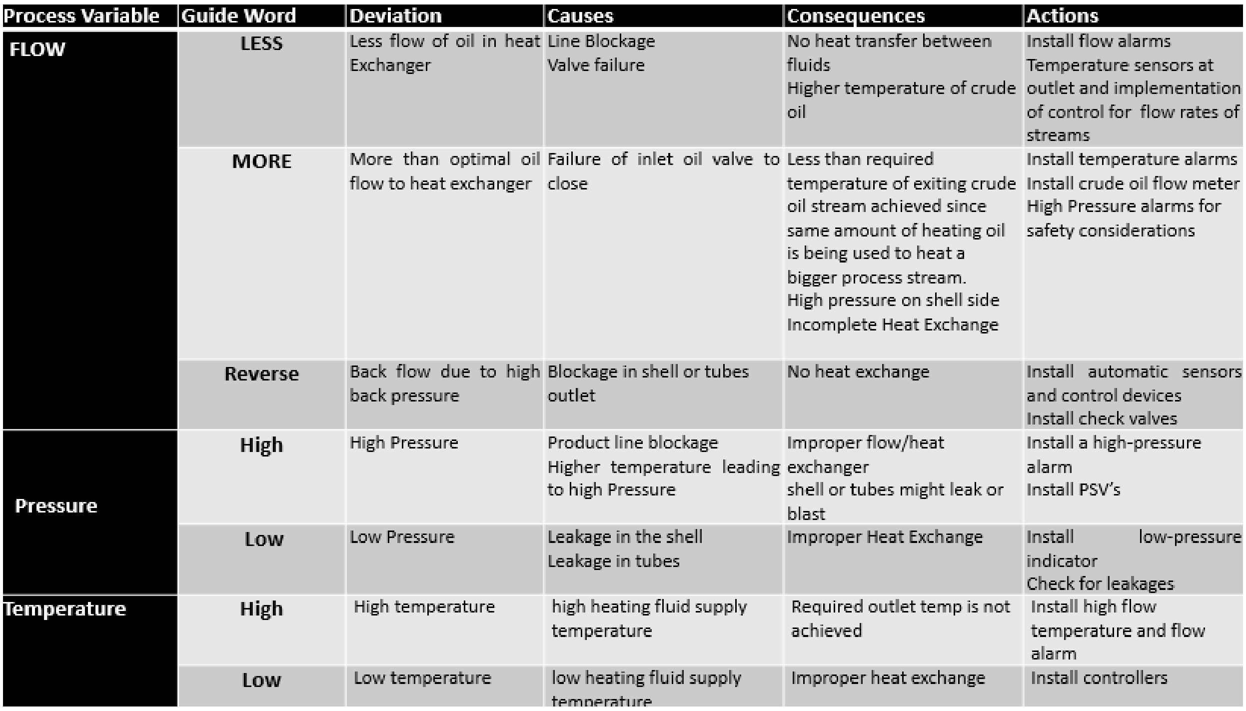
Table 10.1: Hazop for desalter



Parameter	Guide Word	Causes	Consequences	Actions Required
Flow	**NO**	Blockage of pipeline or rupture	No separation, spillage of Crude Oil-Desalting water mixture	Install no flow alarm, strict monitoring of pipeline, check flow meters regularly
	LESS	Flow of water and oil into mixer is out of proportion and unequal, required ratio is not achieved	Inefficient separation of salt from crude, Water not removed properly as there is more unwanted water in desalted oil	Install Flow control on water and crude oil streams in order to maintain desired ratio for mixing
Temperature	**MORE**	Overheating in desalter.	Impurities not removed easily, poor separation	Install temperature alarms
Pressure	**MORE**	Blockage in desalter.	Vessel might leak or blast, Improper separation	Install Pressure Safety valves, install high pressure alarms
	LESS	Leakages in desalter	Improper separation	Check for leakages, Install low pressure alarms
Level	**MORE**	Inlet flows are more than required, discharge pipelines are clogged	Overflow of contents, complete flooding of desalter.	Cleaning of desalted crude oil exit pipes, Monitoring conditions of pipelines

10.1.2 Heat Exchanger:

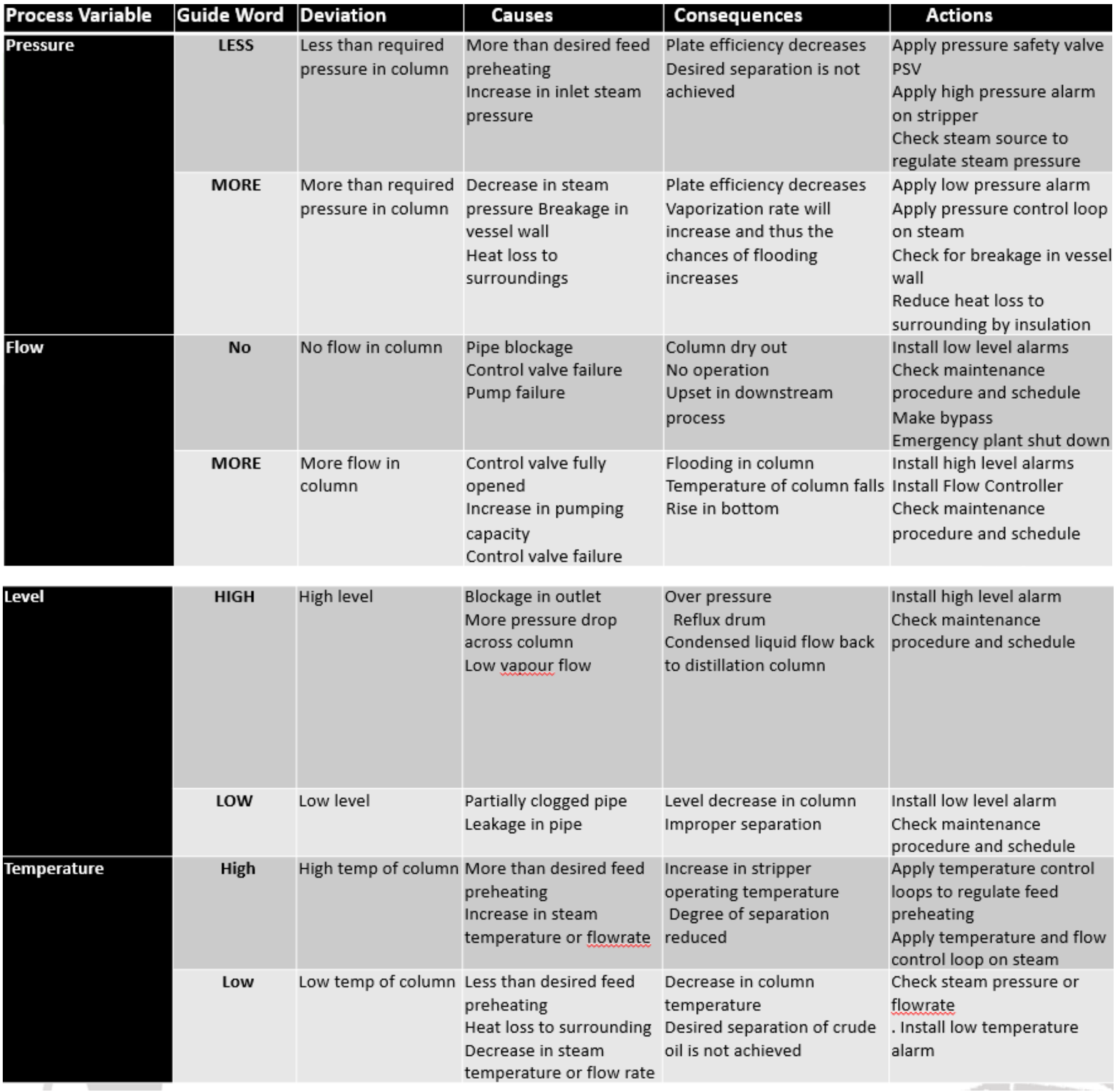
Table 10.2: Hazop for Heat exchanger.



Process Variable	Guide Word	Deviation	Causes	Consequences	Actions
FLOW	**LESS**	Less flow of oil in heat Exchanger	Line Blockage Valve failure	No heat transfer between fluids Higher temperature of crude oil	Install flow alarms Temperature sensors at outlet and implementation of control for flow rates of streams
	MORE	More than optimal oil flow to heat exchanger	Failure of inlet oil valve to close	Less than required temperature of exiting crude oil stream achieved since same amount of heating oil is being used to heat a bigger process stream. High pressure on shell side Incomplete Heat Exchange	Install temperature alarms Install crude oil flow meter High Pressure alarms for safety considerations
	Reverse	Back flow due to high back pressure	Blockage in shell or tubes outlet	No heat exchange	Install automatic sensors and control devices Install check valves
Pressure	**High**	High Pressure	Product line blockage Higher temperature leading to high Pressure	Improper flow/heat exchanger shell or tubes might leak or blast	Install a high-pressure alarm Install PSV's
	Low	Low Pressure	Leakage in the shell Leakage in tubes	Improper Heat Exchange	Install low-pressure indicator Check for leakages
Temperature	**High**	High temperature	high heating fluid supply temperature	Required outlet temp is not achieved	Install high flow temperature and flow alarm
	Low	Low temperature	low heating fluid supply temperature	Improper heat exchange	Install controllers

10.1.3 Distillation Column:

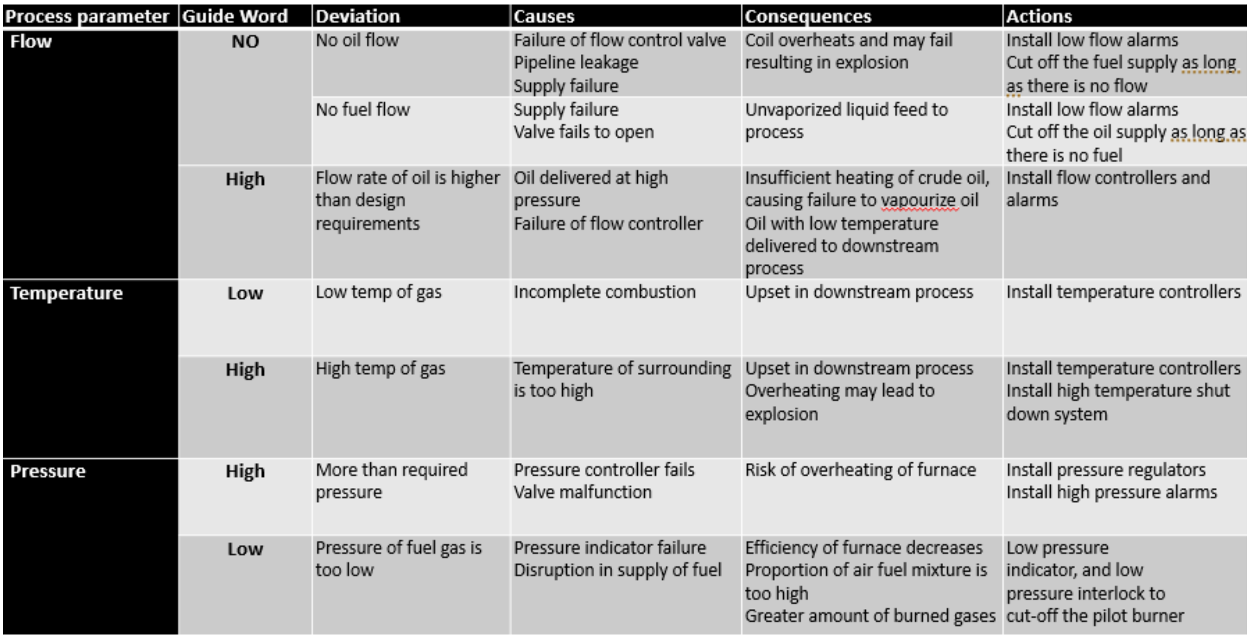
Table 10.3: Hazop for distillation column



Process Variable	Guide Word	Deviation	Causes	Consequences	Actions
Pressure	**LESS**	Less than required pressure in column	More than desired feed preheating Increase in inlet steam pressure	Plate efficiency decreases Desired separation is not achieved	Apply pressure safety valve PSV Apply high pressure alarm on stripper Check steam source to regulate steam pressure
	MORE	More than required pressure in column	Decrease in steam pressure Breakage in vessel wall Heat loss to surroundings	Plate efficiency decreases Vaporization rate will increase and thus the chances of flooding increases	Apply low pressure alarm Apply pressure control loop on steam Check for breakage in vessel wall Reduce heat loss to surrounding by insulation
Flow	**No**	No flow in column	Pipe blockage Control valve failure Pump failure	Column dry out No operation Upset in downstream process	Install low level alarms Check maintenance procedure and schedule Make bypass Emergency plant shut down
	MORE	More flow in column	Control valve fully opened Increase in pumping capacity Control valve failure	Flooding in column Temperature of column falls Rise in bottom	Install high level alarms Install Flow Controller Check maintenance procedure and schedule
Level	**HIGH**	High level	Blockage in outlet More pressure drop across column Low vapour flow	Over pressure Reflux drum Condensed liquid flow back to distillation column	Install high level alarm Check maintenance procedure and schedule
	LOW	Low level	Partially clogged pipe Leakage in pipe	Level decrease in column Improper separation	Install low level alarm Check maintenance procedure and schedule
Temperature	**High**	High temp of column	More than desired feed preheating Increase in steam temperature or flowrate	Increase in stripper operating temperature Degree of separation reduced	Apply temperature control loops to regulate feed preheating Apply temperature and flow control loop on steam
	Low	Low temp of column	Less than desired feed preheating Heat loss to surrounding Decrease in steam temperature or flow rate	Decrease in column temperature Desired separation of crude oil is not achieved	Check steam pressure or flowrate . Install low temperature alarm

10.1.4 Furnace:

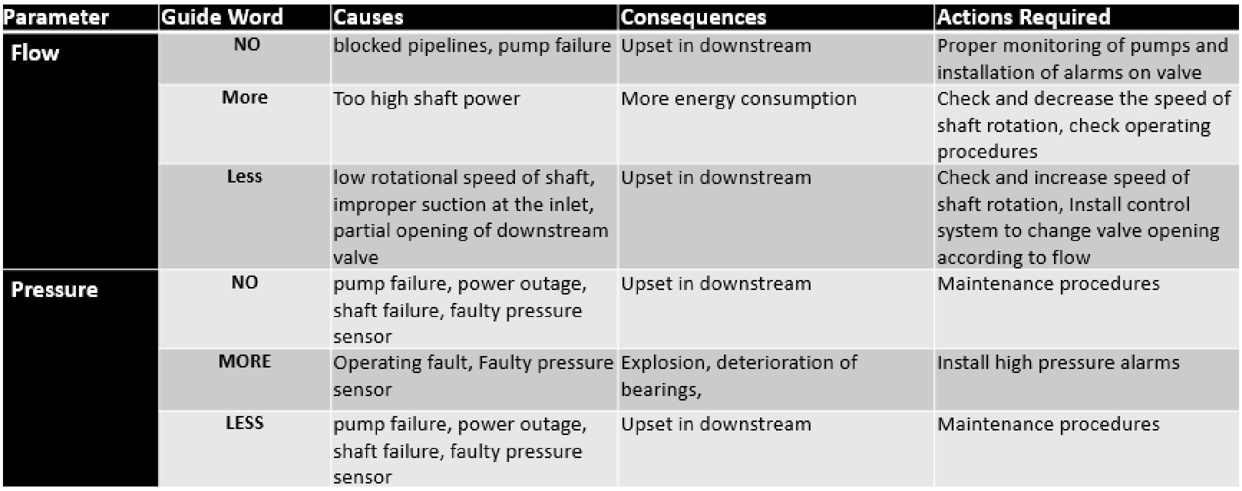
Table 10.4: Hazop for furnace



Process parameter	Guide Word	Deviation	Causes	Consequences	Actions
Flow	**NO**	No oil flow	Failure of flow control valve Pipeline leakage Supply failure	Coil overheats and may fail resulting in explosion	Install low flow alarms Cut off the fuel supply as long as there is no flow
		No fuel flow	Supply failure Valve fails to open	Unvaporized liquid feed to process	Install low flow alarms Cut off the oil supply as long as there is no fuel
	High	Flow rate of oil is higher than design requirements	Oil delivered at high pressure Failure of flow controller	Insufficient heating of crude oil, causing failure to vapourize oil Oil with low temperature delivered to downstream process	Install flow controllers and alarms
Temperature	**Low**	Low temp of gas	Incomplete combustion	Upset in downstream process	Install temperature controllers
	High	High temp of gas	Temperature of surrounding is too high	Upset in downstream process Overheating may lead to explosion	Install temperature controllers Install high temperature shut down system
Pressure	**High**	More than required pressure	Pressure controller fails Valve malfunction	Risk of overheating of furnace	Install pressure regulators Install high pressure alarms
	Low	Pressure of fuel gas is too low	Pressure indicator failure Disruption in supply of fuel	Efficiency of furnace decreases Proportion of air fuel mixture is too high Greater amount of burned gases	Low pressure indicator, and low pressure interlock to cut-off the pilot burner

10.1.5 Pump:

Table 10.5: Hazop for desalter



Parameter	Guide Word	Causes	Consequences	Actions Required
Flow	NO	blocked pipelines, pump failure	Upset in downstream	Proper monitoring of pumps and installation of alarms on valve
	More	Too high shaft power	More energy consumption	Check and decrease the speed of shaft rotation, check operating procedures
	Less	low rotational speed of shaft, improper suction at the inlet, partial opening of downstream valve	Upset in downstream	Check and increase speed of shaft rotation, Install control system to change valve opening according to flow
Pressure	NO	pump failure, power outage, shaft failure, faulty pressure sensor	Upset in downstream	Maintenance procedures
	MORE	Operating fault, Faulty pressure sensor	Explosion, deterioration of bearings,	Install high pressure alarms
	LESS	pump failure, power outage, shaft failure, faulty pressure sensor	Upset in downstream	Maintenance procedures

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