Design of 10,000 Barrels per day Crude Distillation Unit for processing Light Sweet Crude Blend.

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Certificate

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Dedication

This Work is dedicated to our Parents and Teachers, who always had our back when we needed them the most.

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First of all, we would like to express our utmost gratitude to Almighty Allah. Without Allah's mercy, we would never been able to complete our project.

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ABSTRACT

Crude distillation is the most renowned unit operation, and for good reason. The crude is fractionated on the basis of associated volatilities and the mass transfer phenomena occurring at each and every stage of a distillation column. The rising vapors are enriched with more volatile compounds that leave towards the top of the distillation column. The heavier, less volatile compounds leave towards the bottom of the column. However, before crude oil is subjected to this separation, it must undergo pre-treatment processes to remove the impurities it contains at the time of extraction from the well. Sulphur is one of the key impurities that must be removed since it poses a grave threat to the environment. If not removed properly, Sulphur oxidizes during the thermally driven processes forming oxides. These oxides have their own associated health risks and they may combine with rain to form 'acid rain'. Crude oil with a sulphur content of below 1-2% is known as sweet crude. Sweet crude oil is highly sought after due to the less thorough pre-treatment required.

Light, sweet crude oil is crude oil with a low specific gravity. As a result, the API gravity is high. According to NYMEX, API gravity for light sweet crude for non-US countries is between 32° and 42°. Light sweet crude is highly sought after due to its high naphtha yields. This high amount of naphtha leads to increased gasoline production after processes like catalytic reforming. This gasoline is a high-grade reformate with a good enough octane number to meet the growing energy demands of the.

The current distillation techniques, however, are highly energy intensive. So, to increase gasoline production, it is obligatory to increase the energy investment. This project proposes an alternate approach to the conventional distillation mode called the Heat Integrated Distillation Column (HIDC). The HIDC system reduces the amount of energy required by incorporating a thermosiphon flow of contents between the rectifying and stripping sections, reducing the amount of energy wasted. The heat of the rectifying section is used to vaporize a stream from the stripping section, reducing the overall load on the reboiler. The process is designed to be both economically and technically feasible and sustainable for the long-run.

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List of Symbols

ΔH: Change in enthalpy

A: Area

C: Cost

°C: Degree centigrade

D: Diameter

F: Flow Rate

g: Acceleration due to gravity

H: Height

J: Joule

L: *Length*

Mol: Mole

P: Pressure

Q: Energy Flow rate

t: time

V: Volume

W: *Width*

Cp: Specific Heat Capacity at Constant Pressure

T: Temperature

V: *Volumetric Flow Rat*e

K: Thermal Conductivity

h: Thermal Convectivity

CHAPTER 1

Introduction

1.1 overview

Over the course of million years, numerous underwater living organisms like plants and algae lived across shallow seas, died, and sunk into the seafloor. Eventually, these organisms decomposed under immense pressure and temperatures of changing sea environments and earth, transforming into what today know fossil fuels, namely coal, petroleum, and natural gas. As of today, most of the petroleum reserves are found at places where ancient seas stood. They are also found below ocean floors and giant drill machines are accommodated to extract them to the surface.

The oil itself accompanies numerous colors, existing in dark shades like black and dark brown, along with red, yellow, tan or even green shades, resulted of variated in chemical compositions found across the globe. Clear or lighter colors are observed for composition low at metal and sulfur content.

The primary product of petroleum is gasoline, consumed extensively in everyday life. Lately, concerns have been raised in light of wide scale consumption of fossils, influencing the carbon content of air, and leading to global warming.

Massive amounts of petroleum are found below the Earth's surface, generally existing in tar pits, which bubble to the surface. Crude oil also exists at bottoms of deepest pits and wells and developments are being done to extract it. It must be noted that fossil fuels are non-renewable resources developed over million years and are meant to run out. Experts predict that human are yet to hit 'peak oil' which is described as the time period with highest production levels of fossils, which could come as early as 2050, therefore many industries as of today are focusing on more alternative sources of energy.

1.2 Petroleum Formation

Petroleum is formed under varying geological conditions observed at Earth's surface over million years. Initially, living organisms like alae, plants and planktons when die drift towards deeper oceans. As they sank into sea floor at conclusion of their life cycles, they are buried and crush under pressures of sea along with sediments and further layers of dead organisms' debris. With Earth's constant tectonic plate activities, ancient oceans dried and what was left are so to be called sedimentary basins. Under these basins, Earth's mantle compressed the organic material under extreme pressures and temperatures along with absence of oxygen, leading to decomposition or organic matter into a waxy substance named as kerogen. Further exposure to heat and pressure results in further deterioration, called catagenesis, where the kerogen further converts into hydrocarbons. Hydrocarbon itself are just comprised of hydrogen and carbon atoms, and their combination, composition, chain lengths are depended upon the conditions they were formed in.

1.3 Classification and fundamental chemistry of Petroleum

From gasoline in cars to synthetic fabrics to backpack to lube in machinery, all these products are useful for in their areas and are consistent and reliable. Unlike that, the Crude Oil from which these products are derived are not very uniform nor consistent.

Generally, crude oil is majorly hydrocarbons, with almost 12-13% hydrogen and 85-86% carbon by weight. Oxygen, Nitrogen, sulfur and some metals like iron, copper nickel make up the rest of the composition in small amounts. The primary contributor to the molecular arrangement of hydrocarbon is dependent upon the composition of its former source, like algae, plankton, or others. The secondary contribution is the exposure of heat and pressure to the plants and algae. Resultantly, when extracted from the ground, crude oil consists of hundreds of different petroleum compounds. Lighter oils are almost 97% hydrocarbons in contrast to heavier oils, and bitumens, which can have hydrocarbon as low as 50%. Therefore, its desirable to always refine the petroleum feed to produce useful consumable products.

1.3.1 Saturated chains

Carbon compounds range start from the simplest molecule, methane ($CH4$) which comprises of one carbon atom and four hydrogen atoms. Next is line is ethane ($C2H6$) and so on. This expanding relation can be used to define whole class of molecules named as paraffins, distinguished by its chain-like existence which formula of C_nH_{2n-2}. This class of molecules are referred to as saturated chains, since each valance electron on carbon is accommodated by four different hydrogen or carbon molecules, forming only single bonds, leaving no valance electron behind, giving paraffins a very chemically stable structure.

Paraffins exist is multiple structures, they could either be straight chained such as butane in the figure or isoparaffins (branched chains). Generally, crude oil products are naturally occurring normal paraffins and require intervention and processing by refinery to change them to more desirable isoparaffins. Primary concern with the natural paraffins arise from their poor property as motor fuel in contrast to isoparaffins, which exhibit good engine-combustion attributes. Paraffins with longer carbon chains have wax characteristics.

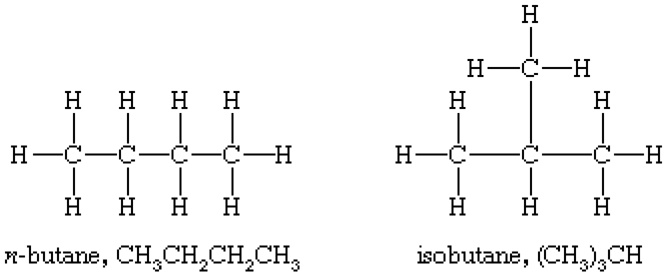


Figure 1:structures of common hydrocarbon compounds

Structures assumed by hydrogen (H) and carbon (C) molecules in four common hydrocarbon compounds. *Once the carbon number of 4 is exceeded, carbon chain has capacity to form a closed cyclic ring structure classified as cyclo-compounds. Cyclic compounds that are saturated in nature are termed as naphthenes. Generally, naphthenes exhibits poor lubricant qualities but on the other side are easier to convert into premium quality gasoline in comparison to paraffins.*

1.3.2 Unsaturated chains

Another important class of petroleum refinery is the compounds consisting of unsaturated molecules. In these, the valance electron of carbon are not separately bonded to hydrogen or other carbon atoms, instead, two or even three electrons are being shared with the neighboring carbon atom, resulting in double and triples bonds. Similar to saturated chains, unsaturated chains can also be branched and cyclic, and are termed as olefins. Crude oil only consist a fraction of olefins and most of them are produced in refineries by further processing. In relative terms, olefins are more reactive compounds and can be converted to longer chains readily.

An extension to unsaturated compound family is called aromatics, which are ring molecules. The most basic aromatic molecule is benzene (C_6H_6). Every atom is benzene is linked via a double bond with the adjacent carbon atom, making benzene a highly chemically active and unstable compound, thereby an important building block of petroleum industry.

These unsaturated carbons demonstrate excellent combustion attributes, but their lack of stability often makes their storage and movement difficult, raising environmental concerns.

1.4 Crude Oil types

As discussed above, crude oil itself is a mixture of numerous carbon-based compounds and its distinguishing process by every single component is not a viable option. Therefore, industries identify crde on basis of most prevalent hydrocarbon compounds namely naphthenes, paraffins and aromatics. Crude oils like the ones which originate is Pennsylvania fields are primarily consisting of paraffins. Rest like the ones in Maxico and Venezuela are heavier crudes rich in naphthenic and bitumen compounds. A chart is demonstrated on composition of some commonly found distillation products of some of the crude oil sources below, ranging from Venezuela's Boscan to lighter oils such as Bass Strait found in Australia. In light of the modern demand curves, that tends to prioritize gasoline, the price of crude is dependent of its lighter components, becoming increasing expensive as lighter component compositions increase. It is however possible to improve light product yields from heavier oils, however, high conversion rates are accompanied by higher operating costs that's makes the effort undesirable.

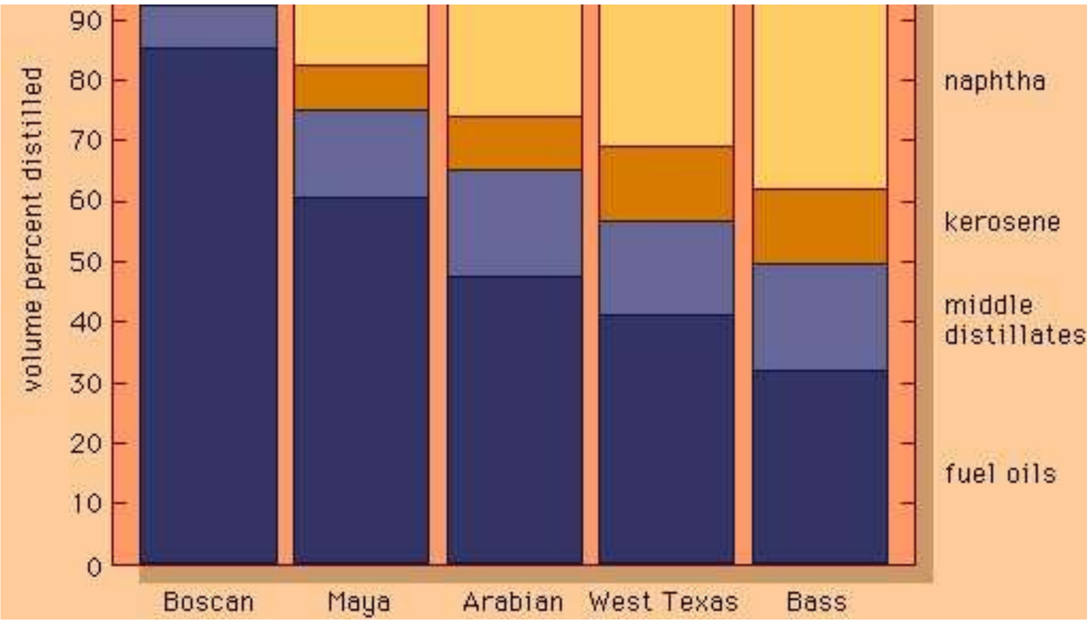


Figure 2:crude oils

Product content of five major crude oils.

In general, there are three basic ways to classify crude oil,

1) By its geological source
2) By its sulfur content
3) Or by its density (in API gravity)

Geographically, oil is categorized in three primary sources, West Texas Intermediate (WTI), Brent Crude, Dubai and Oman. These three are generally used as set reference to rank and price other oil supplies.

WTI is generally a light and sweet version of crude oil, perceived to be of very high quality and are main supplier of oil to North America.

Brent Crude on the other hand is a product of oil fields in Norway and Scotland in the northern sea, and primary supplier of oil to Europe.

Dubai-Oman crude is a sour light crude, recently being produced in Oman along with major production in Dubai and parts of UAE. It is often set to reference for pricing the Persian Gulf Crude, that are majorly consumed in Asia.

Sulfur is an impurity in crude oil and leads to corrosion of metals in the refineries and as well as acts as an air pollutant when burnt. Therefore, in most refineries, the feed of crude is pretreated to reduce sulfur content. The removal of sulfur from crude is called sweetening and crude with high sulfur is termed sour crude.

1.4.1 Density

Density of crude is determined in API gravity. A term introduced by American Petroleum Institute (API) which is a trading association for oil and gas business. The system has standardized different products in oil and gas market such as the likes of pumps, gauges, and drilling machines. They have also numerous measurement units, the most concerned to us being API Gravity. API Gravity is a reference measurement of density of crude to water. A value greater than 10 indicate that the oil is 'Light' and will float on water surface while a value below 10 means that the oil is 'Heavy' and subject to sinking in water.

1.4.2 Crude Oil Reservoirs

Crude Oil is found near the surface of Earth in pockets called reservoirs. Deep beneath the surface of Earth exists high temperatures and pressure, crude oil is generally found in those zones. Naturally, since the pressure under Earth are high, oil moves upwards due to pressure gradient and move closer to the surface until its blocked by impermeable layer of rocks, hundreds of meters under the surface. This crude can be trapped in structures of rocks which are faulted or broken due to Earth's shifting plates and masses. Crude oil can also be trapped stratigraphically, meaning different layers of rocks exhibiting different permeability and porosity for crude oil to migrate, for instance crude can easily move through sandstone layers but can be contained under shale layers.

Scientists and experts generally look for typical topographies features that can trap crude oil beneath the surface. "Seismic Reflection" is a widely applied to locate crude oil underground in suspected rock structures. In the process, sound waves are directed towards the Earth which bounce back from different types of rocks. The returning waves are observed and used to analyze the geological layout underground, for whether, it could possibly hold crude oil reserves or not.

1.4.3 Extraction of crude oil

Crude oil exists in various forms around Earth's surface. Generally, oil is contained underground it reservoirs but in some places like Saudi Arabia, porous upper surface of Earth allows the oil to seep to the surface and form small ponds. The entire quantity of oil in a reservoir is termed as 'Oil-in-Place'. However, is most cases, only a set amount of oil can be extracted from a reservoir as many products are inaccessible, expensive or dangerous to drill. Therefore, the amount of crude that can be extracted from 'Oil-in-Place' is known as the reservoir's oil reserve.

Drilling oil is complicated, and decision are often made on basis of proven oil reserves of an area.

The part of a reservoir's oil-in-place that can be extracted and refined is that reservoir's oil reserves. The decision to invest in complex drilling operations is often made based on a site's proven oil reserves.

There are three methods of drilling applied,

1) Directional
2) Developmental
3) Exploratory

In directional drilling, hole is drilled vertically to access the existing source of oil, and then drill is veered around to find additional useful resources.

In developmental drilling, drilling is done at site where oil reserves have been explored already, it accounts for making new wells at the site and expanding ability to extract more oil.

Lastly, in Exploratory drilling, as the name indicated is done for exploring new reservoirs, at places where there are no proven oil reserves. Its often a very risky task and is accompanied by high failure rates.

1.4.4 Extraction process

Modern extraction process involves use of air rotary drilling rigs and has capacity to work 24 hours. The engine powers a drill bit, which itself is the cutting tool to drill circular holes into the surface. The machine uses hollow drill bits made of steel, while a tungsten rod in applied to cut through rocks. Standard drill bits drill holes around 0.36 meters in dia.

The drill bits function by revolving at speed, cutting through rocks in the Earth surface and chipping them off. A pressurized air stream is pumped into the hollow steel rod and exits at the end of drill bit. The air then rushes upwards to do pressure difference carrying pieces of chipped rocks with it, projecting them at the surface. Experts generally study there rocks to identify the strata of rock the drill is coming across, yielding important bits of information regarding the progress of drilling.

On contact with oil, a gusher is produced as pressured oil moves towards surface due to pressure gradient and project dozens of meters above the surface. The process can be dangerous and usually, equipment like blowout preventer is installed to minimized pressure by redistributing it, resultantly controlling the gusher.

Two types of pumps are used to extract crude oil to the surface, extraction pumps and mud pumps. In mud pump, ‘Mud’ is defined as drilling fluid intended to form boreholes for extraction from reservoir and their primary function is to circulate the drilling fluid in the system. Extraction pumps are more versatile in range, meaning that different variety of pumps exist depending upon the topography, position, and quality of petroleum in the reservoir. For example, a gas pump relies on compressed air, which is pumped downwards, pushing the oil in the reservoir to move upwards due to pressure gradient. A submersible pump on the contrary is submerged promptly into the fluid. The most applied pump is called pumpjack, which relies on a crank that moves a large hammer head, up and down. This movement generated moves a hollow piston beneath the ground as well, pushing oil to the surface.

1.5 Refining petroleum and Refinery Industry

Refining of crude oil is a process involving separation and conversion of petroleum into more useful products. The crude itself comes with numerous impurities, ranging from dissolved components like sulfurs along with insoluble impurities like water and sand. These components have to be separated before the crude oil is further subject to processing where it is partition into numerous components via distillation to yield useful products. These products are then transported through pipelines to required industries before they become available to consumers.

Oil has not been subject to extraction or consumption by humans for a long time. The earliest evidence of extraction of oil date back to 350 CE, China where engineers at the time drilled 244 meters below the ground using bamboo bits to extract oil and transport it via bamboo pipelines. They used it mainly to produce heat in order to evaporate brine and produce salt.

In 7th century, engineers from Japan realized that petroleum can be burned to produce light. Later in 9th century, Persian chemist drafted a way to distill crude oil to produce kerosene. By early 1800s, kerosene extracted from petroleum replaced the use of whale oil in lamps halting the whale hunting process altogether. It was not until 1850s, when modern oil industry was established starting from Poland, where first well was drilled in 1853, and followed by rapid expansion of technology to other countries. 18th century industrial revolution became the driving force demand of petroleum, as steam powered engines became too slow, expensive and scale bound, demand of petroleum rose leading to establishing of a proper industry. In 1858, more oil wells were drilled in Ontario, Canada followed by Pennsylvania, United States In 1859. Once rock oil was discovered in Pennsylvania, Crude oi became accessible in abundance leading to large scale systems for its production and processing. The first refineries relied upon simple units of distillation to sperate petroleum into its constituent by heating the crude oil in a column and then condensing the vapors in different liquid fractions. The first sought product was mainly kerosene, which served as a replace for whale oil, producing much cleaner and brighter flame.

The second sought product was the straight run naphtha, which is a unfinished form of gasoline, and in fact, one of the lighter distillation products. Initially, its application was restricted to its solvent properties due to lack of internal combustion engines at the time. The higher boiling, less volatile compounds were useful lubricants and fuel oils but their existence at that time where mainly novelties.

The oil-drilling procedure reached perfection and quickly spread to Russia AND BY 1890, industries were setup, producing massive quantities of fuel oil and kerosene. Later in the 19th century, first internal combustion engines came to existence establishing a small market for naphtha. However, advancement in automobiles in the early 20th century exponentially grew demand for premium gasoline, providing a room for light petroleum fractions too volatile to be part of kerosene, resulting is a more continuous form of distillation yielding greater yields of naphtha.

By 1910, the market for automobile fuel had grown over market demands for kerosene, and industries were subject to develop new techniques to accommodate increasing demand. The first processes to be applied were thermal cracking, subject to breaking of heavier hydrocarbon under pressurized vessels, yielding lighter components by means of splitting and cracking. The gasoline produced by cracking produced better results in internal combustion engines in comparison with straight run distillation naphtha. By 1930s, airplanes engines with more power output were being used, giving rise to need for gasoline of greater quality, which lead to inclusion of lead-based additives in the fuel to enhance performance. Sophisticated refinery processes were established at the same time, involving use of catalysts to further improve the quality of fuel. The techniques like alkylation, polymerization, catalytic cracking and isomerization, allowed petrochemical industry to combat the high-performance demands which were further accommodated in aircrafts in World War II. In 1950s and 60s, jet fuel and higher quality lubricant demand increased. To meet the continuously increasing demands, petroleum industry had to entertain wider range of processes for higher quality products. New methods like catalytic reforming overshadowed the former thermal reforming and established itself as te primary process to meet fuel quality of higher compression internal combustion engines. Another process named hydrocracking was introduced, which proved to be fundamental process which was versatile and produced improved yields of either jet fuel or kerosene.

As of today, OPEC estimates that around 70million barrels of fuel is produced every day, which accounts for almost 50,000 barrels each minute. Seemingly, these sounds astounding numbers but use of petroleum products have been incorporated into our numerous aspects of life. From shoelaces to fabrics to household gas cylinders, petroleum goods form a fundamental part of our lifestyle.

1.6 Products of petroleum and their usage

1.6.1 Gases

Refineries produce the likes of gases like methane, ethane, propane, butane along with hydrogen. Hydrogen produced is generally consumed within the refinery is desulfurization facilities, removing H_2S gas from the feed stream and then being parted into elemental Sulfur 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 Petrol (Gasoline)

Gasoline is standardized to meet three primary quality demands, steady combustion patterns, ability to ignite in cold conditions, and have minimal additives meeting environmental concerns.

1.6.2.1 Octane Rating

Gasoline is expected in modern engines to provide smooth burning and minimal knocking. That is a major concern because knocking can reduce power output and in return damage the combustion engines. In the stretch of 1920s, engine powers increased and it was observed that some fuels produced more consistent knocking then the rest. Experimental analysis indicated that straight chain heptane was more susceptible to knocking is comparison to isooctane. The determination led to inclusion of standard called octane rating, which is today used to define gasoline quality. Today most motor engines operate at octane rating between 87 to 100 while some gasoline for modern aircraft piston engine is rated between 115 to 130.

Today, each naphtha additive is tested separately for octane rating before its added to the blend. Processes like isomerization, alkylation, reforming, and cracking produces gasoline with octane rating excess of 90 in contrast to straight run naphtha which only has a rating of 70 or less. In late 1920s, it was concluded that addition of lead-based additives like tetraethyl lead used to enhance the rating values of numerous naphtha chains. Some produced a synergistic response to others antagonistic response, uniquely relying upon each naphtha component. Based on this, industries derived a sophisticated methods to produce gasoline blends for desired qualities.

After successful use of lead-based blends by general consumers and their use in World War II, by 1975, environmentalist raised the legislations to restrict the additive use for gasoline in automobiles, inferring to the environmental hazards linked to lead based fuel byproducts. Today, lead additives are banned across Europe, US and numerous other countries around the world. Advancement in techniques to improve low octane naphtha by processes like alkylation and catalytic reforming are applied to produce premium gasoline.

1.6.2.2 Component volatility

The ability for fuel to ignite in lower temperatures is the second criterion the gasoline producers have to meet. This was worked upon by inclusion of butane in the gasoline blend, which itself is a low boiling point paraffin (gaseous at room temperature) and has a high-octane rating. However, heavier hydrocarbons must be accommodated to offset its low energy values to satisfy the energy output of the system. However, environmentalist raised concerns regarding high evaporation rates of butane during refueling and storage, making it a pollutant coupled with the concern that heavier hydrocarbon itself increase the production of unburnt fuel in combustion engine exhaust. This forced legislation reforms in US for gasoline producers to accommodate oxidants in fuel for more complete combustion and reducing pollutant exhausts like unburnt hydrocarbons, carbon monoxide and nitrogen oxides.

Industries tackled that with including oxidants like Ethyl Alcohol, Methyl Tertiary Butyl Ether, and Dimethyl Tertiary Butyl Ether. MTBE and DTBE were later recognized as ground water pollutant causing more reforms in legislations. BY 2005, reformulation of MTBE and DTBE ceased to exist in gasoline blends. Today gasoline has increased amount of Ethyl Alcohol to meet emission requirements while is some countries, MTBE is still used as additive.

1.6.3 Kerosene

Kerosene use has diminished as an illuminant but is still consumed extensively is cooking, space heating and is a predominant fuel is jet engines. Standardization of kerosene is done for its minimum flash point to be at 49 °C for same handling and storage. In domestic consumption, kerosene produces smoke free odorless fuel. The major quality standard for gasoline is in jet engines. The temperatures in higher altitudes where jets fly reach as low as −50 °C, therefore, fuel must be free of water droplets as well as remain liquid, hindering any inclusion of wax particles. Military jet fuels have even stringent specifications to serve their special purposes.

1.6.4 Diesel Fuel

Diesel or gas oil is widely powers energy intensive automobiles like railway trains, buses, trucks or heavy industrial machines. Diesel engines rely on its compression capacities of fuel and air, for gasoline engines, it could of catastrophic as detonation will lead to knocking and damage of engine, however, this process is a necessity for the diesel engine. A optimum diesel engine is expected to produce multiple ignition points simultaneously with in the compression cylinder, and any fuel that enters the cylinder then ignites likewise. Therefore, unlike for gasoline engines, diesel engines have preference for more straight chain hydrocarbons, and has its own cetane rating, which is derived from blends comparison with hexadecane (cetane). An ideal cetane rating is said to be around 50, giving similar combustion attributes like a 50-50 mixture of standard fuel. Slower utilities like ship engines or stationary power plants have tolerance for even heavier gas oils.

Diesel standards have not been stringent until late 1990s. Cetane numbers were still critical, but sulfur levels of fuels were not a concern. Level as high as 5000ppm sulfur were found common across the market. However today, regulations have been placed to contain sulfur concentration to 10 to 15ppm.

1.6.5 Fuel Oil

Fuel oil also termed as furnace oil is the residual product of refinery. Its viscous in nature, has high sulfur content and rich in heavy metals. Therefore, its usage is restricted and often needs to be blended with low viscosity gas oil for make it convenient for handling. Resultantly, furnace oil is the only refinery product that has a lower selling price in comparison to raw crude oil. On burning, furnace oil produces hazardous sulfur trioxide, and therefore has to be regulated by industries by installment of units like cokers and desulfurization units. Heavy metal also impinges its use as they produce ash on combustion and lead to fouling in burners. Removal of these metals is costly as also contribute to low market price of fuel.

1.6.6 Lube Oil

Initially, the lubricating properties of petroleum fraction were depended upon the source of oil it was derived from. The likes of Pennsylvanian crude offered superior properties due to its greater paraffinic content. However, the range of raw material has now been extended with the introduction of techniques like hydrocracking and solvent extraction. The primary characteristic of a lubricant is its viscosity. The demand varies from high-speed spindles in textiles to more viscous material used is wire ropes or open gears. The product range offers numerous characteristics of oils, all serving their special purposes while the automotive industry remains the prime segment in the market. Viscosity rating are allotted by Society of Automotive Engineers, ranging from 5 to 50. These rating are generally applied in United States, but United Kingdom’s Institute of Petroleum offer ratings, while are virtually identical to the ones offered by SAE.

Ordinary mineral oil offers adequate lubricating properties at low temperatures. Nevertheless, over extended temperatures ranges, thinning of oil leads to reduced lubricant properties leaving it unfit for operation. To accommodate these concerns, multigrade polymers wit longer chains are developed, to serve purpose on wider ranges of temperature. For instance, SAE 10W40 rating indicate that the lubricant will have a viscosity of 10W SAE at -18°C and will rise to SAE 40W at 99°C. Thus, providing appropriate lubrication in winters as well as in summers. Antioxidants and detergents are other additives incorporated into lubricating oils to improve performance and maintain engine cleanliness.

Greases

In order to lubricate gears, grease oil must separate surfaces from each other, minimizing friction and reducing wear and tear. Resultantly, it must be able to bear extreme pressures and prevent metal-metal contact. To enable that, often additives like sulfur is added to oils which tends to create a resistant film between the surfaces. Thickening agents like calcium, sodium, aluminum soaps are also added along with special non-soap components like silica, carbon or polyethene.

1.7 Petrochemical products

Petroleum products cover a vast range of products in manufacturing industry. The likes of cosmetics, paints, dry cleaning and others industry extensively use naphtha as its basic solvent. Paper manufacturing and foods employ waxes in their processes. Asphalt is a key ingredient in construction of roads and airfields. Pipeline coating, electrical insulation, waterproof papers and roofing felts all accommodate bitumen in their manufacturing systems. Following are the list of widely employed petrochemical products.

1.7.1 Olefins

After the incorporation of thermal cracking process in refinery in 1920s, the gasoline production increased but the process offered other byproducts which included lower molecular weight olefins. These olefins, particularly ethylene became a valuable building block of petrochemical industry. Polymerization of ethylene along with other components like propylene produced polymers today effectively used in packaging, and everyday goods. Today, ethylene manufacturing by steam cracking is widely applied in the industry.

1.7.2 Polymers

Polymers are plastics, synthetic fibers and rubbers that are manufactured from petrochemical by products. They are called polymers, mainly because they pack long chains of repeated monomer units and therefore have high molecular weights. Major industrial products like polyvinyl chloride, polyester and polyethene are derivatives of their monomers, ethylene, and propylene. Rubbers are composed of petrochemical products like ethylene, benzene, butadiene, and propylene. Similarly, ethyl glycol and terephthalic acid (derived from xylene) is used in creation of polyesters.

1.7.3 Other Chemical

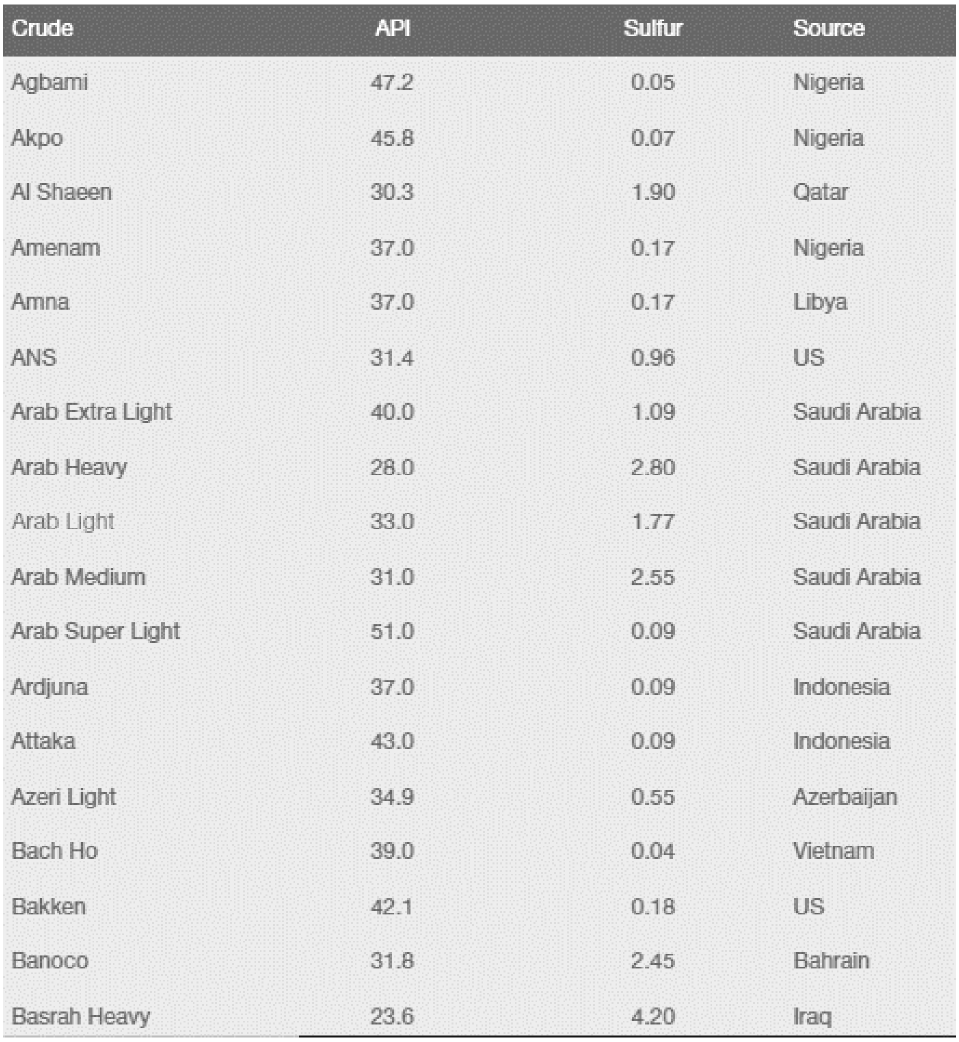
Crude oil has been neglected as source of ammonia and sulfur till the restriction imposed to produce cleaner fuels. Today, they serve as prominent source of inorganic compounds like ammonia and sulfur. Hydrogen derived from hydrocarbon sources like petroleum liquid fractions, natural gas, and residual refinery gases to produce ammonia through harbor process. Similarly, sulfur is derived from desulfurization of oils to produce sulfuric acid. Both sulfur and ammonia today serve as critical components in the fertilizer industry.

CHAPTER 2

Literature Review

2.1 The Light Sweet Crude

Crude oil that when derived from the source having low content of sulfur is said to be sweet crude. Sweet crude is thought to be a valuable source of crude, much more efficient in nature as sulfur tends to reduce yields of numerous refined products like gasoline, gas oil and others. In contrast, crude with higher concentration of sulfur is termed as sour crude. When sweet crude is coupled with lower density, it is termed as light sweet crude, which in fact is one of the most sought-after form of crude oil and often comes with higher price.



Crude	API	Sulfur	Source
Agbami	47.2	0.05	Nigeria
Akpo	45.8	0.07	Nigeria
Al Shaeen	30.3	1.90	Qatar
Amenam	37.0	0.17	Nigeria
Amna	37.0	0.17	Libya
ANS	31.4	0.96	US
Arab Extra Light	40.0	1.09	Saudi Arabia
Arab Heavy	28.0	2.80	Saudi Arabia
Arab Light	33.0	1.77	Saudi Arabia
Arab Medium	31.0	2.55	Saudi Arabia
Arab Super Light	51.0	0.09	Saudi Arabia
Ardjuna	37.0	0.09	Indonesia
Attaka	43.0	0.09	Indonesia
Azeri Light	34.9	0.55	Azerbaijan
Bach Ho	39.0	0.04	Vietnam
Bakken	42.1	0.18	US
Banoco	31.8	2.45	Bahrain
Basrah Heavy	23.6	4.20	Iraq

Figure 3:Geographic variance in WTI crude oil

Second to its sweet or sour nature, another consideration to crude oil is its density which effects its quality. According to API standards of American Petroleum Institute, crude oil can be classified as ‘light’ or ‘heavy’, based upon its relative density to that of water. In API gravity terms, crude oil that possesses the density of water has an API gravity value of 10. The scale itself in inverted, meaning that oil with API gravity value less than 10 will float on water and thereby be termed as ‘Light’. Similarly, crude with API Gravity less than 10 will sink in water and is classed as ‘Heavy’. Generally, heavy crudes are accompanied by high sulfur content along with other impurities like heavy metals, making it less desirable to distil and process.

Once sweet crude is coupled with lower density, it is termed as light sweet crude, which in fact is one of the most sought-after form of crude oil and often comes with higher price. Primarily because of its high efficiency and minimal residual remains, making its refinement easier as well as can be effectively transported due to low density. Below are some of the major sources of crude oil. According to API standard, crude with sulfur content lower than 0.50% is termed as sweet.

2.2 Distillation Techniques

2.2.1 Fractional Distillation

The fundamental process of distinguishing hydrocarbons into more components is done by fractional distillation. The fractions separated move into subsequent processing units like reformers, isomerization unit, cracking unit, alkylation unit or cokers. Furthermore, all these units incorporate their own complex fractionators to separate the products of the process. In modern times, fractionating units operate over extensive periods of time and processing rates are commonly in excess of 100,000 barrels per day reaching up to 200,000 barrels per day for some plants.

The demonstration of a modern crude oil distillation plant is visualized below. A feed of crude is withdrawn from the storage facility, is heated up to 120°C by network of heat exchangers and fed to a desalter. The feed is the passed through a controlled amount of fresh water in desalting drum and components are separated into two different phases, salted water and desalted crude, by applying electric field to the mixtures. Salt removal is critical to avoid corrosion and fouling of tubes in the furnace. The desalted crude in further heated to temperatures upwards of 400°C, dependent upon the crude oil composition and the end product desired. After heating, the crude oil is introduced to the fractionating column, generally 40 to 50m high with plates ranging from 20 to 50 dependent on the reflux ratios and design of distillation column. The common types of trays used in fractionators are either sieve or valve with sieve trays generally perforated with holes of 5 to 6mm is diameter. The former valve trays are similar but include a metal disk that acts as controlling agent for flow through disk form specific process conditions.

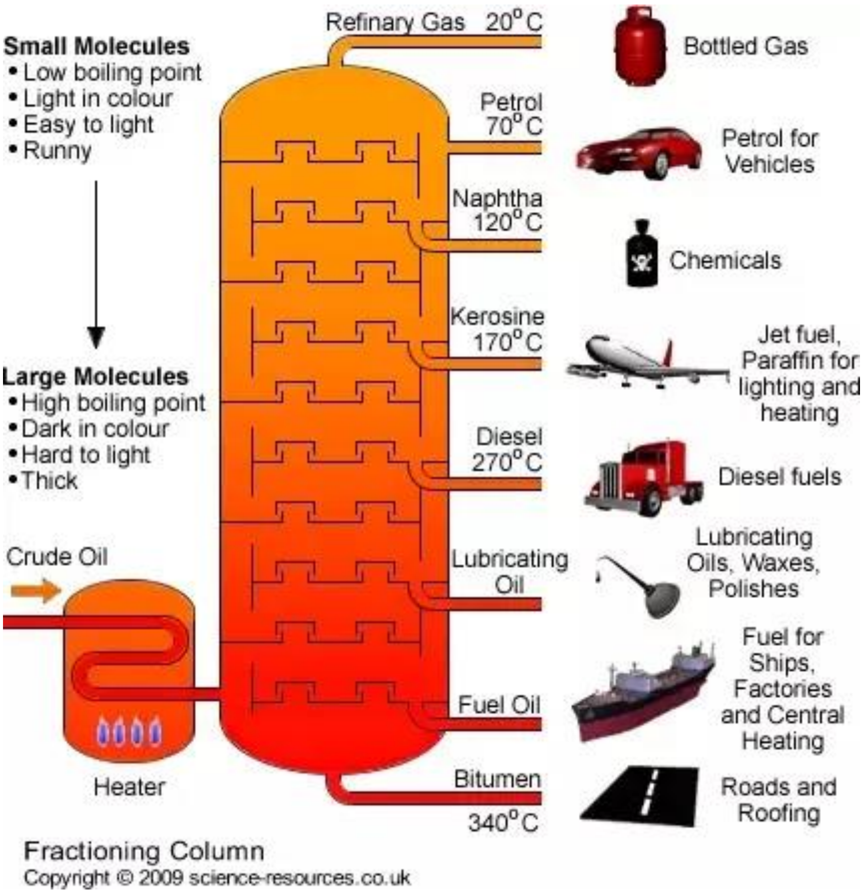


Figure 4: Schematic diagram of a crude-oil fractional distillation column.

The vapors leaving from top of the column are condensed to liquid by a water or air-based cooler. The remaining uncondensed gas becomes part of fuel gas system of the refinery. A pressure control valve is installed on the fuel gas line and serves the purpose of maintaining the pressure conditions of the fractionating column, normally set to be around 1atm. A certain degree of reflux ratio is set according to the distillation specifications, and the reflux drum according to its setting returns the condensed liquid back to the column which descends from tray to tray, exchanging heat with rising vapors, intensifying the distillation process. The exchange of heat leads to condensation and evaporation of different components within the column and with sufficient time, a equilibrium is established between condensing and evaporating vapors leading to gradation of temperatures in the subsequent trays. The presence of trays and reflux is what separates the simple distillation process with fractional distillation.

Side streams are introduced at numerous points of columns to withdraw intermediate products. The products itself are termed intermediate because of their natural properties exist between that of top and bottom product. The products the drawn at variable boiling ranges, some of them as given below:

Light naphtha: 20–95 °C (70–200 °F) from overhead.

Heavy Naphtha: 90–165 °C (195– 330 °F) from first sidestream.

Kerosene Oil: 150–245 °C (300–475 °F) from second sidestream.

Diesel: 215–315 °C (420–600 °F) from third sidestream.

Heavier oils that do not vaporize flow downwards to the lower part of the column often referred to as stripping section as it often operates at higher temperatures and tends to remove light constituent of the crude oil. A stream of steam is often injected at the bottom of the fractionator to lower the pressure of the section, allowing the less volatile components to vaporize, thereby assisting in separation. The final stream that leaves the bottom section often consist of heavier gas oil, with boiling point about 285–370 °C and are often used as suitable additive in industrial fuels. They can however be subjected to further processing in forms of vacuum distillation to extract the remaining oils. Lubricants and gas oil are the major products of the distillation. Gas oil can further be accommodated as feedstock for cracking process producing lighter oil constituents.

2.2.2 Heat Pump System

Distillation remains the most sought after separation technology in industries and refineries as of today, primarily because of its ravishing purification attributes, tremendous production rates and turndown ratio, and a simpler textbook design procedure. With time, more proposed techniques have been established to handle streams that involve less favorable thermodynamic characteristics such as low volatility difference or azeotropic blends. The major concerns surrounding distillation process is very low thermodynamic efficiency (5-10%) as well as high energy requirement for the columns, anywhere between 1MW to 100MW, depending upon the size and specification of column.

Constant improvements are being made in the system, like addition of solvents in extractive distillation to promote relative volatility, separating agents in azeotropic distillation to dissociate the azeotropes resultantly reducing boiler duties, column height and reflux ratio. Further improvement in packing technologies and increased surface area for maximum contact have effectively improved distillation columns. The primary focus in heat pump scheme is to reduce energy consumption of the column and is demonstrated in the text.

2.2.2.1 Industrial Technology

Industries often imply a onion-model to develop technologies in effort to improve industrial process energy efficiency. The representation of model is given below:

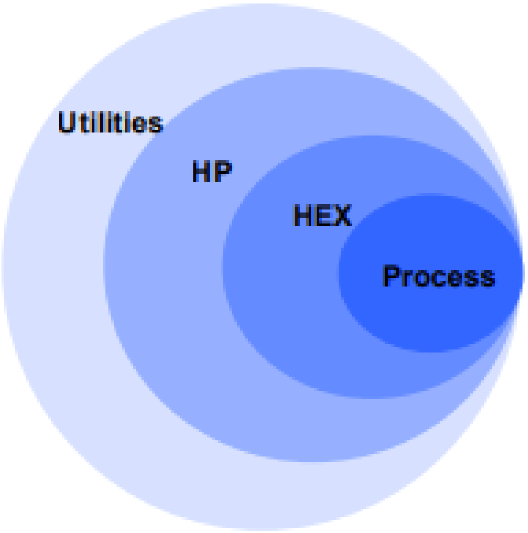


Figure 5 Onion model for energy efficiency improvement

The first shell is characterized as the original process like reaction or in this case, separation and the process is optimized on basis of economics of energy and balancing out operating costs, revenues, and investment. Along with that, improvement in process of distillation is brought, like those in extractive distillation where solvents are introduced. Further optimization is done by finding ideal configuration of column trays, diameter, height, reflux ratio and feed point.

The second shell represents the reduction of energy consumption by use of heat integration by heat exchangers (HEX). The driving force requirement for heat exchanger limits its capacity to a certain set point until further addition becomes economically unviable. Pinch technology rule of thumb often drives the optimization of heat exchanger network and states to not transfer heat across pinch temperatures. Furthermore, enthalpy flow vs temperatures graphs represent a grand composite curve that illustrates the total heating and cooling that is required for a plant. ΔT_{pinch} is the temperature difference is driven by the economy, meaning a greater ΔT_{pinch} value leads to a lower investment cost in the heat exchanger but thereby increase the utility cost. Heat integration has become a standard tool in optimizing industry in process design since 1980s based upon the pinch technology. Heat exchanger technology has dramatically improved offering heat exchange areas of 200-00m^2/m^3. Thereby, the optimum temperature difference has drastically reduced as compact aluminum-based heat exchangers with multi-effect evaporators offering heat exchange at temperature difference as low as 1-2°C. The standard heat integration applied today relies upon preheating of bottom feed streams. Further reduction is energy can be observed with introduction of side reboiler at different sections of columns, limiting the energy demands of main reboilers and condenser.

The third shell indicates a possibility to minimize energy consumption by heat pump scheme. By definition, heat pump is utility that transfers heat from lower temperature source to higher temperature source. The heat pump originated from refrigeration technique and is today implied in air conditioning units too, serving cooling properties by extracting heat from lower temperature source and sending in to higher temperature source in summers and vice versa in winters by reversing the flow of refrigerant in the system.

However, with its effectives, we need to consider the following issues with heat pump:

- Heat pump system requires a temperature lift.
- Thermodynamic cycles and heat pump efficiency as to be taken into account.
- Pinch temperature and adaptability of the process
- The enthalpy balance of the process
- Selection of heat pump utility and its constraint
- System existing configuration
- The capital cost of plant against the utility cost economics.

2.2.2.2 Heat pump configuration in Distillation Column

The serving purpose of heat pump scheme is to extract heat from condensing vapors and deliver it to reboiler for evaporation. Since the reboiler operates at much higher temperature than the column, a heat pump is required. Essentially there are two methods to incorporate heat pump in distillation, a vapor compression column (VC) or a vapor recompression column (VRC), both are demonstrated in the figure below alongside the conventional column (CC).

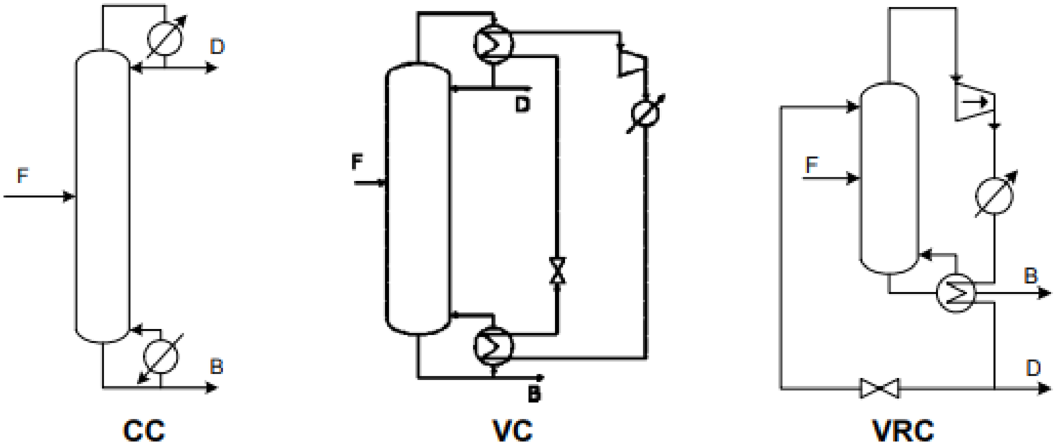


Figure 6: Heat Pump distillation column

In the simple conventional column, the heat addition occurs at the reboiler and is extracted in the condenser while the fractionating column remains adiabatic. The duties for reboiler and condenser are generally in close proximity of magnitude. Ideally, heat pumps are more desirable for small temperature lift, delivering better efficiency and saving greater cost for close boiling systems. Therefore, in there is a critical temperature lift with heat pump system after which, the economics of the plants are no longer beneficial.

A working fluid is incorporated in the vapor compression system, which evaporates in the condenser, consequently compressed to higher saturation temperatures in the compressor, passed through a throttle valve which induces cooling and brings the temperature down under the condenser temperature. An important degree of freedom in design is the selection of heat pump working fluid. Ethylene-ethane separation is an ideal industrial example of vapor compression system that uses propylene as working fluid.

The difference between the vapor recompression column and the vapor compression column in the working fluid. In VRC, the working fluid is the vapor leaving from top of the column and are compressed by the compressor, condensing at the reboiler and then refluxed via pressure reduction by valve to top of the column. The small condenser is ideally desirable to produce a corrective heat input generated by the working compressor. The primary advantage VRC holds over VC is the small condenser and a considerably smaller temperature lift (around 5°C lower), primarily because heat exchanger happens only once. Therefore, VRC produces better thermodynamic efficiency and has been a standard technology over VC in most industries.

Vapor recompression cycle system compression ratio is dependent of saturation curves pT, for bottom and top composition, the difference in temperature across the column, the needed temperature difference needed to operate the heat exchanger and the columns pressure drop. A schematical representation of heat pump embedded in distillation unit is shown below:

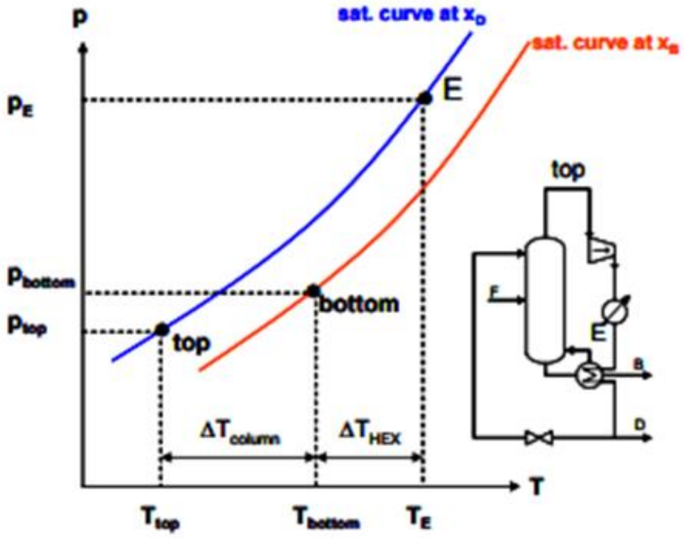


Figure 7 Pressure temperature Diagram for VRC cycle

The distillate and bottom product saturation curves are labeled as x_D and x_B, defined by the thermodynamic model. Normally across most columns, design is based on pressure at the top of column, p_{top}. Likewise for the desired distillate quality x_D, the top temperature is selected and mapped onto the saturation curve. The bottom pressure is however dependent upon the columns overall pressure drop, and T_{Bottom} is mapped onto the saturation curve. Furthermore, the deduction for ideal temperature of trim condenser are made on basis of optimized temperature difference across the heat exchanger, generally to be around 5°C. Assumption is draw that the trim condenser installed works by de-superheating the compressed vapor E and is found again ate the saturation curve x_D. After this, other variables like pressure ratio, temperature lift and shaft work can be deduced:

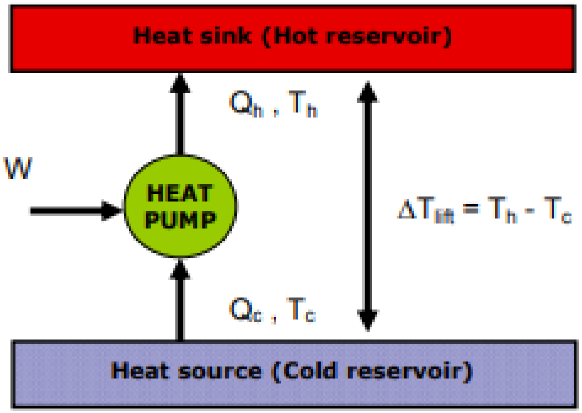
$$T_h - T_c = T_{top} - T_E = \Delta T_{column} + \Delta T_{HEX}$$

Equipment of Heat Pump

Heat pump operation revolves around moving heat from lower temperature source to a higher temperature sink. According to first law of thermodynamic applied on this system, the heat withdrawn from cold source (Q_C) is related to heat delivered to hot source (Q_H) and external work done by following mathematical representation:

Equation 1

$$Q_h = Q_c + W$$



The coefficient of performance (COP) of the system is measure of heat pump performance and is represented by ratio of heat rejected to hot source and the work done on the system:

$$COP = \frac{Q_h}{W}$$

The maximum reachable performance value for any heat pump is COP_C, and is related to Carnot cycle by following equation:

$$COP_c = \frac{T_h}{T_h - T_c}$$

T_h-T_c is defined as temperature difference across the column in addition to temperature difference observed across the heat exchanger as demonstrated by the first equation. Ratio among the two COP values is termed as heat pumps exergetic efficiency, η_e:

Equation 2

$$\eta_e = \frac{W_C}{W_{VRC}}$$

A visualization of Carnot and VRC cycle is demonstrated via a TS-diagram for identical $Q_h = Q_{reboiler}$:

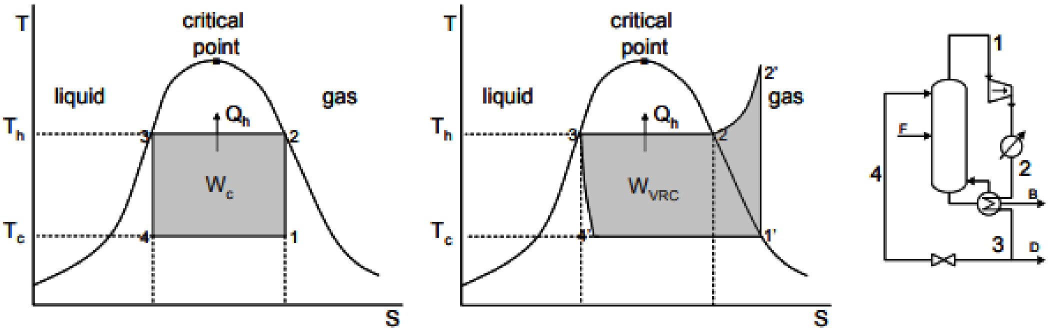


Figure 8 Carnot Cycle and reverse Rankine Cycle represented by TS-diagram for VRC.

The TS-diagram in this case is a simplification as the VRC itself is an open cycle. The determination of cost saving is dependent upon primary energy use of both components, PE_{CC} and PE_{HP}. Along with that, efficiency of steam boilers, $\eta_{Boilers}$ and the total electrical energy provided by power plant to run compressor η_{el} is also to be accounted for.

Conventional columns primary energy consumption can simply be defined as:

$$PE_{CC} = \frac{Q_{reboiler}}{\eta_{boiler}}$$

Now for VRC system, the equation changes to as follow:

Equation 3

$$PE_{VRC} = \frac{W}{\eta_{el}} = \frac{Q_{reboiler}}{COP \cdot \eta_{el}} = \frac{Q_{reboiler} \cdot \left(\Delta T_{column} + \Delta T_{HEX}\right)}{\eta_e \cdot T_h \cdot \eta_{el}}$$

Thus the overall energy saving could be simplified by deducting the PE_{cc} value from PE_{VRC} giving the following overall saving:

Equation 4

$$PES = PE_{CC} - PE_{VRC} = Q_{reboiler} \cdot \left(\frac{1}{\eta_{boiler}} - \frac{\Delta T_{column} + \Delta T_{HEX}}{\eta_e \cdot T_h \cdot \eta_{el}} \right)$$

The equation represents that primary cost saving is inversely proportional to the temperature difference over the column. At certain values of ΔT_{column}, the primary cost saving becomes ineffective as it fails to accommodate the compressor investment. At lower ΔT_{column} , especially when heat exchanger temperature difference coincides with the column temperature difference, The investment on the compact heat exchanger as member of heat pump equipment becomes more attractive.

2.2.3 Heat Integrated Distillation Columns (HIDC)

HIDC is a promising diabatic system which allows heat exchange between a high-pressure rectifier with a low-pressure stripper as shown in the figure.

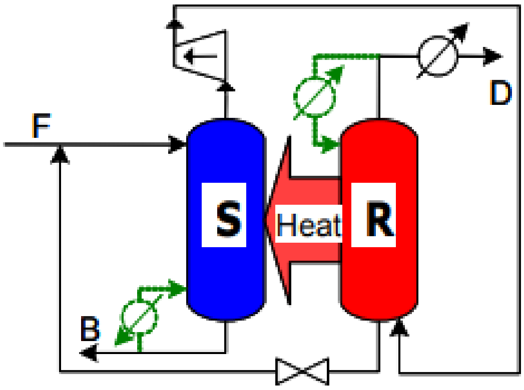


Figure 9 Simple Heat integrated Distillation Model.

In HIDC, vapor stream from the stripping section is extracted, compressed and introduced to the rectifying column, forming an internal reflux which is them sent back to the top of the stripper. Condensation latent heat is utilized to vaporize the liquid on the stripper side. This massively reduces the reboiler duty and requires a small external reflux to adjust desired distillate purity.

The pressure ratio optimization is relied upon the cost of power for compressor and investment costs involved in setting up the compressor and the Heat Integrated Distillation Column. HIDC in return improves the utility cost by addition 25-35% in comparison to vapor recompression column as well as reduce annual operating cost by 10-20%.

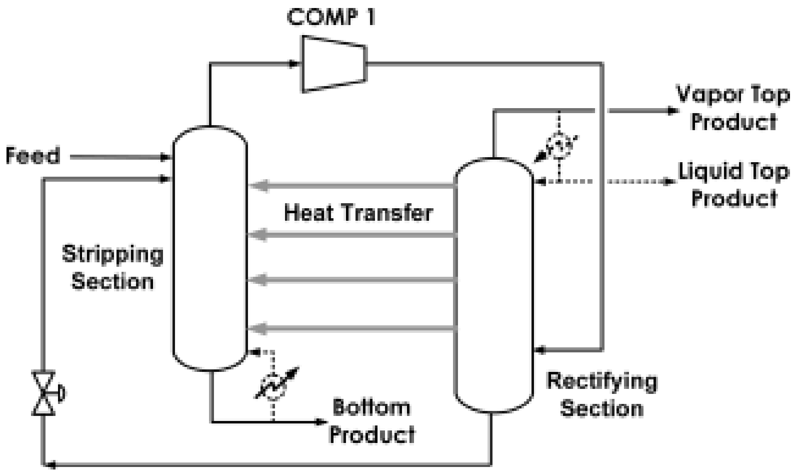


Figure 10 HIDC Configuration

In HIDC, the bottom product of the rectifying column is returned to the stripper after equalizing its pressure accordingly by a throttle valve. Primarily, the pressure difference that exists between the stripping and rectifying section to enable parts of rectifying section to operate at higher temperature than the stripping section, thereby creating a driving force for heat transfer between the columns. The transfer of heat in extended through the length of columns, meaning a constant evaporation at stripping column and likewise, continuous condensation in the rectifying column. Improving heat transfer between the respective columns significantly reduce the reboiler and condenser duties. Theoretically, HIDC enables design of column involving no reboiler or reflux, by appropriately adjusting the pressure ratio among column, feed state and total heat exchange between stripping and rectifying sections.

HIDC and heat pump distillation schemes varying in the set up of how stripping and rectifying sections are constructed. In a simple heat pump representation, the column is set such as that the rectifying section is thermally combined with the bottom of stripping column without the inclusion of heat exchanger. To provide further energy benefits and cost savings, intermediate heat exchanger could be installed along the length of column, where deemed suitable.

In contrast, Heat integration scheme comes with its restriction to the arrangement for heat transfer, with heat transfer occurring between top of stripping section and top of rectifying section, and between intermediate sites along the length of each column. Lastly heat is exchanged from among the bottoms of both columns. The difference between heat pump and HIDC can also be observed at numerous location where the heat integration is installed. For instance, heat pump configuration involving more than 2 or more heat exchanger still has much lest heat integration sites in comparison to HIDC. On the contrary, HIDC scheme coupled with its design complexity can be challenging to operate and control while holding on to its better energy efficiency. The heat pump scheme gives an attractive proposition due to its simpler design of column, easier control, and operation but the energy efficiency attributes of the HIDC system concludes to be the decisive factor in many operations. A graphical representation for HIDC and heap pump system is shown below drawing comparison between plate number (number of stages) and the power consumption of the column.

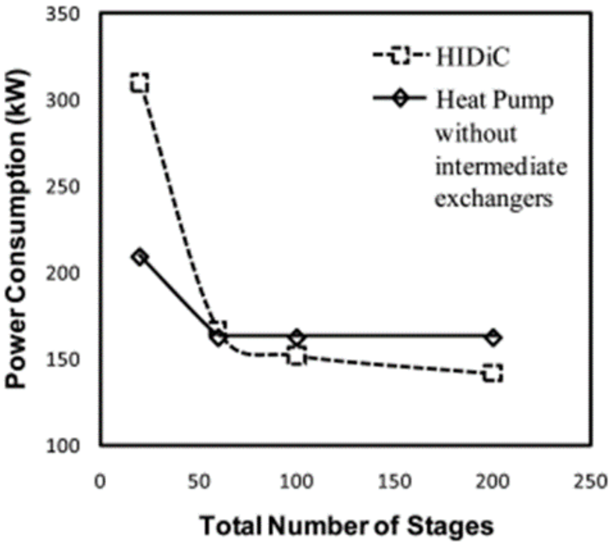


Figure 11 HIDC vs Heat Pump power consumption comparison based on columns number of stages.

Even in its evaluation against conventional units on several pilot plants, Aspen Plus Simulation and bench scale experiments, HIDC has proved to be promising with results indicating energy saving up to 60%.

CHAPTER 3

Process description

3.1 PROCESS DESCRIPTION

The proposed Heat integrated design starts off with a pre-treatment of the crude oil feed. This pre-treatment is achieved by two main components: the desalter and the pre-flash separator. By definition, light sweet crude oil is known to have a very low sulfur and salts content but there are trace amounts still present. Salts, especially, need to be removed in pre-treatment so they do not hinder any processes down the line. Skipping salt removal would lead to increased scaling and pressure drops within heat exchange equipment, lowering their overall performance. Horizontal electrostatic desalters are employed to further reduce the salt content of the light sweet crude. The desalted crude oil is then sent to the pre-flash vessel to be separated into liquid and vapor feed streams.

The liquid stream from the pre flash separator is fed towards the top of the stripping column. The vapor streams from the top of the stripping column and the pre-flash are mixed and fed to a compressor. The pressurized vapor mix from this compressor is fed to the bottom of the rectifying section. Since both the rectifying and stripping sections represent a single distillation unit, there is only one reboiler present at the bottom of the stripping section and only one air-cooler condenser present towards the top of the rectifier. Both the stripper and rectifiers are attached with side-stripping sections to lower the partial pressure of hydrocarbons by incorporating steam. This aids the vaporization of the more volatile components. Heat integration is achieved by withdrawing a liquid stream from the top of the stripping section and allowing it to exchange heat with vapors rising from the bottom of the rectifier.

The product streams obtained from the stripping section are diesel oil and furnace oil. The product streams for the rectifier are the overhead products, gasoline and kerosene. The overhead products are further subjected to a flash separator to yield gaseous and naphtha products. The products from the distillation column are cooled down to workable temperatures by a network of Shell and Tube heat exchangers, all utilizing the same water stream as the coolant.

3.2 Process flow diagram:

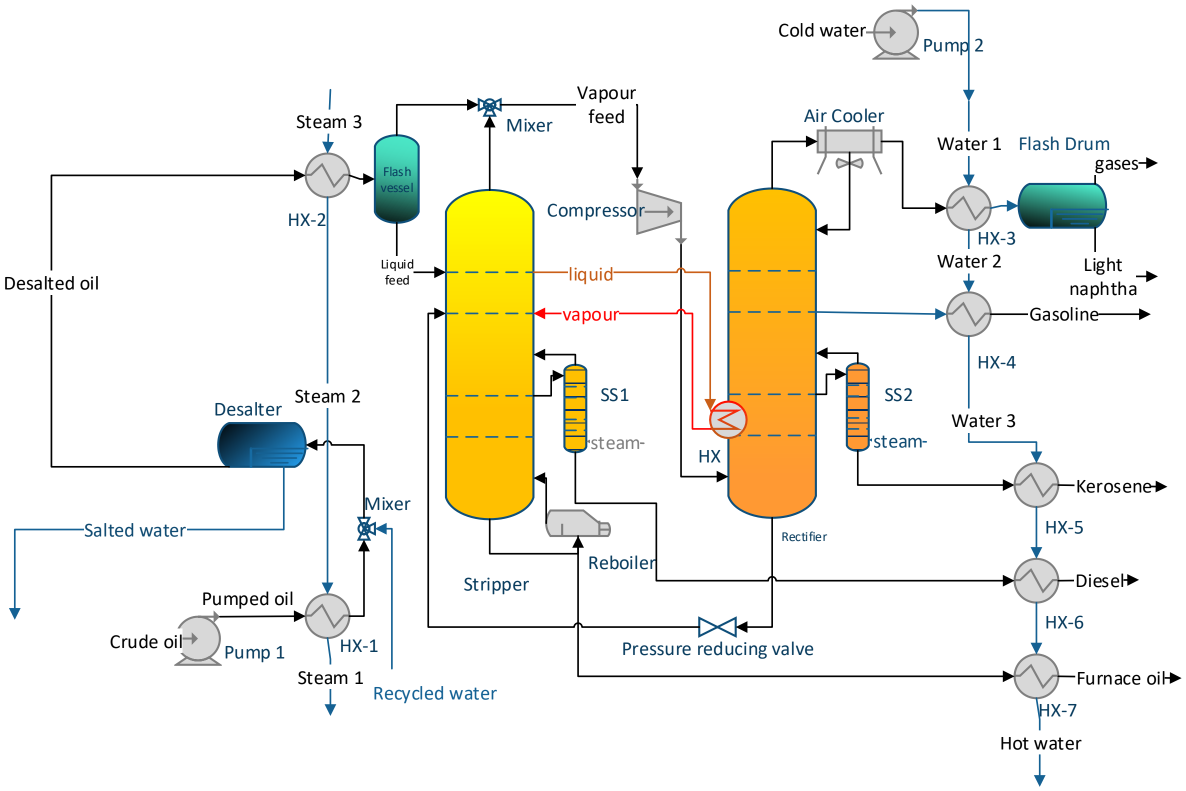


Figure 12: HIDIC process flow diagram

3.3 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 centimeter 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 [1,2]. The proposed design utilized a single stage separator for the light sweet crude blend. Multiple stages were not needed since the feed was light and sweet. The operating conditions were finalized using relevant literature. A typical temperature range for electrostatic desalters ranges from 50 to 150 degrees Celsius. We decided upon an operating temperature of 107.5^{o}C to ensure low enough density and viscosity for better settling rates. Washing water ratio was taken to be 0.255% of the crude oil feed [1]. The settled water exits with the salts and moisture previously present within the crude itself.

3.4 Pre-flash Separator

A flash vessel is employed right after the desalter to split the feed into gaseous and liquid streams. 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 a heat exchanger. The temperature of the flash separation was 150^{o}C and the pressure was reduced to 1 bar via pressure drop of 10 kPa. The purpose of this particular piece of equipment 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 [4].

3.5 Distillation Columns

There are two distillation columns: one represents the stripping section while the other is the rectifying section. The rectifying section operates at a higher pressure (between 5 to 6 bar) as compared to the stripping section (between 0.1 to 1 bar). By maintaining a sufficient pressure ratio, we were able to keep the temperature of the sections within the rectifying column to be higher than their corresponding sections in the stripping column. As a result, heat exchange occurs throughout column lengths. Bottoms reboiler and tops condenser duties are further reduced as a result of this. No external reboil or condensation is needed for the operation of this HIDIC. The operating conditions selected are tabulated as follows:

Table 1: Column parameters

	Stripping Column	***Rectifying Column***
Top Temperature (^{o}C)	143.8	143.9
Top Pressure (bar)	0.1	5
Bottom temperature (^{o}C)	313.8	280.6
Bottom Pressure (bar)	0.2	6.0

3.6 Heat Exchange Network

Products withdrawn from both fractionating columns are sent to a network of heat exchangers to be cooled down. The heat exchanger HX-3 is responsible for cooling off the products from the top of the Rectifying column from 149.3^{o}C to 70^{o}C before they are sent to a flash drum for further phase splitting. HX-4 is employed to cool down the straight run gasoline stream from the rectifier down to 50^{o}C. HX-5 cools down kerosene coming from the rectifier's side stripper to 45^{o}C. HX-6 and HX-7 are responsible from cooling the Diesel and Furnace fuel oil streams from the Stripper, respectively. All these heat exchangers utilize a single cooling water stream to save up on utility costs.

CHAPTER 4

Material Balance

4.1 MATERIAL BALANCE

Since the feed rate of 10,000 BPD was specified within the title of the process itself, balances were just a simple matter of applying law of conservation of mass and energy to each of the important components present within the proposed design. Also, since all of this equipment were unit operations, no chemical reactions had to be considered. This further simplified the task at hand. All the hypothetical components generated from the petroleum assay data are included in the balances as well. Individual component balances are shown, unless otherwise stated. The TBP analysis of the crude oil feed used is as follows:

Table 2: Petroleum Assay

Cut #	Wt.%	Cumulative Wt.%	Cumulative Vol.%	Temp. ^oC	Sp.Gr. 60/60^oF
1	3.00	3.00	3.76	45.0	0.639
2	4.25	7.25	8.89	70.0	0.664
3	8.00	15.25	17.84	95.0	0.716
4	9.65	24.90	28.22	120.0	0.745
5	7.36	32.26	35.98	145.0	0.760
6	4.37	36.63	40.49	160.0	0.776
7	2.45	39.08	43.01	170.0	0.779
8	5.94	45.02	49.06	195.0	0.786
9	5.12	50.14	54.22	220.0	0.794
10	4.03	54.17	58.16	235.0	0.820
11	5.42	59.59	63.39	270.0	0.830
12	6.04	65.63	69.10	295.0	0.847
13	4.04	69.67	72.89	320.0	0.854

14	3.57	73.24	76.22	345.0	0.859
15	3.82	77.06	79.73	371.1	0.872
16	19.98	-----	-----	-----	0.930

4.1.1 Mixer 1

From law of conservation of mass, we get:

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

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

Since $\dot{m}_1 = 132.2\frac{kg}{h}\ and\ \dot{m}_2 = 51800\frac{kg}{h}\ , \dot{m}_3 = \dot{m}_1 + m_2^{\dot{}} = 51932\frac{kg}{h}$

Results are tabulated as follows:

Table 3: Mixer 1

	PROCESSED WATER		**WARM OIL**		**OIL-WATER MIXTURE**	
	y_i	Mass flow (kg/h)	y_i	Mass flow (kg/h)	y_i	Mass flow (kg/h)
Methane	0.00	0.00	0.00	0.00	0.00	0.00
Ethane	0.00	0.00	0.00	0.00	0.00	0.00
Propane	0.00	0.00	0.00	258.19	0.00	254.26
i-Butane	0.00	0.00	0.00	170.73	0.00	168.13
n-Butane	0.00	0.00	0.01	586.00	0.01	577.08
i-Pentane	0.00	0.00	0.01	439.80	0.01	433.10
n-Pentane	0.00	0.00	0.02	872.20	0.02	858.92
H2O	1.00	132.20	0.00	0.00	0.02	920.73
NACL	0.00	0.00	0.00	133.91	0.00	131.87
NBP[0]49*	0.00	0.00	0.02	918.93	0.02	904.94
NBP[0]63*	0.00	0.00	0.03	1513.34	0.03	1490.30
NBP[0]77*	0.00	0.00	0.42	21865.61	0.41	21532.67
NBP[0]90*	0.00	0.00	0.05	2731.23	0.05	2689.64
NBP[0]104*	0.00	0.00	0.05	2778.57	0.05	2736.26
NBP[0]118*	0.00	0.00	0.04	2267.33	0.04	2232.80

NBP[0]133*	0.00	0.00	0.03	1778.59	0.03	1751.51
NBP[0]147*	0.00	0.00	0.03	1683.55	0.03	1657.92
NBP[0]161*	0.00	0.00	0.03	1405.36	0.03	1383.96
NBP[0]176*	0.00	0.00	0.02	1220.25	0.02	1201.67
NBP[0]190*	0.00	0.00	0.02	1054.04	0.02	1037.99
NBP[0]205*	0.00	0.00	0.02	895.15	0.02	881.52
NBP[0]220*	0.00	0.00	0.02	995.02	0.02	979.87
NBP[0]233*	0.00	0.00	0.02	865.19	0.02	852.02
NBP[0]248*	0.00	0.00	0.01	592.23	0.01	583.21
NBP[0]262*	0.00	0.00	0.01	563.10	0.01	554.52
NBP[0]277*	0.00	0.00	0.01	736.73	0.01	725.51
NBP[0]290*	0.00	0.00	0.01	694.31	0.01	683.74
NBP[0]305*	0.00	0.00	0.01	490.67	0.01	483.20
NBP[0]319*	0.00	0.00	0.01	407.99	0.01	401.78
NBP[0]334*	0.00	0.00	0.01	374.64	0.01	368.94
NBP[0]348*	0.00	0.00	0.01	356.11	0.01	350.69
NBP[0]362*	0.00	0.00	0.01	332.26	0.01	327.20
NBP[0]377*	0.00	0.00	0.01	295.05	0.01	290.56
NBP[0]391*	0.00	0.00	0.01	275.69	0.01	271.49
NBP[0]405*	0.00	0.00	0.00	246.94	0.00	243.18
NBP[0]420*	0.00	0.00	0.00	211.16	0.00	207.94
NBP[0]440*	0.00	0.00	0.01	390.46	0.01	384.51
NBP[0]468*	0.00	0.00	0.01	290.78	0.01	286.35
NBP[0]497*	0.00	0.00	0.00	237.94	0.00	234.32
NBP[0]523*	0.00	0.00	0.00	226.56	0.00	223.11
NBP[0]551*	0.00	0.00	0.00	146.00	0.00	143.78
NBP[0]580*	0.00	0.00	0.00	121.90	0.00	120.05
NBP[0]606*	0.00	0.00	0.00	103.79	0.00	102.21
NBP[0]634*	0.00	0.00	0.00	72.13	0.00	71.03
NBP[0]672*	0.00	0.00	0.00	82.77	0.00	81.51
NBP[0]722*	0.00	0.00	0.00	58.05	0.00	57.17
NBP[0]769*	0.00	0.00	0.00	59.75	0.00	58.84
Total	1.00	132.20	1.00	51800.00	1.00	51932.00

4.1.2 Desalter

From law of conservation of mass, we get:

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

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

Since

$$\dot{m}_1 = 51932\frac{kg}{h}, \dot{m}_4 = 52.57\frac{kg}{h}\ and\ \dot{m}_2 = 40830\frac{kg}{h}, \dot{m}_3 = \dot{m}_1 - \dot{m}_2 - \dot{m}_3$$

$$Thus\ \dot{m}_3 = 11040\frac{kg}{h}$$

4.1.3 Pre-Flash Separator

Since this was a multicomponent flash separation of the liquid and vapor phases, using the K-values obtained from HYSYS, we solved for the ratio of resultant vapor and liquid flow rates (β) first using the Rachford-Rice equation on MS Excel. An initial value of β was fed to calculate the Rachford-Rice Solutions:

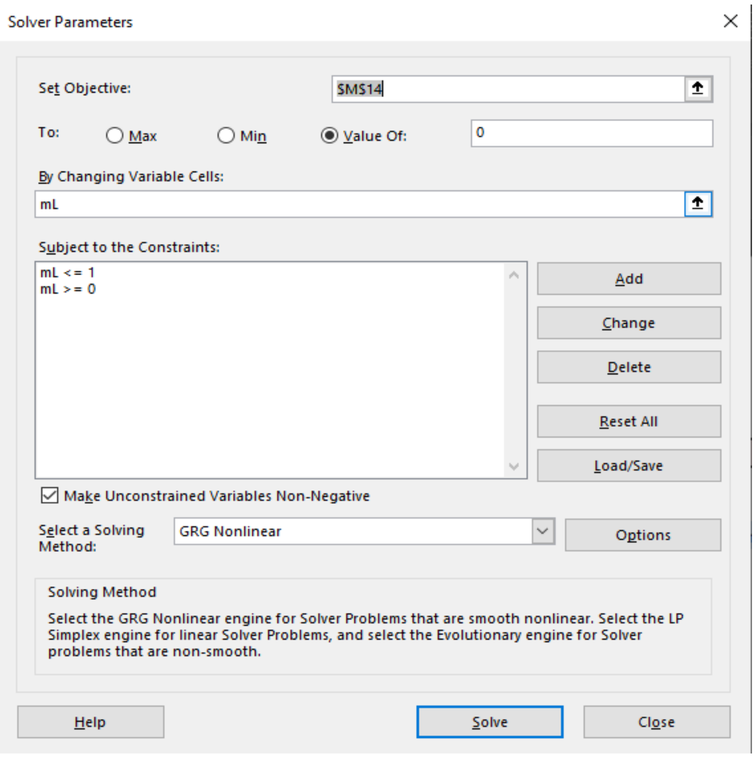
$$\sum_i \frac{z_i(K_i - 1)}{1 + \beta(K_i - 1)} = 0$$

Table 4:Pre-Flash Separator

	Eq Const (K value)	Mole fraction in liquid feed	Rachford-Rice equation Solution
Propane	57.34903911163250	0.00053	0.00080
i-Butane	33.01049600991910	0.00063	0.00093
n-Butane	27.14221329983710	0.00262	0.00386
i-Pentane	15.46322523484120	0.00331	0.00466

n-Pentane	13.45417165804950	0.00745	0.01033
H2O	157.72618792178100	0.00056	0.00087
NBP[0]49*	9.93625198432051	0.00983	0.01519
NBP[0]63*	7.50517026931785	0.01967	0.02614
NBP[0]77*	5.58860969988705	0.33669	0.42380
NBP[0]90*	4.05216216479080	0.04873	0.05676
NBP[0]104*	2.92301748043442	0.05595	0.05778
NBP[0]118*	2.11795691143355	0.05007	0.04314
NBP[0]133*	1.48748558892855	0.04231	0.02756
NBP[0]147*	1.04439262062024	0.04219	0.01567
NBP[0]161*	0.72940215713311	0.03652	0.00158
NBP[0]176*	0.49828717273454	0.03254	-0.01065
NBP[0]190*	0.34112241183090	0.02858	-0.02114
NBP[0]205*	0.22729161914240	0.02456	-0.02800
NBP[0]220*	0.14897990729909	0.02750	-0.04208
NBP[0]233*	0.10062236458907	0.02401	-0.04493
NBP[0]248*	0.06291806676560	0.01648	-0.03499
NBP[0]262*	0.03928792218072	0.01570	-0.03682
NBP[0]277*	0.02419056693978	0.02056	-0.05139
NBP[0]290*	0.01542249572690	0.01939	-0.05049
NBP[0]305*	0.00915714664413	0.01371	-0.03656
NBP[0]319*	0.00536791223658	0.01140	-0.03094
NBP[0]334*	0.00310874567670	0.01047	-0.02871
NBP[0]348*	0.00177442317508	0.00995	-0.02746
NBP[0]362*	0.00099116858355	0.00929	-0.02572
NBP[0]377*	0.00053849424579	0.00825	-0.02289
NBP[0]391*	0.00028458068055	0.00771	-0.02142
NBP[0]405*	0.00014936691121	0.00690	-0.01920
NBP[0]420*	0.00007475823356	0.00590	-0.01642
NBP[0]440*	0.00002734818703	0.01091	-0.03037
NBP[0]468*	0.00000694458255	0.00813	-0.02262
NBP[0]497*	0.00000151466710	0.00665	-0.01851
NBP[0]523*	0.00000037533449	0.00633	-0.01763
NBP[0]551*	0.00000007957925	0.00408	-0.01136
NBP[0]580*	0.00000001391144	0.00341	-0.00948
NBP[0]606*	0.00000000274249	0.00290	-0.00807
NBP[0]634*	0.00000000042170	0.00202	-0.00561
NBP[0]672*	0.00000000002629	0.00231	-0.00644
NBP[0]722*	0.00000000000054	0.00162	-0.00452
NBP[0]769*	0.00000000000001	0.00167	-0.00465
		TOTAL	*-3.85905E-09*

Using the Solver add-on for Excel the value of β was adjusted such that the sum of all Rachford-Rich solutions equals zero. The solver parameters entered are as follows:



Solver Parameters

Set Objective: `$M$14`

To: ○ Max ○ Min ● Value Of: 0

By Changing Variable Cells: mL

Subject to the Constraints:
mL <= 1
mL >= 0

☑ Make Unconstrained Variables Non-Negative

Select a Solving Method: GRG Nonlinear

Solving Method
Select the GRG Nonlinear engine for Solver Problems that are smooth nonlinear. Select the LP Simplex engine for linear Solver Problems, and select the Evolutionary engine for Solver problems that are non-smooth.

Figure 13: Excel Solver

The answer report generated showed that at a value of β=1.4 satisfied the constraints. We know from mole balance that

Molar flow of Feed in (F)= Molar flow out as vapor (V)+ Molar flow out as Liquid (L)

since β=V/L, F = V + (V/ β)

Plugging in F=280.9 kmol/h, β=1.40 and solving for V yields

V=164 kmol/h and thus, L=116.9 kmol/h

Component flows (li and vi) are then obtained by vapor and liquid compositions:

$$For\ liquid\ composition, x_i = \frac{z_i}{1+\beta(K_i-1)}$$

$$For\ vapor\ composition, y_i = K_i x_i$$

Table 5: Preflash Separator

	HEATED OIL			**LIQUID FEED**		**VAPOR FEED**	
	z_i	$f_i = F. z_i$ (kmol/h)	K_i	x_i	$l_i = L. x_i$ (kmol/h)	y_i	$v_i = V. y_i$ (kmol/h)
Propane	**0.001**	**0.148**	**57.349**	**0.000**	**0.002**	**0.001**	**0.146**
i-Butane	**0.001**	0.176	33.011	0.000	0.003	0.001	0.173
n-Butane	**0.003**	0.737	27.142	0.000	0.017	0.004	0.721
i-Pentane	**0.003**	0.930	15.463	0.000	0.038	0.005	0.893
n-Pentane	**0.007**	2.092	13.454	0.001	0.098	0.012	1.998
H2O	**0.001**	0.158	157.726	0.000	0.001	0.001	0.157
NBP[0]49*	**0.010**	2.761	9.936	0.001	0.173	0.016	2.593
NBP[0]63*	**0.020**	5.526	7.505	0.004	0.449	0.031	5.086
NBP[0]77*	**0.337**	94.576	5.589	0.086	10.065	0.516	84.659
NBP[0]90*	**0.049**	13.690	4.052	0.017	1.935	0.072	11.774
NBP[0]104*	**0.056**	15.716	2.923	0.025	2.925	0.078	12.810
NBP[0]118*	**0.050**	14.064	2.118	0.029	3.379	0.065	10.698
NBP[0]133*	**0.042**	11.886	1.487	0.032	3.696	0.050	8.196
NBP[0]147*	**0.042**	11.852	1.044	0.040	4.643	0.044	7.210
NBP[0]161*	**0.037**	10.258	0.729	0.042	4.927	0.032	5.328
NBP[0]176*	**0.033**	9.141	0.498	0.045	5.260	0.024	3.874
NBP[0]190*	**0.029**	8.029	0.341	0.046	5.337	0.016	2.683
NBP[0]205*	**0.025**	6.898	0.227	0.044	5.162	0.011	1.724
NBP[0]220*	**0.027**	7.724	0.149	0.054	6.328	0.008	1.380
NBP[0]233*	**0.024**	6.744	0.101	0.050	5.867	0.005	0.862
NBP[0]248*	**0.016**	4.630	0.063	0.036	4.232	0.002	0.387
NBP[0]262*	**0.016**	4.411	0.039	0.036	4.162	0.001	0.237
NBP[0]277*	**0.021**	5.777	0.024	0.048	5.566	0.001	0.194
NBP[0]290*	**0.019**	5.447	0.015	0.045	5.314	0.001	0.118
NBP[0]305*	**0.014**	3.851	0.009	0.032	3.790	0.000	0.050
NBP[0]319*	**0.011**	3.203	0.005	0.027	3.169	0.000	0.024
NBP[0]334*	**0.010**	2.941	0.003	0.025	2.920	0.000	0.013
NBP[0]348*	**0.010**	2.796	0.002	0.024	2.781	0.000	0.007
NBP[0]362*	**0.009**	2.609	0.001	0.022	2.597	0.000	0.004
NBP[0]377*	**0.008**	2.317	0.001	0.020	2.308	0.000	0.002
NBP[0]391*	**0.008**	2.165	0.000	0.018	2.157	0.000	0.001
NBP[0]405*	**0.007**	1.939	0.000	0.017	1.933	0.000	0.000
NBP[0]420*	**0.006**	1.658	0.000	0.014	1.653	0.000	0.000
NBP[0]440*	**0.011**	3.066	0.000	0.026	3.057	0.000	0.000
NBP[0]468*	**0.008**	2.283	0.000	0.019	2.276	0.000	0.000

NBP[0]497*	**0.007**	1.868	0.000	0.016	1.863	0.000	0.000
NBP[0]523*	**0.006**	1.779	0.000	0.015	1.774	0.000	0.000
NBP[0]551*	**0.004**	1.146	0.000	0.010	1.143	0.000	0.000
NBP[0]580*	**0.003**	0.957	0.000	0.008	0.954	0.000	0.000
NBP[0]606*	**0.003**	0.815	0.000	0.007	0.813	0.000	0.000
NBP[0]634*	**0.002**	0.566	0.000	0.005	0.565	0.000	0.000
NBP[0]672*	**0.002**	0.650	0.000	0.006	0.648	0.000	0.000
NBP[0]722*	**0.002**	0.456	0.000	0.004	0.454	0.000	0.000
NBP[0]769*	**0.002**	0.469	0.000	0.004	0.468	0.000	0.000
TOTAL:	**1.000**	280.900	341.009	1.000	116.900	1.000	164.000

4.1.4 Stripper

To find the component distribution, the Hengstebeck-Geddes Equation was used. This required the knowledge of the relative volatilities of the components (courtesy of Aspen HYSYS). The constants A and C were solved for, and the results are tabulated as follows. The hypotheticals have been summed up to give factions in terms of carbon numbers (inferred from their boiling point data).

$$\log\left(\frac{d_i}{b_i}\right) = A + C\ log(\alpha_i)$$

Table 6: Stripper

	In			Out		
(kg/h)	Liquid feed	from rectifier	from ss1	vapToRect	HSDFO	Bottoms
Methane	**0.000**	**0.000**	**0.000**	**0.000**	**0.000**	**0.000**
Ethane	**0.000**	0.000	0.000	0.000	0.000	0.000
Propane	**0.074**	0.014	0.000	0.088	0.000	0.000
n-Butane	**1.016**	0.139	0.000	1.155	0.000	0.000
i-Butane	**0.200**	0.030	0.000	0.230	0.000	0.000
n-Pentane	**7.094**	0.724	0.001	7.818	0.001	0.000
i-Pentane	**2.756**	0.304	0.001	3.060	0.001	0.000
salts	**0.027**	0.000	0.000	0.000	0.008	0.019
water	**0.012**	0.003	0.000	0.015	0.000	0.000
C4-C5	**1769.252**	124.329	0.696	1893.581	0.697	0.000
C6-C11	**3142.084**	188.822	6.156	3330.604	6.458	0.000
C12-C16	**8651.447**	1306.433	1680.156	6155.922	5324	158
C17-C21	**3773.140**	105.768	117.953	102.844	1606	2287
C22-C28	**7389.863**	4.226	9.461	3.584	2134	5265
Column Total=	**24736.965**	1730.791	1814.424	11498.900	9072	7711
Total	**28282.181**			**28282.181**		

4.1.5 Mixer 2

From law of conservation of mass, we get:

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

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

Since $\dot{m}_1 = 16090\frac{kg}{h} and\ \dot{m}_2 = 11500\frac{kg}{h}$, $\dot{m}_3 = \dot{m}_1 + m_2^{\cdot} = 27590\frac{kg}{h}$

Results are tabulated as follows:

Table 7: Mixer 2

	VAPOR FEED		**VAPTORECT**		**SATURATED MIXTURE**	
	y_i	Mass flow (kg/h)	y_i	Mass flow (kg/h)	y_i	Mass flow (kg/h)
Methane	0.00	**0.00**	**0.00**	**0.00**	**0.00**	**0.00**
Ethane	0.00	0.00	0.00	0.00	0.00	0.00
Propane	0.00	14.35	0.00	0.29	0.00	16.78
i-Butane	0.00	16.93	0.00	0.57	0.00	19.98
n-Butane	0.00	70.70	0.00	2.87	0.00	83.83
i-Pentane	0.01	87.65	0.00	6.13	0.00	105.94
n-Pentane	0.01	196.02	0.00	15.66	0.01	238.46
H2O	0.00	15.44	0.00	0.12	0.00	17.90
NACL	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]49*	0.02	254.38	0.00	27.31	0.01	314.94
NBP[0]63*	0.03	498.99	0.01	70.54	0.02	631.09
NBP[0]77*	0.52	8305.84	0.14	1569.50	0.39	10815.00
NBP[0]90*	0.07	1155.13	0.03	299.91	0.06	1568.16
NBP[0]104*	0.08	1256.77	0.04	451.13	0.07	1804.11
NBP[0]118*	0.07	1049.54	0.05	519.21	0.06	1618.45
NBP[0]133*	0.05	804.09	0.05	566.39	0.05	1372.27
NBP[0]147*	0.04	707.35	0.06	710.86	0.05	1374.04
NBP[0]161*	0.03	522.73	0.07	754.92	0.04	1195.61
NBP[0]176*	0.02	380.08	0.07	808.42	0.04	1072.99
NBP[0]190*	0.02	263.21	0.07	824.98	0.03	951.16
NBP[0]205*	0.01	169.11	0.07	805.56	0.03	827.34
NBP[0]220*	0.01	135.41	0.09	998.70	0.03	940.04
NBP[0]233*	0.01	84.54	0.08	926.66	0.03	824.81
NBP[0]248*	0.00	37.98	0.05	629.87	0.02	538.17
NBP[0]262*	0.00	23.24	0.04	504.30	0.02	422.60
NBP[0]277*	0.00	19.06	0.04	464.38	0.01	386.45
NBP[0]290*	0.00	11.56	0.03	294.88	0.01	244.76
NBP[0]305*	0.00	4.87	0.01	129.55	0.00	107.30
NBP[0]319*	0.00	2.38	0.01	62.62	0.00	51.89

NBP[0]334*	0.00	1.26	0.00	28.93	0.00	24.16
NBP[0]348*	0.00	0.68	0.00	13.28	0.00	11.21
NBP[0]362*	0.00	0.35	0.00	6.37	0.00	5.41
NBP[0]377*	0.00	0.17	0.00	3.04	0.00	2.59
NBP[0]391*	0.00	0.08	0.00	1.55	0.00	1.32
NBP[0]405*	0.00	0.04	0.00	0.77	0.00	0.65
NBP[0]420*	0.00	0.02	0.00	0.36	0.00	0.30
NBP[0]440*	0.00	0.01	0.00	0.28	0.00	0.23
NBP[0]468*	0.00	0.00	0.00	0.06	0.00	0.05
NBP[0]497*	0.00	0.00	0.00	0.02	0.00	0.01
NBP[0]523*	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]551*	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]580*	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]606*	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]634*	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]672*	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]722*	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]769*	0.00	0.00	0.00	0.00	0.00	0.00
Total	1.00	16090.00	1.00	11500.00	1.00	27590.00

4.1.6 Rectifier

The balance was carried out in a manner similar to that of the stripper.

Table 8: Rectifier

	In			Out				
(kg/h)	Compressed vapors	PA 3	from ss2	Overhead	Side product 1	To Strip1	PA 1	Naphtha
Methane	0	**0**	**0**	**0**	**0**	**0**	**0**	**0**
Ethane	0	0	0	0	0	0	0	0
Propane	7	0	0	6	0	0	0	0
n-Butane	43	1	1	41	1	0	1	2
i-Butane	10	0	0	10	0	0	0	0
n-Pentane	152	3	3	141	3	1	3	10
i-Pentane	67	1	1	63	1	0	1	4
salts	0	0	0	0	0	0	0	0
water	3	0	0	3	0	0	0	0
C4-C5	13441	521	462	9060	580	124	521	4138
C6-C11	6757	792	539	59	1079	189	792	5969
C12-C16	7022	5481	1000	0	5002	1308	5478	1715
C17-C21	110	441	0	0	1	106	443	0
C22-C28	4	18	0	0	0	4	18	0
Total=	27615	7258	2006	9383	6668	1732	7257	11839
Total in=	36880	36880						

4.1.7 Flash Drum

Similar to the pre-flash separator, the Rachford-Rice solutions for the feed composition were summated and then brought to 0 using the solver add-on in MS Excel. The constraints were similar to last time as well ($0 \leq \beta \leq 1$). The answer report generated gave a value of β=0.1962.

Molar flow of Feed in (F)= Molar flow out as vapor (V)+ Molar flow out as Liquid (L)

since β=V/L, F = V + (V/ β)

Plugging in F=106.4 kmol/h, β=0.1962 and solving for V yields

V=17.45 kmol/h and thus, L=88.95 kmol/h

Component flows (li and vi) are then obtained by vapor and liquid compositions:

$$For\ liquid\ composition, x_i = \frac{z_i}{1 + \beta(K_i - 1)}$$

$$For\ vapor\ composition, y_i = K_i x_i$$

Table 9: Flash Drum

	OVERHEAD SATURATED			**LIGHT NAPHTHA**		**OVERHEAD GASES**	
	z_i	$f_i = F. z_i$ (kmol/h)	K_i	x_i	$l_i = L. x_i$ (kmol/h)	y_i	$v_i = V. y_i$ (kmol/h)
Propane	0.00	**0.14**	**22.36**	**0.00**	**0.02**	**0.01**	**0.12**
i-Butane	0.00	0.17	10.47	0.00	0.05	0.01	0.12
n-Butane	0.01	0.71	7.93	0.00	0.25	0.03	0.46
i-Pentane	0.01	0.87	3.71	0.01	0.48	0.02	0.39
n-Pentane	0.02	1.95	2.98	0.01	1.19	0.04	0.76
H2O	0.00	0.16	111.16	0.00	0.01	0.01	0.15
NBP[0]49*	0.02	2.51	2.06	0.02	1.75	0.04	0.76
NBP[0]63*	0.05	4.84	1.37	0.04	3.77	0.06	1.07
NBP[0]77*	0.72	76.86	0.92	0.73	64.80	0.69	12.05
NBP[0]90*	0.09	9.20	0.61	0.09	8.21	0.06	0.99
NBP[0]104*	0.06	6.37	0.40	0.07	5.92	0.03	0.45
NBP[0]118*	0.02	2.11	0.26	0.02	2.01	0.01	0.10
NBP[0]133*	0.00	0.41	0.16	0.00	0.40	0.00	0.01
NBP[0]147*	0.00	0.08	0.10	0.00	0.08	0.00	0.00
NBP[0]161*	0.00	0.01	0.06	0.00	0.01	0.00	0.00
NBP[0]176*	0.00	0.00	0.04	0.00	0.00	0.00	0.00

NBP[0]190*	0.00	0.00	0.02	0.00	0.00	0.00	0.00
NBP[0]205*	0.00	0.00	0.01	0.00	0.00	0.00	0.00
NBP[0]220*	0.00	0.00	0.01	0.00	0.00	0.00	0.00
NBP[0]233*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]248*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]262*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]277*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]290*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]305*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]319*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]334*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]348*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]362*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]377*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]391*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]405*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]420*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]440*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]468*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]497*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]523*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]551*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]580*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]606*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]634*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]672*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]722*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NBP[0]769*	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL:	1.00	106.40	164.62	1.00	88.95	1.00	17.45

CHAPTER 5

Energy Balance

5.1 ENERGY BALANCE

Once the required material balances were carried out, we were able to conduct energy balances on appropriate equipment. Since the proposed process is only a series of unit operations, it was only a matter of using the energies of the streams themselves.

5.1.1 PUMPS

Two main pumps were utilized for this process:

Crude oil feed pump,

Cold water supply pump to the product heat exchangers network.

The equations required were:

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

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

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

Required data and results tabulated as follows:

Table 10: Pumps

	Pump	
Property	1	2
Temperature (^oC)	25.0	25.0
Specific Enthalpy (kJ/kg)	2195	15888
Inlet Pressure (kPa)	101.3	100
Outlet Pressure (kPa)	200	200
Density (kg/m^3)	772.7	1007
Mass Flow Rate (kg/h)	51800	124200
Pump Efficiency	0.75	0.75
Pump Duty (kW)	2450	4570

5.1.2 COMPRESSOR

There is only one main compressor within the proposed process, and it is responsible for pressurizing the vapors to the bottom of the rectifying section.

The equations required were:

$$W(isentropic) = \frac{\gamma R T_1}{\gamma - 1}\left[\left(\frac{P_2}{P_1}\right)^{\frac{\gamma-1}{\gamma}} - 1\right]$$

$$Compressor\ Duty = \frac{W(isentropic)}{\eta}$$

Required data and results tabulated as follows:

Table 11:Compressor

Parameter	Inlet (heated vapors)	Outlet (ToRect)
Temperature (^{o}C)	200	290.10
Pressure (kPa)	80	600
C_p (kJ/kmol.^{o}C)	258.74	301.96
Mass flow (kg/h)	60831.38	
Efficiency,	0.75	
Enthalpy (kJ/kg)	1288.25	1496.11
Pump Duty (kW)	1593.11	

5.1.3 HEAT EXCHANGERS

Energy balance was used to work out the heat loads involved in each heat exchanger employed for the proposed process.

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

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

Table 12:Hx-1

		Shell side		Tube side	
	units	Pumped oil	warm oil	steam 2	steam 3
Vapor Fraction		0.000	0.312	1.000	1.000
Temperature	*°C*	25.034	110.000	180.000	120.000
Pressure	*bar*	2.000	1.400	1.500	1.100
Mass Flow	*lb/hr*	1.14E+05	1.14E+05	2.43E+05	2.43E+05
Heat Flow	*kJ/h*	-1.14E+08	-1.01E+08	-1.45E+09	-1.46E+09
Mass Heat Capacity	*kJ/kg-°C*	1.903	2.195	1.948	1.916
Q	*kW*	-3514.876		3514.876	

Table 13:Hx-2

		Shell side		Tube side	
	units	Desalted oil	heated oil	steam 1	steam 2
Vapor Fraction		0.000	0.569	1.000	1.000
Temperature	*°C*	107.456	150.000	220.234	180.000
Pressure	*bar*	1.400	1.100	1.600	1.500
Mass Flow	*lb/hr*	9.00E+04	9.00E+04	2.43E+05	2.43E+05
Heat Flow	*kJ/h*	-8.18E+07	-7.31E+07	-1.44E+09	-1.45E+09
Mass Heat Capacity	*kJ/kg-°C*	2.237	2.239	1.969	1.948
Q	*kW*	-2408.409		2408.409	

Table 14: Hx-3

		Shell side		Tube side	
	units	water1	water 2	Overhead	overhead saturated
Vapor Fraction		0.000	0.569	1.000	1.000
Temperature	*°C*	107.456	150.000	220.234	180.000
Pressure	*bar*	1.400	1.100	1.600	1.500
Mass Flow	*lb/hr*	9.00E+04	9.00E+04	2.43E+05	2.43E+05
Heat Flow	*kJ/h*	-8.18E+07	-7.31E+07	-1.44E+09	-1.45E+09
Mass Heat Capacity	*kJ/kg-°C*	2.237	2.239	1.969	1.948
Q	*kW*	-1150.995		1150.995	

Table 15: Hx-4

		Shell side		**Tube side**	
	units	water 2	water 3	Naphtha	straight run Gasoline
Vapor Fraction		0.000	0.000	0.050	0.000
Temperature	*°C*	32.736	40.740	197.055	50.000
Pressure	*bar*	1.950	1.550	5.325	4.725
Mass Flow	*lb/hr*	2.74E+05	2.74E+05	2.61E+04	2.61E+04
Heat Flow	*kJ/h*	-1.97E+09	-1.96E+09	-2.09E+07	-2.52E+07
Mass Heat Capacity	*kJ/kg-°C*	4.314	4.316	2.761	2.012
Q	*kW*	-1189.945		1189.945	

Table 16: Hx-5

		Shell side		**Tube side**	
	units	water 3	water 4	hot kero oil	Kerosene
Vapor Fraction		0.000	0.000	0.000	0.000
Temperature	*°C*	40.740	43.857	203.746	45.000
Pressure	*bar*	1.550	1.900	1.200	1.250
Mass Flow	*lb/hr*	2.74E+05	2.74E+05	1.03E+04	1.03E+04
Heat Flow	*kJ/h*	-1.96E+09	-1.96E+09	-8.21E+06	-9.89E+06
Mass Heat Capacity	*kJ/kg-°C*	4.316	4.317	2.616	1.909
Q	*kW*	-465.296		465.296	

Table 17; Hx-6

		Shell side		**Tube side**	
	units	water 4	water 5	HS diesel	Diesel
Vapor Fraction		0.00	0.00	0.00	0.00
Temperature	*°C*	43.86	50.00	306.74	127.02
Pressure	*bar*	1.90	1.80	1.00	0.90
Mass Flow	*lb/hr*	2.74E+05	2.74E+05	1.60E+04	1.60E+04
Heat Flow	*kJ/h*	-1.96E+09	-1.96E+09	-1.09E+07	-1.42E+07
Mass Heat Capacity	*kJ/kg-°C*	4.32	4.32	2.86	2.18
Q	*kW*	-915.03		915.03	

Table 18: Hx-7

		Shell side		**Tube side**	
	units	water 5	water 6	Bottoms	Furnace fuel oil
Vapor Fraction		0.000	0.000	0.000	0.000
Temperature	*°C*	50.000	58.400	313.815	63.270
Pressure	*bar*	1.800	1.800	0.200	1.050
Mass Flow	*lb/hr*	2.74E+05	2.74E+05	1.70E+04	1.70E+04
Heat Flow	*kJ/h*	-1.96E+09	-1.96E+09	-1.16E+07	-1.61E+07
Mass Heat Capacity	*kJ/kg-°C*	4.321	4.328	2.764	1.874
Q	*kW*	-1253.116		1253.116	

CHAPTER 6

Equipment Design

6.1 EQUIPMENT DESIGN

This section includes the design of the major components within the proposed process. There are 5 key equipment designs tackled:

Desalter

Pre-flash Separator

HIDC System

Flash Drum

Heat Exchangers

The Desalter and Flash drum were designed using the same set of equations since both of the are vessels demonstrating 3-phase separation. Pre-flash separator was designed with a slightly different methodology since it involves the separation of only 2 phases. The HIDC system was designed by reducing the multicomponent system to a pseudo-binary one and carrying out necessary calculations afterwards.

6.1.1 Desalter

The desalter is assumed to be a horizontal vessel. Material balance has proved that there is a 3-phase separation taking place within; the 3 phases being the desalted oil, the used-up wash-water and the gases escaping from the top. The diameter of the vessel would thus be a function of the vapor flow rate while the length of the vessel would depend on the allowable level of liquid.

Superficial velocity of the gas is given by the Sourdes-Brown equation:

$$u_t = 0.07 \sqrt{\frac{\rho_l - \rho_v}{\rho_v}}$$

Where ρv= 3.859x10-3 g/cm3, ρl = 17.8x10-3 g/cm3 (obtained from Desalter Workbook in HYSYS). The constant 0.07 is the vapor velocity factor chosen from literature.

Substituting given values yields:

$$u_t = 0.07\sqrt{\frac{17.8-3.86}{3.86}} = 0.133\frac{m}{s} = 13.3\frac{cm}{s}$$

The next step is to calculate the drag coefficient of the dispersed phase particles within the flowing gas. This is done by solving a balance between the drag and buoyant forces acting on a single particle:

Coefficient of drag,

$$C_D\rho_g.\frac{\pi(d_m)^2}{4}.\frac{(u_t)^2}{2} = \left|\rho_g-\rho_l\right|.g.\frac{\pi(d_m)^3}{6}$$

Assuming the diameter of the settling particles to be 200 microns, the following data will then be plugged in to solve for Cd:

ut=13.3 cm/s (as calculated earlier)

ρg= 3.859x10-3 g/cm3, ρl = 17.8x10-3 g/cm3

Qg= 7.23 MMscfd (from material balance and HYSYS)

$$\xRightarrow{solve\ for\ C_D} C_D = \frac{1}{3.859\text{x}10^{-3}.\frac{\pi(0.02)^2}{4}.\frac{(13.3)^2}{2}}.(17.8$$
$$-3.86)(981).\frac{\pi(0.02)^3}{6}$$
$$C_D = 0.54$$

Design equation for a 3-phase separator is:

$$\frac{L_{eff}d^2F_g}{h_g} = 421\left(\frac{TZQ_g}{P}\right)\sqrt{\frac{\rho_v}{\rho_l-\rho_v}.\frac{C_D}{d_m}}$$

To simplify the equation the following assumptions are made:

Length of the vessel is 3 times its diameter, Leff=3d,

The liquid pool takes up at least half the space (Fg), and

The height of the liquid pool takes up almost, hg=0.5d

Other values to be used include T= 685.1 R, Z=0.9591, and P=20.31 psia.

Substituting these values allows us to solve for diameter very easily:

$$d^2 = 421\left(\frac{TZQ_g}{P}\right)\sqrt{\frac{\rho_v}{\rho_l - \rho_v}.\frac{C_D}{d_m}}$$

$$3d^2 = 421\left(\frac{685.1(0.9591)(7.23)}{20.31}\right)\sqrt{\frac{0.241}{1.111 - 0.241}.\frac{0.542}{200}}$$

$$3d^2 = 2698$$

Thus, diameter of vessel= 30 inches ≈ 0.762 meters

Consequently, Length of vessel, Leff = 3d = 2.286 meters.

The results are tabulated as follows:

Table 19: Desalter

Superficial Gas Velocity, u_t (cm/s)	13.3
Coefficient of Drag	0.54
Diameter of Horizontal vessel, d (m)	0.762
Length of the Horizontal Vessel, L_{eff} (m)	2.286

6.1.2 Pre-Flash Separator

This unit comes before the stripping section of the HIDC system. The governing phenomena is a 2-phase separation between liquid and vapor phase. The superficial gas velocity will be obtained in a similar manner to previous design, but the dimensions of the vessel will be approached differently.

Superficial velocity of the gas is given by the Sourdes-Brown equation:

$$u_t = 0.07\sqrt{\frac{\rho_l - \rho_v}{\rho_v}}$$

Where ρl= 771.8 kg.m-3, ρv= 2.890 kg.m-3(obtained from Desalter Workbook in HYSYS). The constant 0.07 is the vapor velocity factor chosen from literature.

Substituting given values yields:

$$u_t = 0.07\sqrt{\frac{771.8 - 2.89}{2.89}} = 1.1418\, m/s$$

The diameter of the vessel is given by:

$$D_v = \sqrt{\frac{4V_v}{\pi u_t}}$$

From Rachford-Rice equation and density,

Vv= 1.569 m3s-1, Vl= 0.0089 m3s-1

Substituting necessary values:

$$D_v = \sqrt{\frac{4(1.569)}{\pi(1.1418)}} \approx 1.323\, meters$$

For the total height of the column, we are going to use the volume of the liquid pool in the vessel. For a holdup time of 10 minutes:

$$V_{pool} = V_l\, .\, (holdup\ time\ allowed)$$

$$V_{pool} = 0.0089\, .\, (10 * 60)$$

Thus, the volume of liquid holdup is calculated to be 5.34 m3.

Standard correlations, courtesy of Coulson & Richardson Volume 6, give us the following heights:

Height of liquid pool:

$$h_l = \frac{V_{pool}}{\pi {D_v}^2/4}$$

For Dv=1.323 m and Vpool=5.34 m3,

$$h_l = \frac{5.34}{\pi(1.323)^2/4} = 5.303\, m$$

Height of column above inlet nozzle:

$$h_v = D_v + 0.4\ meters$$

$$h_v = 1.323 + 0.4 = 1.72\ m$$

Height of inlet nozzle from liquid pool surface:

$$h_f = D_v/2$$

$$h_f = \frac{1.323}{2} = 0.662m$$

The total height of the column:

$$L = h_l + h_v + h_f$$

$$L = 5.303 + 1.72 + 0.662 = 7.685\ meters$$

The results are tabulated as follows:

Table 20: Pre-Flash Separator

Superficial Gas Velocity, u_t (m/s)	1.1418
Height of liquid pool, h_l (m)	5.303
Diameter of Vertical vessel, d (m)	1.323
Height of the Vertical Vessel, L (m)	7.685

6.1.3 Flash Drum

The flash drum downstream of the rectifier was designed using the same method as the one used for the desalter. The flash drum is used to cool down the top products of the rectifier so that the gases and light naphtha may be separated better. It also acts as a way of removal of any residual moisture.

Superficial velocity of the gas is given by the Sourdes-Brown equation:

$$u_t = 0.07\sqrt{\frac{\rho_l - \rho_v}{\rho_v}}$$

Where ρv= 0.3098x10-3 g/cm3, ρl=31.22x10-3 g/cm3 (obtained from Desalter Workbook in HYSYS). The constant 0.07 is the vapor velocity factor chosen from literature.

Substituting given values yields:

$$u_t = 0.07\sqrt{\frac{\rho_l - \rho_v}{\rho_v}} = 0.07\sqrt{\frac{17.8 - 3.86}{3.86}} = 0.699\frac{m}{s} = 69.9\frac{cm}{s}$$

The next step is to calculate the drag coefficient of the dispersed phase particles within the flowing gas. This is done by solving a balance between the drag and buoyant forces acting on a single particle:

Coefficient of drag,

$$C_D\rho_g.\frac{\pi(d_m)^2}{4}.\frac{(u_t)^2}{2} = \left|\rho_g - \rho_l\right|.g.\frac{\pi(d_m)^3}{6}$$

Assuming the diameter of the settling particles to be 200 microns, the following data will then be plugged in to solve for Cd:

ut=13.3 cm/s (as calculated earlier)

ρg= 0.3098x10-3 g/cm3, ρl=31.22x10-3 g/cm3

Qg= 9.34 MMscfd (from material balance and HYSYS)

$$C_D = \frac{1}{3.859\text{x}10^{-3}.\frac{\pi(0.02)^2}{4}.\frac{(13.3)^2}{2}}.(17.8 - 3.86)(981).\frac{\pi(0.02)^3}{6}$$

$$C_D = 0.55$$

Design equation for a 3-phase separator is:

$$\frac{L_{eff}d^2F_g}{h_g} = 421\left(\frac{TZQ_g}{P}\right)\sqrt{\frac{\rho_v}{\rho_l - \rho_v}.\frac{C_D}{d_m}}$$

To simplify the equation the following assumptions are made:

Length of the vessel is 3 times its diameter, Leff=3d,

The liquid pool takes up at least half the space (Fg), and

The height of the liquid pool takes up almost, hg=0.5d

Other values to be used include T= 617.1 R, Z=0.9957, and P=13.05 psia,

Substituting these values allows us to solve for diameter very easily:

$$d^2 = 421\left(\frac{TZQ_g}{P}\right)\sqrt{\frac{\rho_v}{\rho_l - \rho_v}.\frac{C_D}{d_m}}$$

$$d^2 = 421\left(\frac{617.1(0.9957)(9.34)}{13.05}\right)\sqrt{\frac{0.01934}{1.949 - 0.01934}.\frac{0.55}{200}}$$

$$3d^2 = 971.98$$

Thus, diameter of vessel= 18 inches ≈ 0.4572 meters

Consequently, Length of vessel, Leff = 3d = 1.372 meters.

The results are tabulated as follows:

Table 21: Flash Drum

Superficial Gas Velocity, u_t (cm/s)	69.9
Coefficient of Drag	0.55
Diameter of Horizontal vessel, d (m)	0.457
Length of the Horizontal Vessel, L_{eff} (m)	1.372

6.1.4 HIDC System

The HIDC System design was approached as a simple multicomponent distillation column design. First of all, the system was reduced to a pseudo-binary form by picking the high -key and low-key components respectively. Once they were chosen and their tops and bottoms compositions were fixed, the temperature at the column ends were to be 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:

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

The values of composition and K for every component were obtained easily through HYSYS workbooks for the process streams. The value of temperature that satisfied this relation came out to be 143°C.

For the column bottom temperature, bubble point calculations were carried out at different temperatures. For 313 °C, the following relation was satisfied:

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

Once these temperatures have been calculated, the minimum number of stages were to be calculated. This was achieved using Fenske's Equation for minimum number of plates required for given distillate and bottoms compositions:

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

Substituting the required values:

The relative volatility values of the components were then used to obtain the minimum reflux ratio required through the Erbar-Maddox approach:

6.1.5 Heat Exchanger designing:

Table 22: Hx-1

Crude oil	Steam
$\boldsymbol{K = 0.1121\ \frac{W}{m\,°C}}$	$K = 0.03144\frac{W}{m\,°C}$
$\boldsymbol{\mu = 1.236e-3\ \frac{Ns}{m^2}}$	$\mu = 1.53e-5\frac{Ns}{m^2}$
$\boldsymbol{c_p = 1.903\frac{kJ}{kg\,°C}}$	$c_p = 1.948\frac{kJ}{kg\,°C}$
$\boldsymbol{\rho = 772.7\frac{kg}{m^3}}$	$\rho = 0.7227\frac{kg}{m^3}$
$\boldsymbol{\dot{m} = 5.18e4\ kg/s}$	$\dot{m} = 1.1e5\ kg/s$
$\boldsymbol{T_h = 110}$ °C	$T_h' = 180$ °C
$\boldsymbol{T_c = 25.03}$ °C	$T_c' = 120$ °C

Exchanger Duty:

$Q = \dot{m}c_p\Delta T$ = 1.265e7 kJ/h

Water flow rate:

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

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

$\dot{m} = 1.1\text{e}5$ kg/h

Logarithmic mean of Temperature Difference (LMTD):

$$LMTD = \frac{\theta_2 - \theta_1}{\ln(\frac{\theta_2}{\theta_1})} = 81.849\ ^\circ\text{C}$$

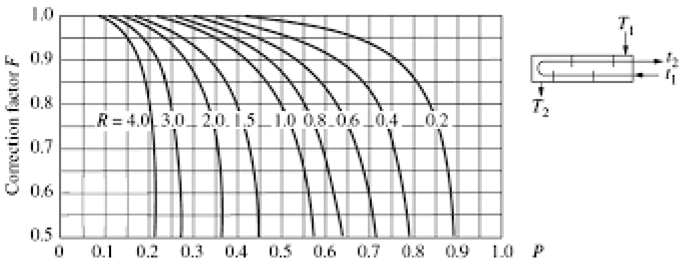
Where $\theta = T_h - T_c$

$$R = \frac{T_h' - T_c'}{t_c - t_h} = 1$$

$$P = \frac{t_c - t_h}{T_h' - t_c} = 0.59$$

Shell side 1 pass

Tube side double pass



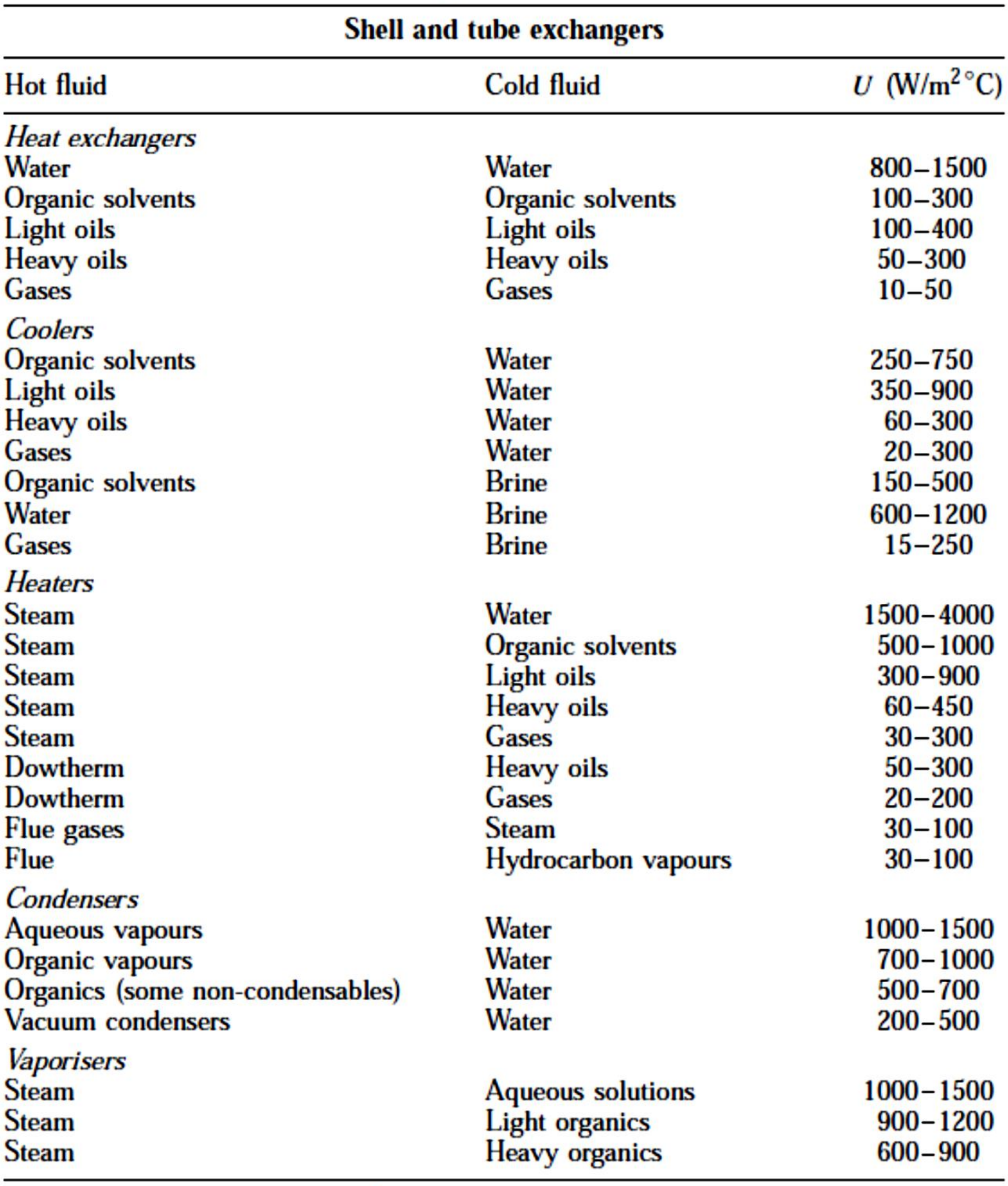
(a) One-shell pass and 2, 4, 6, etc. (any multiple of 2), tube passes

Figure 14: Ft factor

$F_t = 0.856$

So, Corrected temperature difference Δt =70.06 °C

Table 23: overall heat transfer coefficient



Shell and tube exchangers		
Hot fluid	Cold fluid	U (W/m^2°C)
Heat exchangers		
Water	Water	800–1500
Organic solvents	Organic solvents	100–300
Light oils	Light oils	100–400
Heavy oils	Heavy oils	50–300
Gases	Gases	10–50
Coolers		
Organic solvents	Water	250–750
Light oils	Water	350–900
Heavy oils	Water	60–300
Gases	Water	20–300
Organic solvents	Brine	150–500
Water	Brine	600–1200
Gases	Brine	15–250
Heaters		
Steam	Water	1500–4000
Steam	Organic solvents	500–1000
Steam	Light oils	300–900
Steam	Heavy oils	60–450
Steam	Gases	30–300
Dowtherm	Heavy oils	50–300
Dowtherm	Gases	20–200
Flue gases	Steam	30–100
Flue	Hydrocarbon vapours	30–100
Condensers		
Aqueous vapours	Water	1000–1500
Organic vapours	Water	700–1000
Organics (some non-condensables)	Water	500–700
Vacuum condensers	Water	200–500
Vaporisers		
Steam	Aqueous solutions	1000–1500
Steam	Light organics	900–1200
Steam	Heavy organics	600–900

$U = 2994\ \frac{W}{m^2 °C}$ from table in Supra

Area:

$$A = \frac{Q}{U\Delta t}$$

$$A = 60.32\ m^2$$

BWG-17: 0.01 m outer-side radius and 0.075 m inside radius

Length of tube $= l = 3.66\ m$

Area of single tube $= a = \pi(l)(do) = \ \pi \times 3.66 \times 0.03 = 0.334\ m^2$

Number of tubes $= n = \frac{A}{a} = 241\ tubes$

Tube bundle diameter = 0.32 m

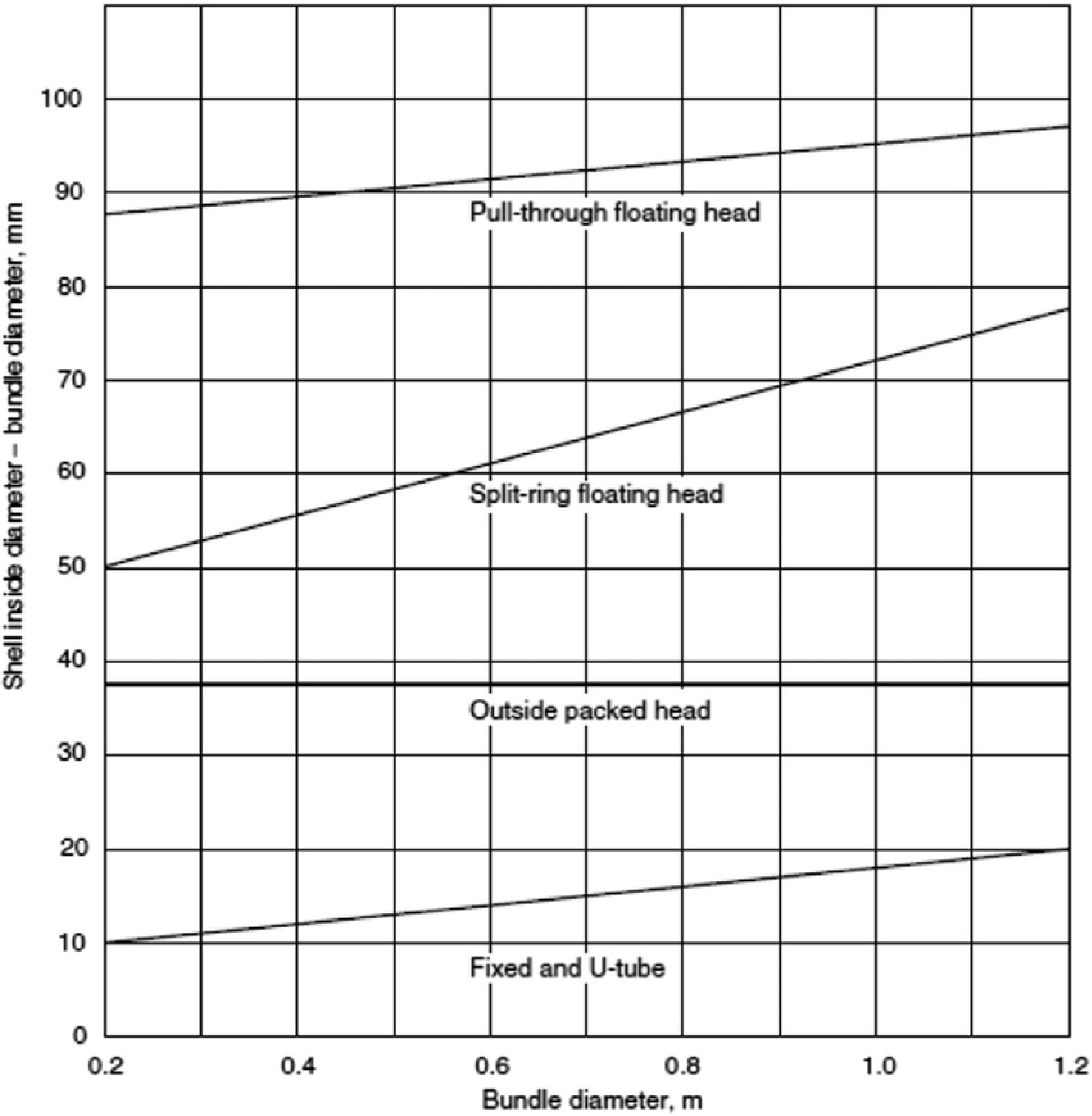
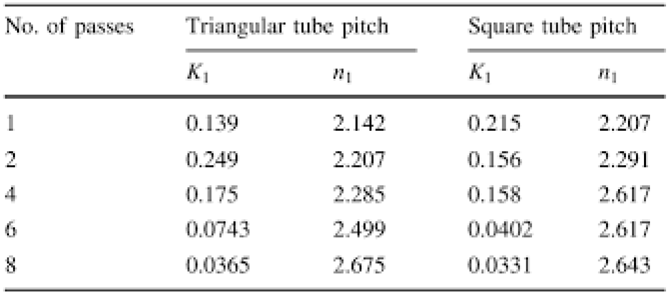


Figure 15: Bundle diameter

Clearance = 0.0093 m

Table 24: Pitch



No. of passes	Triangular tube pitch		Square tube pitch	
	K_1	n_1	K_1	n_1
1	0.139	2.142	0.215	2.207
2	0.249	2.207	0.156	2.291
4	0.175	2.285	0.158	2.617
6	0.0743	2.499	0.0402	2.617
8	0.0365	2.675	0.0331	2.643

Shell side co-efficient:

Baffle spacing (Ib) = $0.5 \times ds = 0.46557$

Pt = 0.0009 m

As = $\frac{(P_t - P_o) \times D_s}{P_t} = 60.32$ m2

Mass velocity 'Ga' =$\frac{\omega}{\text{As}} = 1.2\text{e}5 \, \frac{kg}{m^2 s}$

$$D_e = \frac{1.10}{d_o}\left(P_t^2 - 0.917{d_o}^2\right)$$

Re = 1539

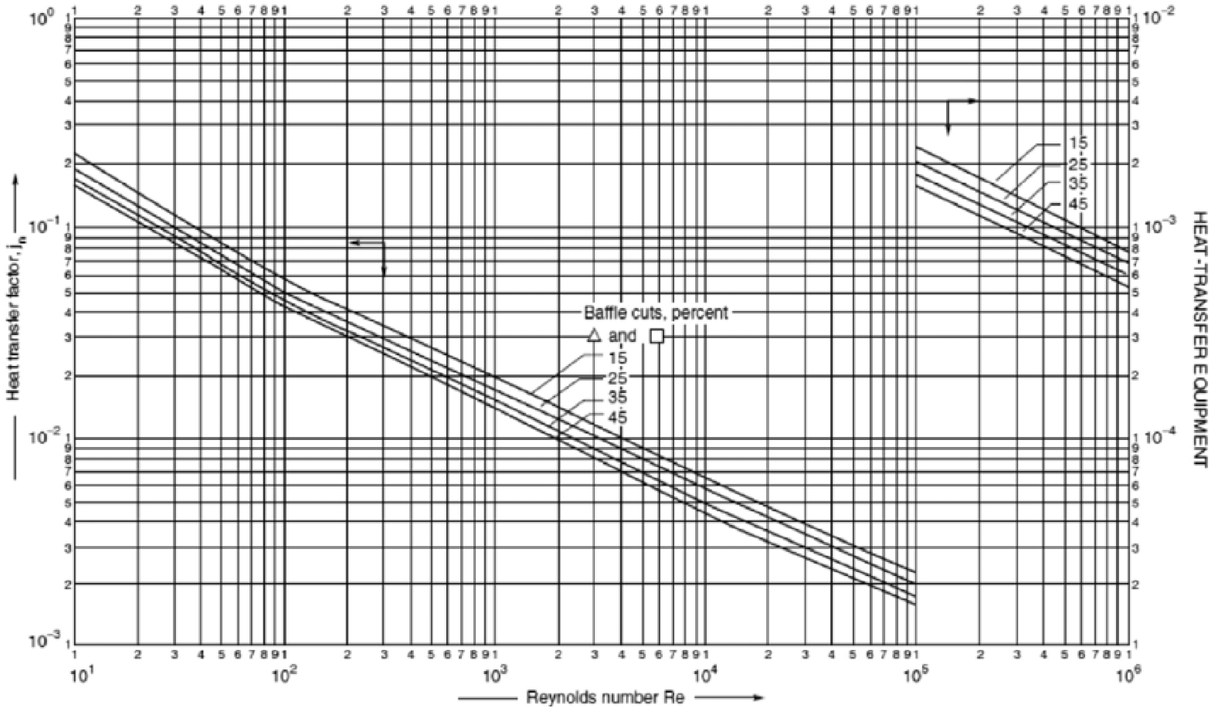


Figure 16: Heat transfer coefficient

By using figure above $J_h = 0.12$

$$Pr = \frac{\mu \times c_p}{K}$$

Pr = 1

$$\frac{h_o}{\theta_s} = \frac{J_h\,(Re)(Pr)^{0.33}(K_f)}{De} =$$

$h_o = 0.242\,\frac{W}{m^2\,℃}$

Tube side Coefficient:

Cross section area of tube $= \frac{\pi}{4}\,d_i^2 = 0.21\,m^2$

Number of tubes per pass n = 241

Overall area = 60. 32 m^2

Mass velocity 'Ga' = $\frac{\omega}{\mathrm{As}} = 1\text{e}4\,\frac{kg}{m^2 s}$

$$Re = \frac{D_i \times Ga}{\mu} = 35500$$

$$Pr = \frac{\mu \times c_p}{K} = 0.1$$

L/D=0.0005

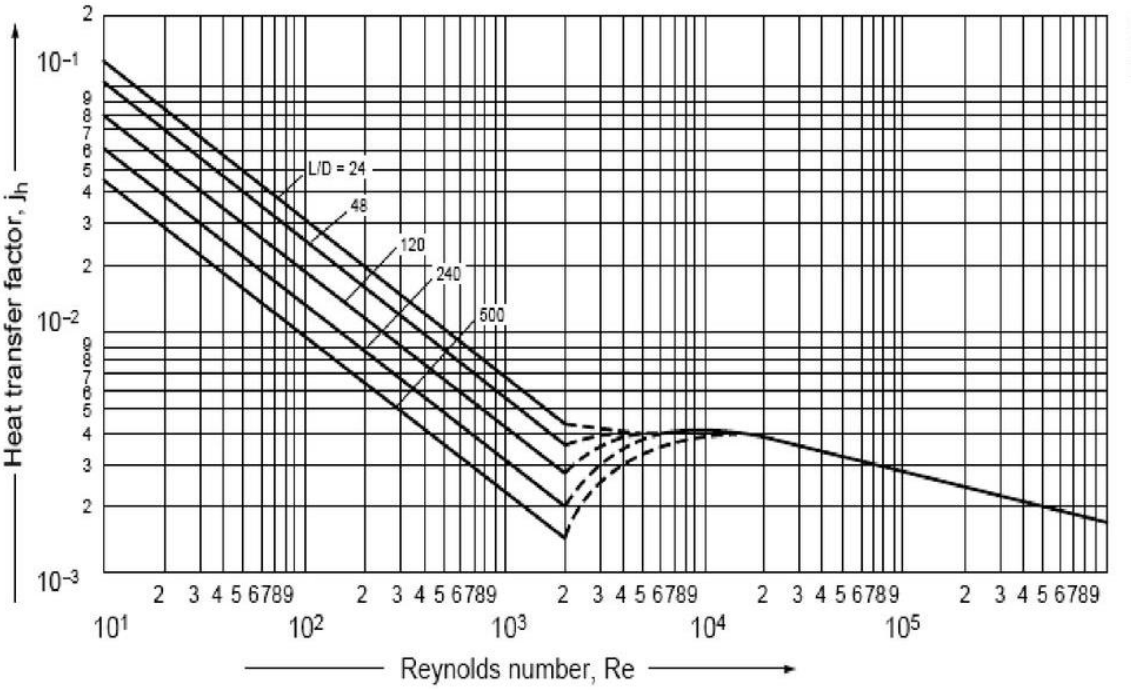


Figure 17: Heat transfer factor

$$\frac{h_i}{\theta_t} = \frac{J_h\,(Re)(Pr)^{0.33}(K_f)}{De} = 0.2$$

$\frac{h_o}{\theta_t} = 0.2\,\frac{W}{m^2\,°C}$

Mean Wall temperature θ_t:

$\left(\frac{\mu}{\mu_w}\right)^{0.14}$=1

$h_i = \quad \frac{W}{m^2\,°C}$

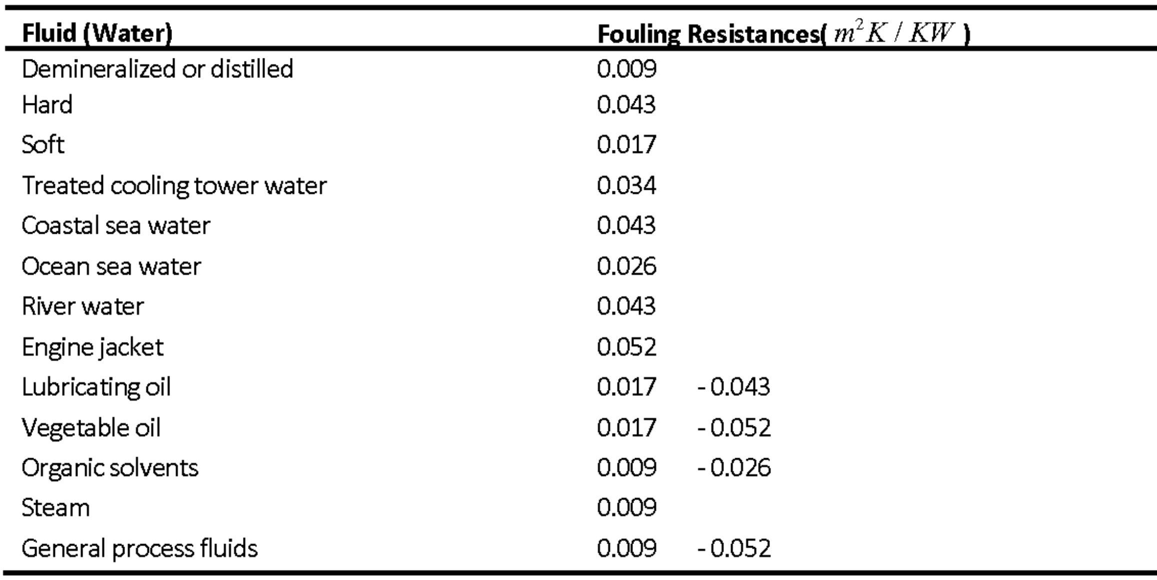
Overall convective heat transfer Coefficient:

$h_{io} = h_i \times \frac{ID}{OD} = 0.242 \frac{W}{m^2\,°C}$

$U_c = \frac{h_{io} \times h_o}{h_{io} + h_o} = 2994 \frac{W}{m^2\,°C}$

For Dirt factor Rd =0.003

Table 25: Fouling resistance



Fluid (Water)	Fouling Resistances(m^2K/KW)
Demineralized or distilled	0.009
Hard	0.043
Soft	0.017
Treated cooling tower water	0.034
Coastal sea water	0.043
Ocean sea water	0.026
River water	0.043
Engine jacket	0.052
Lubricating oil	0.017 - 0.043
Vegetable oil	0.017 - 0.052
Organic solvents	0.009 - 0.026
Steam	0.009
General process fluids	0.009 - 0.052

$$R_D = \frac{U_c - U_D}{U_c \times U_D}$$

$U_D = 2800 \frac{W}{m^2\,°C}$

Pressure drop Calculation:

Pressure drop in shell:

$$\Delta P_s = 8 \times J_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{L_B}\right)\left(\frac{\rho\,\mu^2}{2}\right)\left(\frac{\mu}{\mu_w}\right)^{-0.14} = 8.7\,psi$$

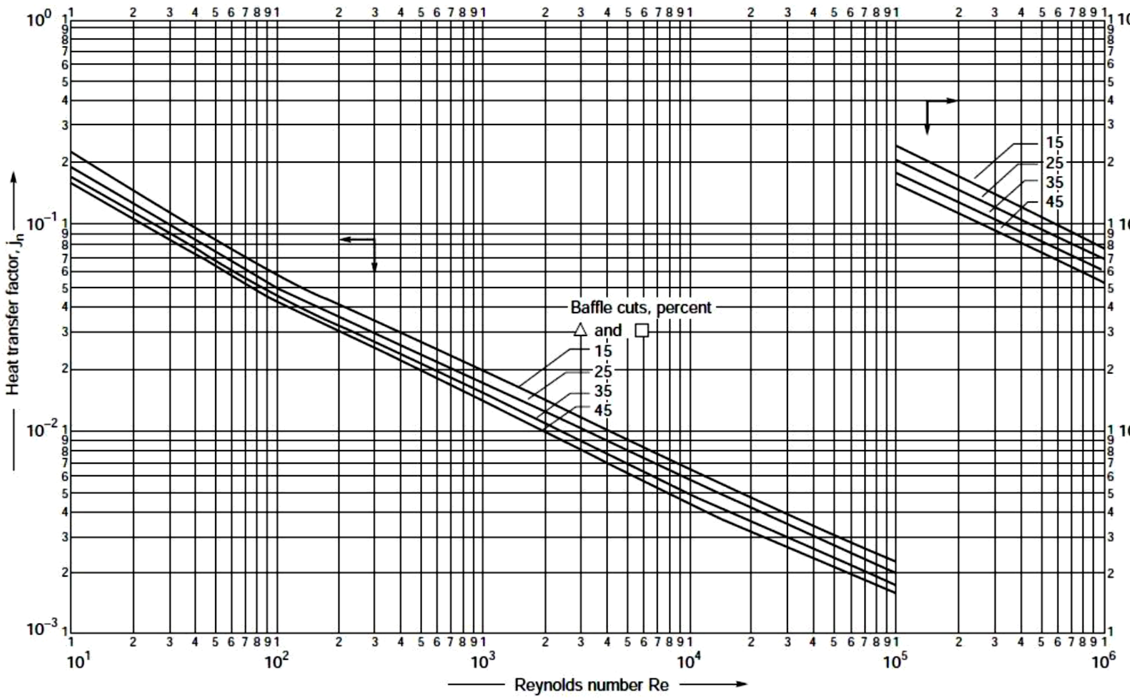


Figure 18: Jh factor

Pressure drop in tubes:

$$\Delta P_t = N_P\left[8\times J_h\left(\frac{L}{d_i}\right)\left(\frac{\mu}{\mu_w}\right)^{-m}+2.5\right]\times\frac{\rho u_t^2}{2}$$

$$\Delta P_t = 5.8\ psi$$

Heat Exchanger 2

Table 26: Hx-2

Crude oil	Steam
$K = 0.1121\ \frac{W}{m\ ℃}$	$K = 0.03144\frac{W}{m\ ℃}$
$\mu = 1.236e-3\ \frac{Ns}{m^2}$	$\mu = 1.53e-5\frac{Ns}{m^2}$
$c_p = 1.903\frac{kJ}{kg\ ℃}$	$c_p = 1.948\frac{kJ}{kg\ ℃}$
$\rho = 772.7\frac{kg}{m^3}$	$\rho = 0.7227\frac{kg}{m^3}$
$\dot{m} = 5.18e4\ kg/s$	$\dot{m} = 1.1e5\ kg/s$
$T_h = 150$ ℃	$T_h' = 220.2$ ℃
$T_c = 107$ ℃	$T_c' = 180$ ℃

Exchanger Duty:

$Q = \dot{m}c_p\Delta T$ = 1.265e7 kJ/h

Water flow rate:

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

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

$\dot{m} = 1.1\text{e}5$ kg/h

Logarithmic mean of Temperature Difference (LMTD):

$$LMTD = \frac{\theta_2 - \theta_1}{\ln(\frac{\theta_2}{\theta_1})} = 71.38\ ℃$$

Where $\theta = T_h - T_c$

$$R = \frac{T_h' - T_c'}{t_c - t_h} = 4$$

$$P = \frac{t_c - t_h}{T_h' - t_c} = 0.1$$

Shell side 1 pass

Tube side double pass

$F_t = 1$

So, Corrected temperature difference Δt =71.383℃

$U = 2065\ \frac{W}{m^2℃}$ from table in Supra

Area:

$$A = \frac{Q}{U\Delta t}$$

$$A = \ 58.81\ m^2$$

BWG-17: 0.01 m outer-side radius and 0.075 m inside radius

Length of tube $= l = 3.66\ m$

Area of single tube $= a = \pi(l)(do) = \ \pi \times 3.66 \times 0.03 = 0.334\ m^2$

Number of tubes $= n = \frac{A}{a} = 234\ tubes$

Tube bundle diameter = 0.32 m

Clearance = 0.0093 m

Shell side co-efficient:

Baffle spacing (Ib) = 0.5 × $ds\ = 0.46557$

Pt = 0.0009 m

As = $\frac{(P_t - P_o) \times D_s}{P_t}$ = 60.32 m2

Mass velocity 'Ga' =$\frac{\omega}{\text{As}}$ = 1.2e5 $\frac{kg}{m^2 s}$

$$D_e = \frac{1.10}{d_o}\left(P_t^2 - 0.917{d_o}^2\right)$$

Re = 1823

By using figure above $\quad J_h = 0.12$

$$Pr = \frac{\mu \times c_p}{K}$$

Pr = 1

$$\frac{h_o}{\theta_s} = \frac{J_h\,(Re)(Pr)^{0.33}\left(K_f\right)}{De} =$$

$h_o = 0.242\,\frac{W}{m^2\ ℃}$

Tube side Coefficient:

Cross section area of tube $= \frac{\pi}{4}\ d_i^2\ = 0.21\ m^2$

Number of tubes per pass n = 241

Overall area = $60.32\ m^2$

Mass velocity ‘Ga’ = $\frac{\omega}{\mathrm{As}} = 1\mathrm{e}4\,\frac{kg}{m^2 s}$

$$Re = \frac{D_i \times Ga}{\mu} = 35500$$

$$Pr = \frac{\mu \times c_p}{K} = 0.1$$

L/D=0.0005

$$\frac{h_i}{\theta_t} = \frac{J_h\,(Re)(Pr)^{0.33}\left(K_f\right)}{De} = \ 0.2$$

$\frac{h_o}{\theta_t} = 0.2\,\frac{W}{m^2\ ℃}$

Mean Wall temperature $\boldsymbol{\theta_t}$:

$\left(\frac{\mu}{\mu_w}\right)^{0.14}$=1

$h_i = \ \ \frac{W}{m^2\ ℃}$

Overall convective heat transfer Coefficient:

$h_{io} = h_i \times \frac{ID}{OD} = 0.242\, \frac{W}{m^2\ ℃}$

$U_c = \frac{h_{io} \times h_o}{h_{io} + h_o} = 2994\, \frac{W}{m^2\ ℃}$

For Dirt factor Rd =0.003

$$R_D = \frac{U_c - U_D}{U_c \times U_D}$$

$U_D = 2800\, \frac{W}{m^2\ ℃}$

Pressure drop Calculation:

Pressure drop in shell:

$$\Delta P_s = 8 \times J_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{L_B}\right)\left(\frac{\rho\, \mu^2}{2}\right)\left(\frac{\mu}{\mu_w}\right)^{-0.14} = 4.35\, psi$$

Pressure drop in tubes:

$$\Delta P_t = N_P \left[8 \times J_h \left(\frac{L}{d_i}\right)\left(\frac{\mu}{\mu_w}\right)^{-m} + 2.5\right] \times \frac{\rho u_t^2}{2}$$

$$\Delta P_t = 1.45\, psi$$

Heat Exchanger 3

Table 27: Hx-3

water	Overheads
$\boldsymbol{K = 0.6110\ \frac{W}{m\ ℃}}$	$K = 2.278e - 2\frac{W}{m\ ℃}$
$\boldsymbol{\mu = 8.9e - 4\ \frac{Ns}{m^2}}$	$\mu = 8.78e - 6\frac{Ns}{m^2}$
$\boldsymbol{c_p = 4.313\frac{kJ}{kg\ ℃}}$	$c_p = 2.18\frac{kJ}{kg\ ℃}$
$\boldsymbol{\rho = 1007\frac{kg}{m^3}}$	$\rho = 14.69\frac{kg}{m^3}$
$\boldsymbol{\dot{m} = 1.2e5\ kg/s}$	$\dot{m} = 9383\ kg/s$
$\boldsymbol{T_h = 32.74}$ ℃	$T_h' = 143$ ℃
$\boldsymbol{T_c = 25}$ ℃	$T_c' = 70$ ℃

Exchanger Duty:

$Q = \dot{m} c_p \Delta T$ = 1.2e5 kJ/h

Water flow rate:

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

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

$\dot{m} = 1.1\text{e}5$ kg/h

Logarithmic mean of Temperature Difference (LMTD):

$$LMTD = \frac{\theta_2 - \theta_1}{\ln(\frac{\theta_2}{\theta_1})} = 73.170\ ℃$$

Where $\theta = T_h - T_c$

$$R = \frac{T_h' - T_c'}{t_c - t_h} = 1$$

$$P = \frac{t_c - t_h}{T_h' - t_c} = 0.59$$

Shell side 1 pass

Tube side double pass

$F_t = 0.981$

So, Corrected temperature difference Δt =71.79 ℃

$U = 956.9\ \frac{W}{m^2℃}$ from table in Supra

Area:

$$A = \frac{Q}{U\Delta t}$$

$$A = 60.32\ m^2$$

BWG-17: 0.01 m outer-side radius and 0.075 m inside radius

Length of tube $= l = 6\ m$

Area of single tube = $a = \pi(l)(do) = \ \pi \times 3.66 \times 0.03 = 0.334\ m^2$

Number of tubes = $n = \frac{A}{a} = 160\ tubes$

Tube bundle diameter = 0.7 m

Clearance = 0.0043 m

Shell side co-efficient:

Baffle spacing (Ib) = 0.5 × $ds\ = 0.8\ m$

Pt = 0.005 m

As = $\frac{(P_t - P_o) \times D_s}{P_t}$ = 60.32 m2

Mass velocity 'Ga' =$\frac{\omega}{\text{As}} = 1.2\text{e}5\ \frac{kg}{m^2 s}$

$$D_e = \frac{1.10}{d_o}\left(P_t^2 - 0.917{d_o}^2\right)$$

Re = 1539

By using figure above $\ \ J_h = 0.12$

$$Pr = \frac{\mu \times c_p}{K}$$

Pr = 1

$$\frac{h_o}{\theta_s} = \frac{J_h\ (Re)(Pr)^{0.33}\left(K_f\right)}{De} =$$

$h_o = 0.242 \frac{W}{m^2\,°C}$

Tube side Coefficient:

Cross section area of tube $= \frac{\pi}{4}\, d_i^2 = 0.21\, m^2$

Number of tubes per pass n = 241

Overall area = $60.32\, m^2$

Mass velocity 'Ga' = $\frac{\omega}{\mathrm{As}} = 1e4 \frac{kg}{m^2 s}$

$$Re = \frac{D_i \times Ga}{\mu} = 35500$$

$$Pr = \frac{\mu \times c_p}{K} = 0.1$$

L/D=0.0005

$$\frac{h_i}{\theta_t} = \frac{J_h\,(Re)(Pr)^{0.33}\left(K_f\right)}{De} = 0.2$$

$\frac{h_o}{\theta_t} = 0.2 \frac{W}{m^2\,°C}$

Mean Wall temperature $\boldsymbol{\theta_t}$:

$\left(\frac{\mu}{\mu_w}\right)^{0.14}$=1

$h_i = \quad \frac{W}{m^2\,°C}$

Overall convective heat transfer Coefficient:

$h_{io} = h_i \times \frac{ID}{OD} = 0.242 \frac{W}{m^2\,°C}$

$U_c = \frac{h_{io} \times h_o}{h_{io} + h_o} = 2994\,\frac{W}{m^2\ ℃}$

For Dirt factor Rd =0.003

$$R_D = \frac{U_c - U_D}{U_c \times U_D}$$

$U_D = 2800\,\frac{W}{m^2\ ℃}$

Pressure drop Calculation:

Pressure drop in shell:

$$\Delta P_s = 8 \times J_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{L_B}\right)\left(\frac{\rho\,\mu^2}{2}\right)\left(\frac{\mu}{\mu_w}\right)^{-0.14} = 8.7\,psi$$

Pressure drop in tubes:

$$\Delta P_t = N_P\left[8 \times J_h\left(\frac{L}{d_i}\right)\left(\frac{\mu}{\mu_w}\right)^{-m} + 2.5\right] \times \frac{\rho u_t^2}{2}$$

$$\Delta P_t = 5.8\,psi$$

CHAPTER 7

Simulation

7.1 Simulation Main Flow-Sheet

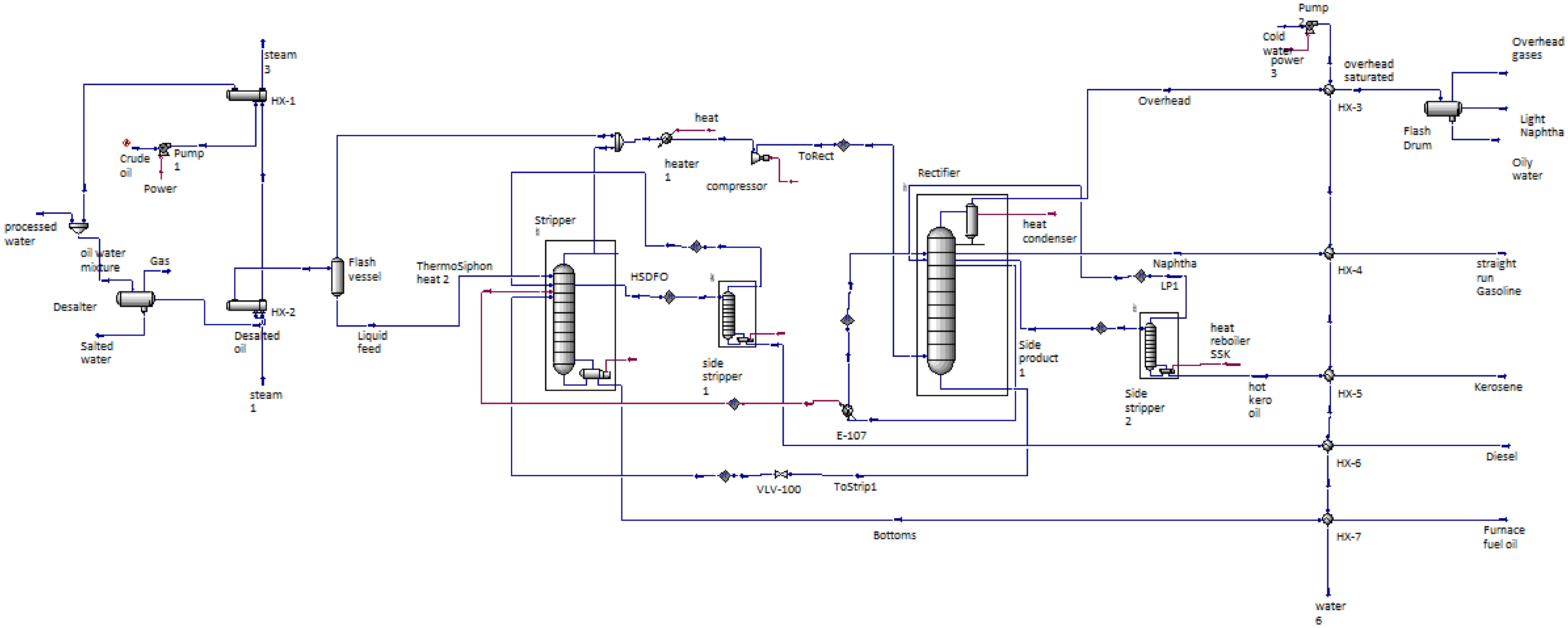


Figure 19: Simulation Flow Sheet

7.2 Petroleum Essay input

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

a) Open Aspen Hysys V11.

b) Click on Properties tab at left bottom corner.

c) Click on component list and add all the lighter fractions one by one.

d) Now click on petroleum management tab and click Lighter Fractions.

e) Add data given in table below.

Table 28: Light components

Light Components		
Components	**Mole Fraction**	**Boiling Point (C)**
Methane	0	-161.525
Ethane	0	-88.59999695
Propane	0.146	-42.10199585
i-Butane	0.127	-11.72998657
n-Butane	0.702	-0.501989746
i-Pentane	6.54E-01	27.87801514
n-Pentane	1.30E+00	36.05901489
H2O	0.00E+00	99.99801025
NACL	0.00E+00	326.85

a) Now click on Properties tab at left bottom corner.

b) Click on component list and add all the lighter fractions one by one.

c) Now click on petroleum management tab and click Lighter Fractions.

d) Add data given in table below.

Table 29: Hypothetical components

Hypothetical components	Boiling Point (C)	Molecular weight(g/mol)	Density (kg/m3)	Viscosity (cP)	Molar Fraction
NBP_49	49.03	75.39	658.96	0.2482	0.0289
NBP_63	63.17	81.83	671.14	0.2731	0.0475
NBP_77	76.57	88.00	691.36	0.2939	0.0688
NBP_90	90.23	93.08	722.75	0.3439	0.0858
NBP_104	104.37	99.22	747.95	0.3989	0.0873
NBP_118	118.20	106.16	761.94	0.4522	0.0712
NBP_133	133.01	113.73	770.54	0.5113	0.0559
NBP_147	147.35	121.52	782.03	0.5837	0.0529
NBP_161	161.48	128.90	795.33	0.6634	0.0442
NBP_176	176.07	138.05	800.00	0.7515	0.0383
NBP_190	190.17	147.59	803.84	0.8678	0.0331
NBP_205	204.84	157.73	808.39	1.0159	0.0281

Hypothetical components	**Boiling Point (C)**	**Molecular weight(g/mol)**	**Density (kg/m3)**	**Viscosity (cP)**	**Molar Fraction**
NBP_220	219.61	169.25	815.79	1.2212	0.0313
NBP_233	232.77	176.98	834.50	1.4661	0.0272
NBP_248	247.59	185.91	845.28	1.7382	0.0186
NBP_262	262.32	198.66	849.28	2.0977	0.0177
NBP_277	276.98	214.03	855.07	2.6403	0.0231
NBP_290	290.10	225.68	864.78	3.2602	0.0218
NBP_305	304.82	237.79	871.11	3.9733	0.0154
NBP_319	319.40	252.02	874.63	4.8660	0.0128
NBP_334	333.83	267.76	877.53	6.0338	0.0118
NBP_348	348.13	283.87	881.57	7.6257	0.0112
NBP_362	362.38	298.67	888.23	9.8845	0.0104
NBP_377	376.72	311.37	896.94	13.1389	0.0093
NBP_391	391.12	323.17	907.18	18.1815	0.0087
NBP_405	405.17	334.92	918.67	26.0995	0.0078
NBP_420	419.82	346.79	930.61	38.5754	0.0066
NBP_440	440.46	366.84	949.56	78.1394	0.0123
NBP_468	467.77	393.90	975.74	2.67E+02	0.0091
NBP_497	496.83	414.37	1000.33	1.09E+03	0.0075
NBP_523	523.07	434.07	1025.44	6.53E+03	0.0071
NBP_551	550.83	449.60	1048.10	4.65E+04	0.0046
NBP_580	579.82	474.01	1066.22	5.85E+05	0.0038
NBP_606	605.88	500.00	1083.01	5.57E+06	0.0033
NBP_634	633.90	523.77	1096.98	3.95E+07	0.0023
NBP_672	672.04	552.29	1109.66	3.73E+08	0.0026
NBP_722	722.42	588.85	1124.82	2.55E+10	0.0018
NBP_769	769.39	632.15	1143.01	2.01E+12	0.0019

7.3 Product Analysis

After adding all these data

a) Press on calculate assay.
b) Confirm that table is converged.
c) Click on output blend and Install crude oil

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

Table 30: Product distribution

Products Analysis	
Products	Temeprature (C)
Off Gas	10
Lt St Run	70
Naphtha	180
Kerosene	240
Light Diesel	290
Heavy Diesel	340
Atm Gas Oil	370
Residue	1200

Product distribution chart shows that out Blend has 37 percent Gasoline in it. Which can be extracted at cut point of 180°C.

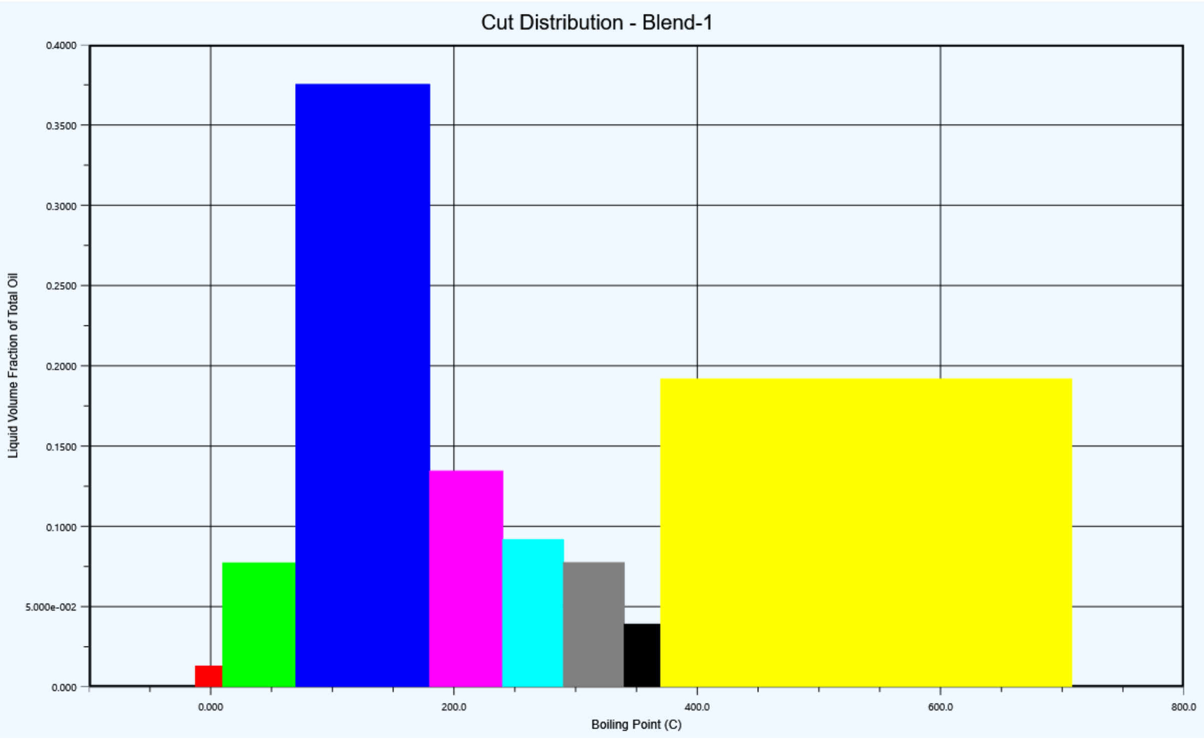


Figure 20: Cut distribution

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

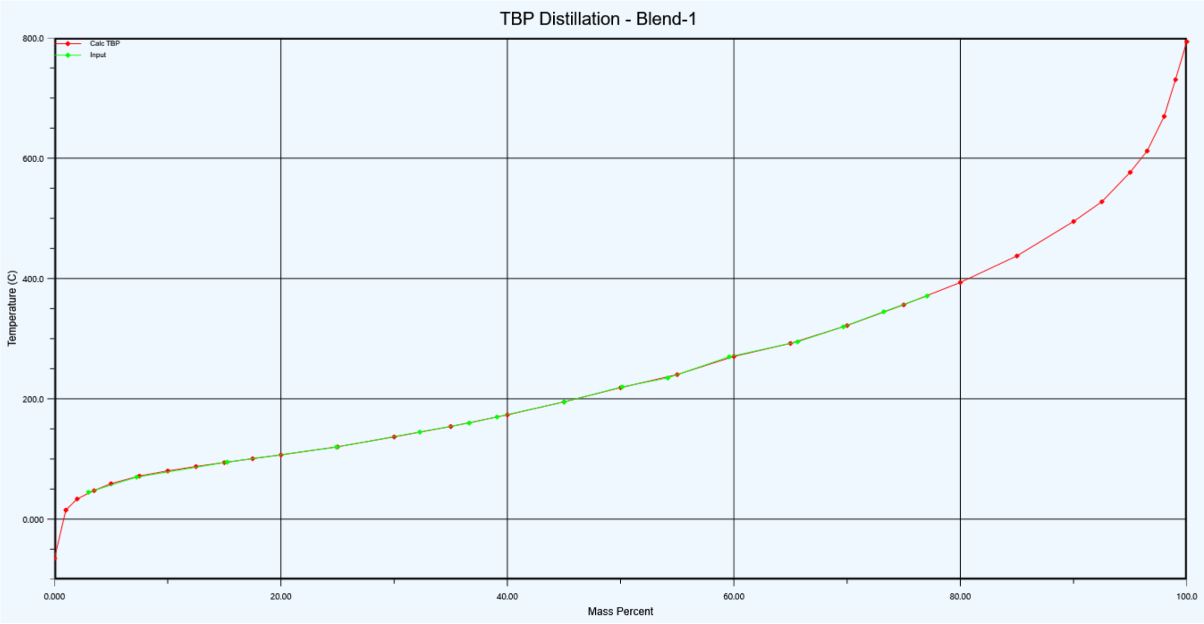


Figure 21: True boiling point curve

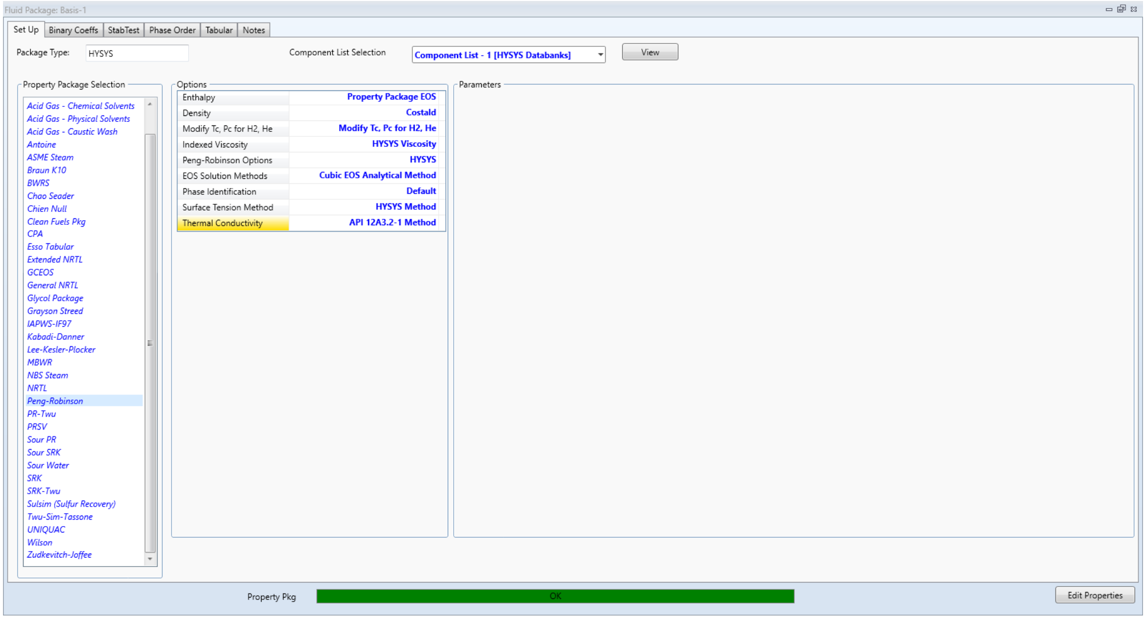
7.4 Binary Coefficients

Table 31: Binary Coefficients

Methane	Ethane	Propane	i-Butane	n-Butane	i-Pentane	n-Pentane	NBP[0]49*	NBP[0]77*	NBP[0]90*	NBP[0]104*	NBP[0]118*	NBP[0]133*	NBP[0]147*	NBP[0]161*	NBP[0]262*	NBP[0]277*	NBP[0]290*	NBP[0]319*	NBP[0]348*	NBP[0]362*	NBP[0]377*	NBP[0]391*	NBP[0]405*	NBP[0]420*	NBP[0]468*	NBP[0]497*	NBP[0]523*	NBP[0]551*
Methane	---	0.0022	0.0068	0.0131	0.0123	0.0176	0.0000	0.02263	0.02425	0.02537	0.02683	0.02891	0.03162	0.03395	0.05082	0.05387	0.05663	0.06101	0.06717	0.07005	0.07241	0.07432	0.07586	0.07696	0.07894	0.07978	0.08123	0.08198
Ethane	0.0022	---	0.0013	0.0046	0.0041	0.0074	0.0000	0.01082	0.01196	0.01277	0.01383	0.01536	0.01739	0.01916	0.03258	0.03508	0.03738	0.04104	0.04626	0.04873	0.05076	0.05241	0.05374	0.05470	0.05642	0.05715	0.05842	0.05907
Propane	0.0068	0.0013	---	0.0010	0.0008	0.0026	0.0000	0.00474	0.00552	0.00607	0.00681	0.00791	0.00940	0.01073	0.02137	0.02344	0.02535	0.02844	0.03288	0.03501	0.03676	0.03819	0.03935	0.04019	0.04169	0.04233	0.04345	0.04402
i-Butane	0.0131	0.0046	0.0010	---	0.0000	0.0003	0.0000	0.00135	0.00178	0.00210	0.00255	0.00324	0.00422	0.00513	0.01313	0.01478	0.01633	0.01885	0.02255	0.02434	0.02583	0.02705	0.02804	0.02875	0.03005	0.03060	0.03157	0.03206
n-Butane	0.0123	0.0041	0.0008	0.0000	---	0.0005	0.0000	0.00163	0.00210	0.00244	0.00293	0.00366	0.00470	0.00566	0.01397	0.01567	0.01725	0.01984	0.02363	0.02545	0.02697	0.02822	0.02923	0.02996	0.03128	0.03184	0.03282	0.03333
i-Pentane	0.0176	0.0074	0.0026	0.0003	0.0005	---	0.0000	0.00033	0.00055	0.00074	0.00102	0.00147	0.0216	0.00282	0.00926	0.01067	0.01199	0.01418	0.01743	0.01902	0.02034	0.02144	0.02233	0.02297	0.02414	0.02464	0.02551	0.02596
n-Pentane	0.0179	0.0076	0.0027	0.0004	0.0005	0.0000	0.0000	0.00029	0.00050	0.00068	0.00095	0.00138	0.00205	0.00270	0.00905	0.01043	0.01175	0.01391	0.01714	0.01871	0.02003	0.02111	0.02200	0.02264	0.02380	0.02429	0.02516	0.02561
NBP[0]49*	0.0206	0.0094	0.0038	0.0009	0.0011	0.0001	0.0000	0.00005	0.00016	0.00026	0.00044	0.00075	0.00126	0.00178	0.00729	0.00855	0.00974	0.01173	0.01471	0.01617	0.01740	0.01842	0.01925	0.01985	0.02094	0.02140	0.02222	0.02265
NBP[0]63*	0.0226	0.0108	0.0047	0.0013	0.0016	0.0003	0.0000	---	0.00003	0.00008	0.00019	0.00041	0.00080	0.00123	0.00613	0.00729	0.00840	0.01025	0.01305	0.01444	0.01561	0.01657	0.01736	0.01793	0.01897	0.01942	0.02020	0.02060
NBP[0]77*	0.0242	0.0120	0.0055	0.0018	0.0021	0.0006	0.0000	0.00003	---	0.00001	0.00007	0.00022	0.00052	0.00088	0.00531	0.00639	0.00744	0.00919	0.01186	0.01318	0.01430	0.01522	0.01598	0.01653	0.01753	0.01796	0.01872	0.01911
NBP[0]90*	0.0254	0.0128	0.0061	0.0021	0.0024	0.0007	0.0000	0.00008	0.00001	---	0.00002	0.00012	0.00037	0.00067	0.00480	0.00582	0.00682	0.00850	0.01108	0.01237	0.01345	0.01435	0.01509	0.01562	0.01660	0.01701	0.01775	0.01813
NBP[0]104*	0.0268	0.0138	0.0068	0.0025	0.0029	0.0010	0.0000	0.00019	0.00007	0.00002	---	0.00004	0.00021	0.00045	0.00417	0.00514	0.00608	0.00767	0.01013	0.01136	0.01240	0.01327	0.01398	0.01449	0.01544	0.01584	0.01655	0.01692
NBP[0]118*	0.0289	0.0154	0.0079	0.0032	0.0037	0.0015	0.0000	0.00041	0.00022	0.00012	0.00004	---	0.00007	0.00022	0.00338	0.00426	0.00512	0.00659	0.00889	0.01004	0.01102	0.01184	0.01252	0.01300	0.01390	0.01428	0.01496	0.01531
NBP[0]133*	0.0316	0.0174	0.0094	0.0042	0.0047	0.0022	0.0000	0.00080	0.00052	0.00037	0.00021	0.00007	---	0.00005	0.00251	0.00327	0.00403	0.00535	0.00744	0.00850	0.00941	0.01017	0.01080	0.01125	0.01209	0.01244	0.01308	0.01340
NBP[0]147*	0.0340	0.0192	0.0107	0.0051	0.0057	0.0028	0.0000	0.00123	0.00088	0.00067	0.00045	0.00022	0.00005	---	0.00188	0.00255	0.00323	0.00442	0.00634	0.00732	0.00817	0.00888	0.00946	0.00989	0.01067	0.01101	0.01161	0.01192
NBP[0]161*	0.0360	0.0207	0.0119	0.0060	0.0066	0.0035	0.0000	0.00166	0.00125	0.00100	0.00073	0.00042	0.00015	0.00003	0.00142	0.00201	0.00261	0.00369	0.00546	0.00638	0.00717	0.00784	0.00839	0.00879	0.00954	0.00985	0.01042	0.01072
NBP[0]233*	0.0487	0.0308	0.0199	0.0120	0.0128	0.0083	0.0000	0.00535	0.00459	0.00411	0.00353	0.00281	0.00202	0.00146	0.00003	0.00015	0.00035	0.00080	0.00172	0.00226	0.00274	0.00316	0.00352	0.00378	0.00428	0.00449	0.00488	0.00508
NBP[0]248*	0.0508	0.0326	0.0214	0.0131	0.0140	0.0093	0.0000	0.00613	0.00531	0.00480	0.00417	0.00338	0.00251	0.00188	---	0.00005	0.00018	0.00054	0.00132	0.00179	0.00223	0.00261	0.00293	0.00317	0.00363	0.00383	0.00419	0.00438
NBP[0]262*	0.0539	0.0351	0.0234	0.0148	0.0157	0.0107	0.0000	0.00729	0.00639	0.00582	0.00514	0.00426	0.00327	0.00255	0.00005	---	0.00004	0.00026	0.00086	0.00124	0.00161	0.00193	0.00221	0.00242	0.00283	0.00300	0.00332	0.00349
NBP[0]277*	0.0566	0.0374	0.0254	0.0163	0.0173	0.0120	0.0000	0.00840	0.00744	0.00682	0.00608	0.00512	0.00403	0.00323	0.00018	0.00004	---	0.00009	0.00053	0.00084	0.00114	0.00142	0.00166	0.00184	0.00220	0.00235	0.00264	0.00279
NBP[0]290*	0.0584	0.0388	0.0266	0.0173	0.0183	0.0129	0.0000	0.00912	0.00812	0.00748	0.00670	0.00569	0.00454	0.00369	0.00030	0.00011	0.00002	0.00003	0.00036	0.00063	0.00089	0.00114	0.00136	0.00152	0.00184	0.00199	0.00225	0.00239
NBP[0]319*	0.0640	0.0436	0.0306	0.0207	0.0217	0.0158	0.0000	0.01161	0.01048	0.00975	0.00886	0.00769	0.00635	0.00533	0.00088	0.00051	0.00027	0.00004	0.00004	0.00016	0.00031	0.00046	0.00060	0.00071	0.00094	0.00104	0.00123	0.00133
NBP[0]348*	0.0701	0.0487	0.0350	0.0243	0.0255	0.0190	0.0000	0.01444	0.01318	0.01237	0.01136	0.01004	0.00850	0.00732	0.00179	0.00124	0.00084	0.00037	0.00004	---	0.00002	0.00008	0.00014	0.00020	0.00032	0.00038	0.00050	0.00057
NBP[0]362*	0.0724	0.0508	0.0368	0.0258	0.0270	0.0203	0.0000	0.01561	0.01430	0.01345	0.01240	0.01102	0.00941	0.00817	0.00223	0.00161	0.00114	0.00058	0.00012	0.00002	---	0.00002	0.00005	0.00008	0.00017	0.00022	0.00031	0.00036
NBP[0]377*	0.0743	0.0524	0.0382	0.0270	0.0282	0.0214	0.0000	0.01657	0.01522	0.01435	0.01327	0.01184	0.01017	0.00888	0.00261	0.00193	0.00142	0.00078	0.00022	0.00008	0.00002	---	0.00001	0.00003	0.00008	0.00012	0.00019	0.00023
NBP[0]391*	0.0759	0.0537	0.0394	0.0280	0.0292	0.0223	0.0000	0.01736	0.01598	0.01509	0.01398	0.01252	0.01080	0.00946	0.00293	0.00221	0.00166	0.00096	0.00032	0.00014	0.00005	0.00001	---	0.00000	0.00004	0.00006	0.00011	0.00015
NBP[0]405*	0.0770	0.0547	0.0402	0.0288	0.0300	0.0230	0.0000	0.01793	0.01653	0.01562	0.01449	0.01300	0.01125	0.00989	0.00317	0.00242	0.00184	0.00110	0.00040	0.00020	0.00008	0.00003	0.00000	---	0.00002	0.00003	0.00007	0.00010
NBP[0]420*	0.0780	0.0556	0.0410	0.0294	0.0307	0.0236	0.0000	0.01849	0.01707	0.01614	0.01500	0.01348	0.01170	0.01031	0.00341	0.00264	0.00203	0.00125	0.00049	0.00026	0.00013	0.00005	0.00002	0.00000	0.00000	0.00001	0.00004	0.00006
NBP[0]468*	0.0798	0.0571	0.0423	0.0306	0.0318	0.0246	0.0000	0.01942	0.01796	0.01701	0.01584	0.01428	0.01244	0.01101	0.00383	0.00300	0.00235	0.00150	0.00066	0.00038	0.00022	0.00012	0.00006	0.00003	0.00000	---	0.00001	0.00002
NBP[0]497*	0.0812	0.0584	0.0434	0.0316	0.0328	0.0255	0.0000	0.02020	0.01872	0.01775	0.01655	0.01496	0.01308	0.01161	0.00419	0.00332	0.00264	0.00173	0.00081	0.00050	0.00031	0.00019	0.00011	0.00007	0.00002	0.00001	---	0.00000
NBP[0]523*	0.0820	0.0591	0.0440	0.0321	0.0333	0.0260	0.0000	0.02060	0.01911	0.01813	0.01692	0.01531	0.01340	0.01192	0.00438	0.00349	0.00279	0.00186	0.00089	0.00057	0.00036	0.00023	0.00015	0.00010	0.00004	0.00002	0.00000	---
NBP[0]551*	0.0835	0.0604	0.0452	0.0331	0.0344	0.0269	0.0000	0.02145	0.01992	0.01892	0.01769	0.01604	0.01409	0.01257	0.00478	0.00385	0.00311	0.00212	0.00108	0.00072	0.00049	0.00033	0.00023	0.00017	0.00008	0.00005	0.00002	0.00001

7.5 Fluid Package

Peng Robinson fluid package was used for the calculation Enthalpy and Cubic equation of state analytical method applied for properties calculation and all other parameters and set to be default.



Package Type: HYSYS

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

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 22: Fluid Package

The fluid package is the set of equations that helps the program come up with relevant fluid properties at given states. Since a process is nothing but a change of states, the simulation needs underlying relationships to come up with values of properties pertaining to the system. The selection of the fluid package depends on the type of components involved.

For Hydrocarbon systems where most components are nonpolar and the process is distillation , Peng Robinson fluid package is enough .

7.6 Column Environments

Stripper column environment is represented as follows which include reboiler and its condenser is attached to rectifier. So, its reflux can be controlled by condenser duty at rectifier. Pressure drop can be observed on the graph below with is linear with the column tray number which can be non-linear in real case which is nearly impossible to model.

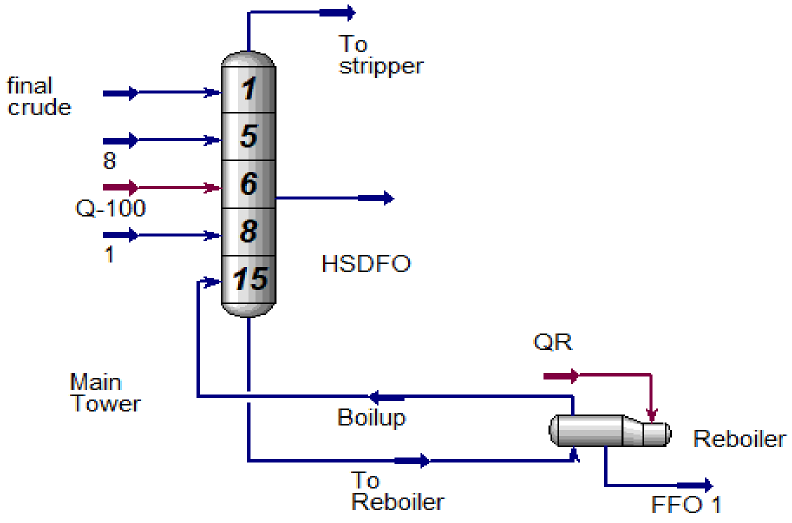
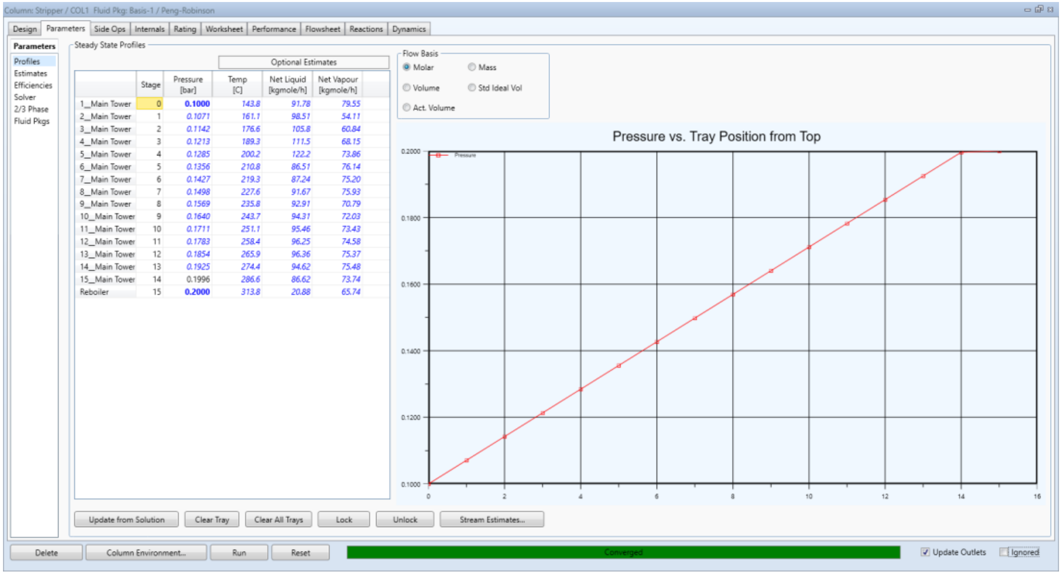


Figure 23:Stripper



Column: Stripper / COL1 Fluid Pkg: Basis-1 / Peng-Robinson

Steady State Profiles

	Stage	Pressure [bar]	Temp [C]	Net Liquid [kgmole/h]	Net Vapour [kgmole/h]
1__Main Tower	0	0.1000	143.8	91.78	79.55
2__Main Tower	1	0.1071	161.1	98.51	54.11
3__Main Tower	2	0.1142	176.6	105.8	60.84
4__Main Tower	3	0.1213	189.3	111.5	68.15
5__Main Tower	4	0.1285	200.2	122.2	73.86
6__Main Tower	5	0.1356	210.8	86.51	76.14
7__Main Tower	6	0.1427	219.3	87.24	75.20
8__Main Tower	7	0.1498	227.6	91.67	75.93
9__Main Tower	8	0.1569	235.8	92.91	70.79
10__Main Tower	9	0.1640	243.7	94.31	72.03
11__Main Tower	10	0.1711	251.1	95.46	73.43
12__Main Tower	11	0.1783	258.4	96.25	74.58
13__Main Tower	12	0.1854	265.9	96.36	75.37
14__Main Tower	13	0.1925	274.4	94.62	75.48
15__Main Tower	14	0.1996	286.6	86.62	73.74
Reboiler	15	0.2000	313.8	20.88	65.74

Flow Basis: ● Molar ○ Mass ○ Volume ○ Std Ideal Vol ○ Act. Volume

Converged

☑ Update Outlets ☐ Ignored

Figure 24: Stripper performance

Rectifier column environment is represented as follows which include condenser and its reboiler is attached to stripper shown in supra. So, its boil-up ratio can be controlled by reboiler duty at stripper. Pressure drop can be observed on the graph below with is linear with the column tray number which can be non-linear in real case which is nearly impossible to model.

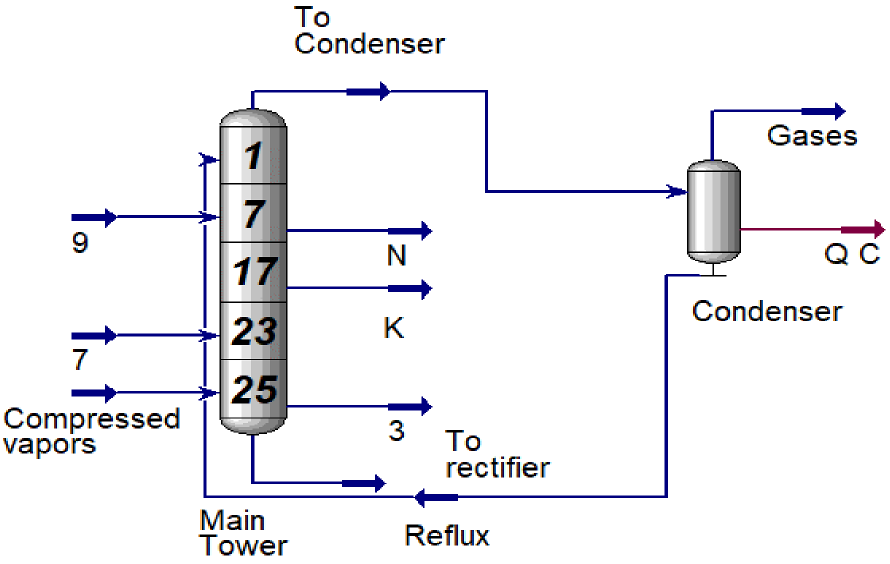
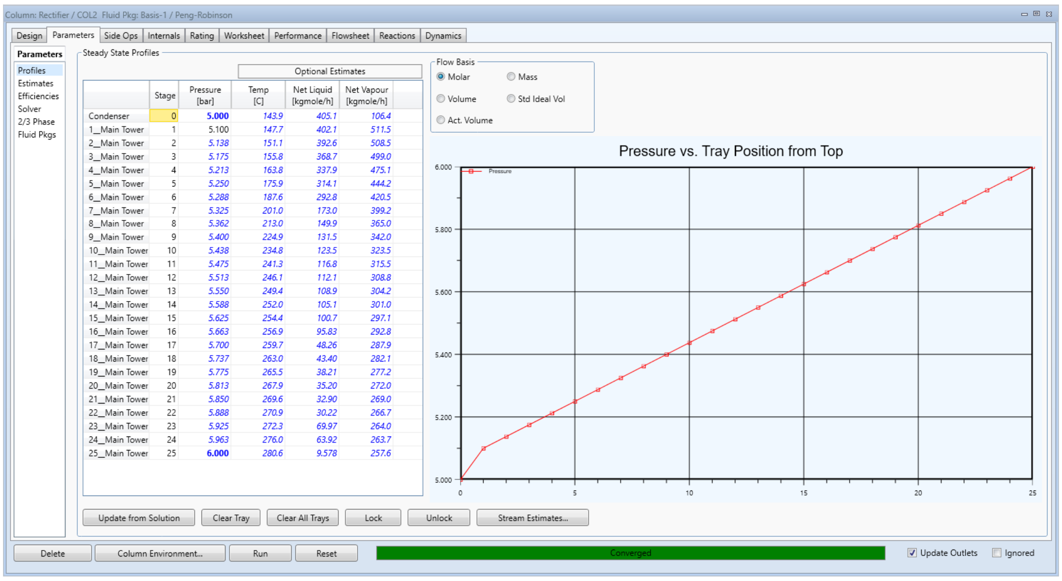


Figure 25: Rectifier


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

Steady State Profiles

Flow Basis: ● Molar ○ Mass ○ Volume ○ Std Ideal Vol ○ Act. Volume

	Stage	Pressure [bar]	Temp [C]	Net Liquid [kgmole/h]	Net Vapour [kgmole/h]
Condenser	0	5.000	143.9	405.1	106.4
1__Main Tower	1	5.100	147.7	402.1	511.5
2__Main Tower	2	5.138	151.1	392.6	508.5
3__Main Tower	3	5.175	155.8	368.7	499.0
4__Main Tower	4	5.213	163.8	337.9	475.1
5__Main Tower	5	5.250	175.9	314.1	444.2
6__Main Tower	6	5.288	187.6	292.8	420.5
7__Main Tower	7	5.325	201.0	173.0	399.2
8__Main Tower	8	5.362	213.0	149.9	365.0
9__Main Tower	9	5.400	224.9	131.5	342.0
10__Main Tower	10	5.438	234.8	123.5	323.5
11__Main Tower	11	5.475	241.3	116.8	315.5
12__Main Tower	12	5.513	246.1	112.1	308.8
13__Main Tower	13	5.550	249.4	108.9	304.2
14__Main Tower	14	5.588	252.0	105.1	301.0
15__Main Tower	15	5.625	254.4	100.7	297.1
16__Main Tower	16	5.663	256.9	95.83	292.8
17__Main Tower	17	5.700	259.7	48.26	287.9
18__Main Tower	18	5.737	263.0	43.40	282.1
19__Main Tower	19	5.775	265.5	38.21	277.2
20__Main Tower	20	5.813	267.9	35.20	272.0
21__Main Tower	21	5.850	269.6	32.90	269.0
22__Main Tower	22	5.888	270.9	30.22	266.7
23__Main Tower	23	5.925	272.3	69.97	264.0
24__Main Tower	24	5.963	276.0	63.92	263.7
25__Main Tower	25	6.000	280.6	9.578	257.6

Converged

☑ Update Outlets ☐ Ignored

Figure 26: Rectifier Performance

Table 32: Overall material Balance by Aspen Hysys

Overall Material Balance											
	IN				Out						
(kg/h)	Crude oil	processed water	Cold water	steam 1	steam 3	Salted water	Overhead gases	Light Naphtha	straight run Gasoline	Kerosene	Diesel
Methane	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ethane	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Propane	89.38	0.00	0.00	0.00	0.00	0.00	5.39	0.98	0.13	0.00	0.00
n-Butane	267.38	0.00	0.00	0.00	0.00	0.00	26.63	14.61	1.56	0.03	0.00
i-Butane	77.90	0.00	0.00	0.00	0.00	0.00	7.02	2.86	0.32	0.01	0.00
n-Pentane	494.01	0.00	0.00	0.00	0.00	0.00	55.08	85.60	10.03	0.30	0.00
i-Pentane	249.10	0.00	0.00	0.00	0.00	0.00	28.31	34.69	3.99	0.11	0.00
salts	61.43	0.00	0.00	0.00	0.00	61.41	0.00	0.00	0.00	0.00	0.01
water	32.00	1900.00	9324.00	4398.00	4398.00	1800.94	2.73	0.09	0.02	0.00	0.01
C4-C5 (hydrocarbons)	22669.56	0.00	0.00	0.00	0.00	0.00	1353.14	7707.34	4138.12	117.57	0.00
C6-C11 (hydrocarbons)	7159.66	0.00	0.00	0.00	0.00	0.00	1.50	57.33	5968.90	539.83	0.31
C12-C16 (hydrocarbons)	9565.84	0.00	0.00	0.00	0.00	0.00	0.00	0.01	1715.43	4001.23	3644.04
C17-C21 (hydrocarbons)	3780.63	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.01	1.14	1487.79
C22-C28 (hydrocarbons)	7390.11	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	2125.54
Total	51837.00	1900.00	9324.00	4398.00	4398.00	1862.35	1479.79	7903.51	11838.52	4660.22	7257.68

Table 33: : Overall Energy Balance by Aspen Hysys

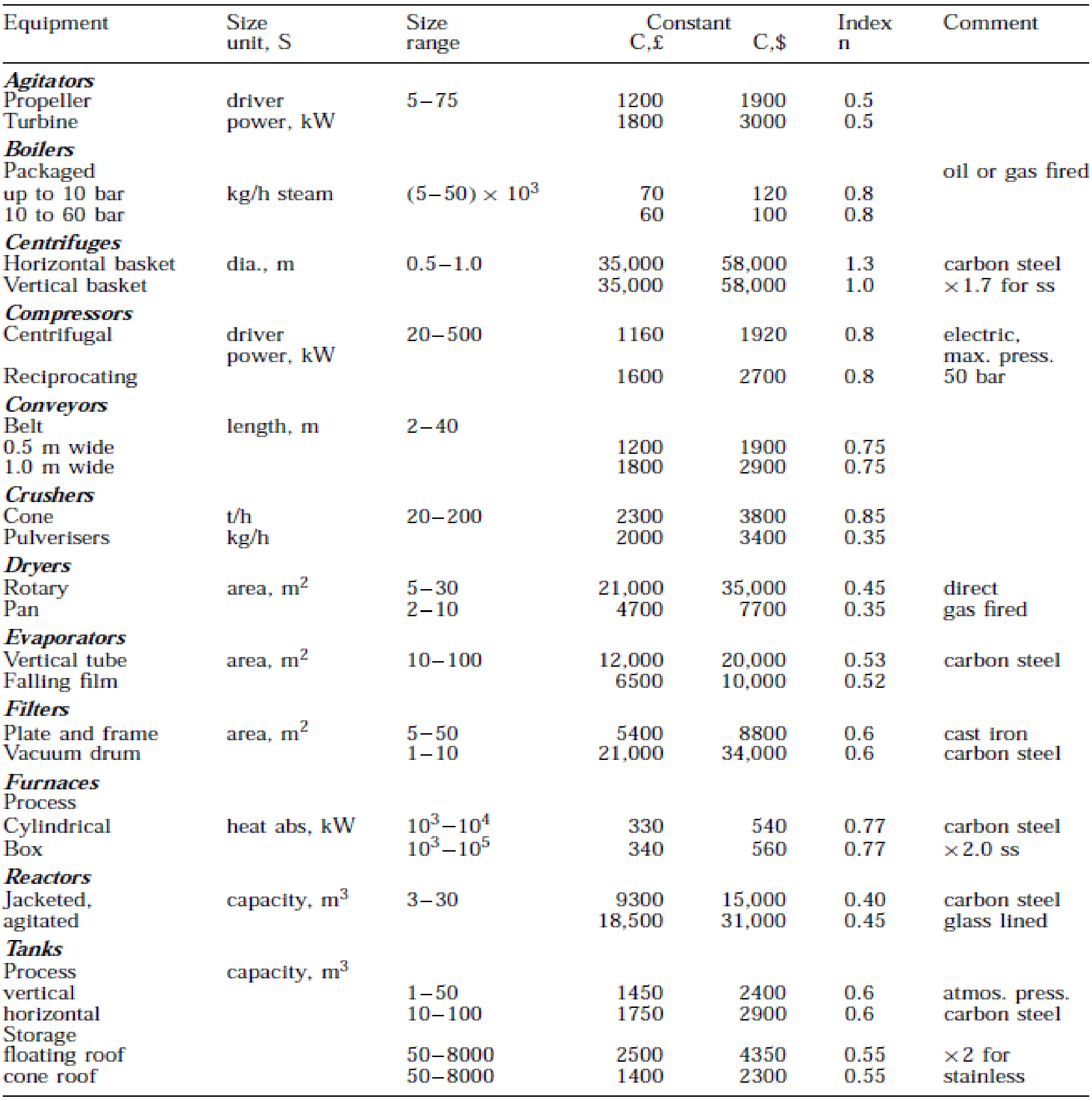
Overall Energy Balance																			
		IN							OUT										
	units	Crude oil	processed water	Cold water	steam 1	Power pump	power com	Heat reboiler	heat condenser	steam 3	Salted water	water 6	Gas	Overhead gases	Light Naphtha	straight run Gasoline	Kerosene	Diesel	Furnace fuel oil
Vapour Fraction		0.00	0.00	0.00E+00	1.00E+00	0.00	0.00	0.00	0.00	1.00E+00	0.00E+00	0.00E+00	1.00E+00	1.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Temperature	*C*	2.50E+01	2.50E+01	2.50E+01	2.20E+02	0.00	0.00	0.00	0.00	1.20E+02	1.07E+02	5.84E+01	1.07E+02	5.75E+01	5.86E+01	5.00E+01	4.50E+01	1.27E+02	6.33E+01
Pressure	*bar*	1.01E+00	2.00E+00	1.00E+00	1.60E+00	0.00	0.00	0.00	0.00	1.10E+00	1.40E+00	1.80E+00	1.40E+00	1.00E-01	6.00E-01	4.73E+00	1.25E+00	9.00E-01	1.05E+00
Heat Flow	*kJ/h*	-1.14E+08	-2.10E+06	-1.97E+09	-1.44E+09	0.00	0.00	0.00	0.00	-1.46E+09	-1.00E+06	-1.96E+09	-2.04E+07	-2.69E+06	-1.68E+07	-2.52E+07	-9.89E+06	-1.42E+07	-1.61E+07
Mass Heat Capacity	*kJ/kg-C*	1.90E+00	4.31E+00	4.31E+00	1.97E+00	0.00	0.00	0.00	0.00	1.92E+00	5.38E+00	4.33E+00	1.98E+00	1.75E+00	2.24E+00	2.01E+00	1.91E+00	2.18E+00	1.87E+00
Duty	*kW*					2.45	1593.11	1770.37	-1641.69										

CHAPTER 8

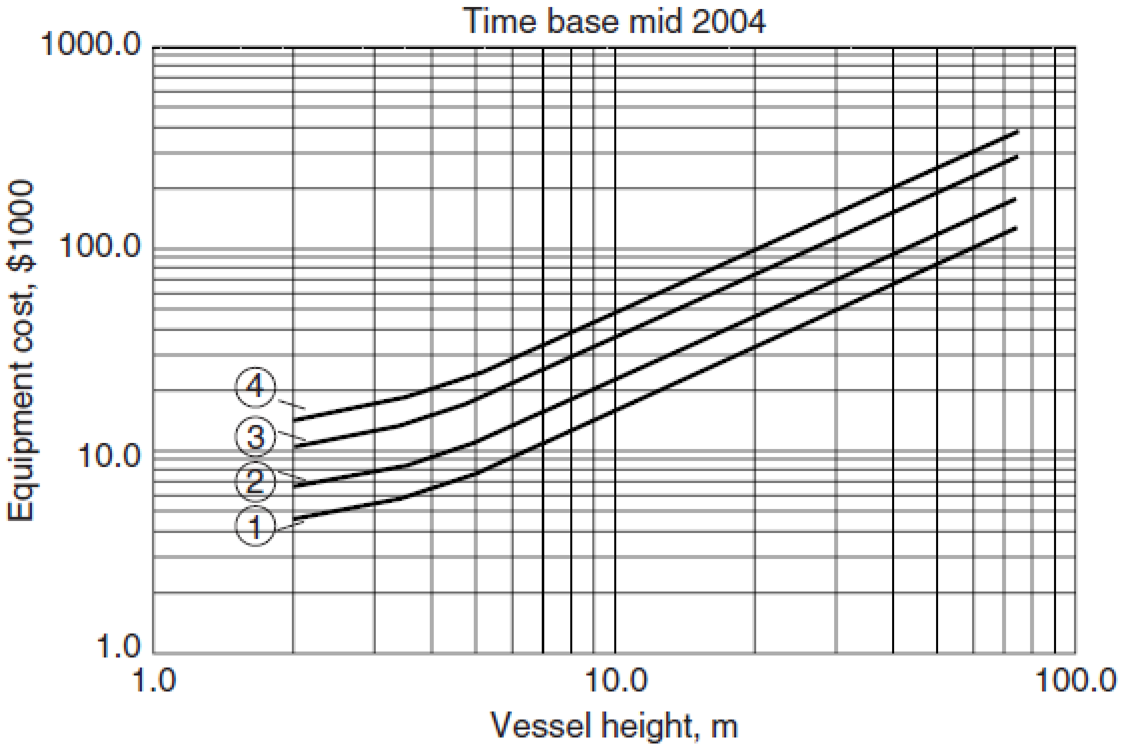
Economic Analysis

Costing of a project decides whether the project is feasible or not. Economic analysis is very important for the financially evaluate the project and all or its components. Since prime goal of a project is to do business and produce revenue So, that's why cost is a major factor for the decision of an equipment installation. Cost is one of the most important parameters in efficient management of a project. Time and quality are usually trade off with price of a project for desired results. We adapted Coulson and Richardson method for our Project economic analysis.

Table 34: Purchased equipment cost



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



(*b*) US dollars

Diameter, m

① — 0.5 ③ — 2.0
② — 1.0 ④ — 3.0

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

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

Figure 27: Vessel PCE

8.1 Desalter

Diameter (d) = 0.762 m

Length (l) = 2.286 m

- Bottom two curve’s data were interpolated for 0.762 m diameter.
- Since we used carbon steel material because of cost factor, therefore **material factor** is 1
- **Pressure factor** is also 1 for 1-5 bar pressure

Total cost of desalter =$ 24500

8.2 Flash Vessel

Diameter (d) = 1.524 m

Length (l) = 5.334 m

- Middle two curve’s data were interpolated for 1.524 m diameter.
- Since we used carbon steel material because of cost factor, therefore **material factor** is 1
- **Pressure factor** is also 1 for 1-5 bar pressure

Total cost of desalter =$ 102000

8.3 Stripper Column

Diameter (d) = 1.78 m

Length (l) = 9 m

- Middle two curve’s data were interpolated for 1.78 m diameter.
- Since we used carbon steel material because of cost factor, therefore **material factor** is 1
- **Pressure factor** is also 1 for 1-5 bar pressure

Total cost of desalter =$ 10700000

8.4 Rectifier Column

Diameter (d) = 1.78 m

Length (l) = 15 m

- Middle two curve’s data were interpolated for 1.78 m diameter.
- Since we used carbon steel material because of cost factor, therefore **material factor** is 1
- **Pressure factor** is also 1 for 1-5 bar pressure

Total cost of desalter =$ 17600000

8.5 Side Stripper 1

Diameter (d) = 1 m

Length (l) = 2.5 m

- Second curve’s data was calculated for 1 m diameter.
- Since we used carbon steel material because of cost factor, therefore **material factor** is 1
- **Pressure factor** is also 1 for 1-5 bar pressure

Total cost of desalter =$ 400000

8.6 Side Stripper 2

Diameter (d) = 1 m

Length (l) = 3 m

- Second curve’s data was calculated for 1 m diameter.
- Since we used carbon steel material because of cost factor, therefore **material factor** is 1
- **Pressure factor** is also 1 for 1-5 bar pressure

Total cost of desalter =$ 430000

8.7 Pumps

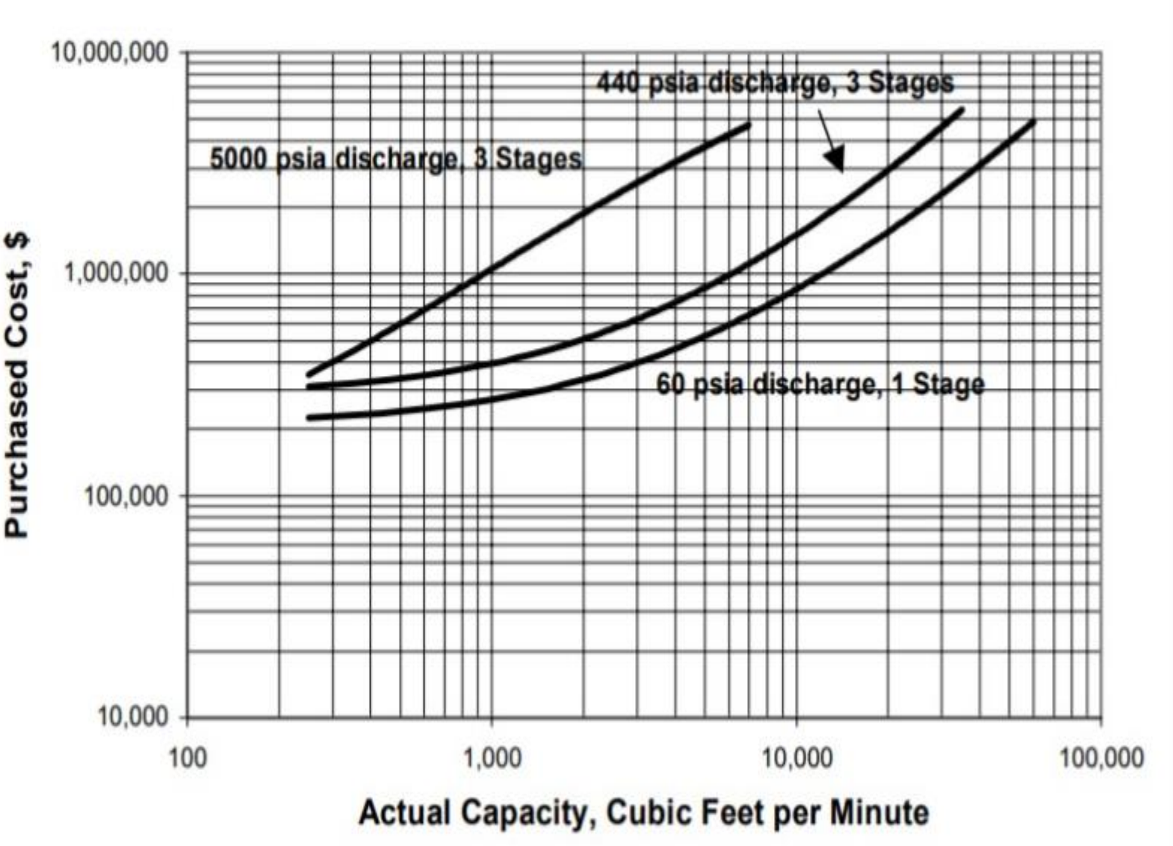


Figure 28: Pumps PCE

8.7.1 Pump 1

Actual Capacity: 112.5 ft^3/min

- Less than 60 psi discharge
- Single stage Centrifugal pump

Total pump cost = $ 28500

Cost index in 1998 = 5920

Cost index today = 6342

Cost in 2021 = 28500 x 6342/5920

Cost in 2021 = $ 30000

8.7.2 Pump 2

Actual Capacity: 112.5 ft^3/min

- Less than 60 psi discharge
- Single stage Centrifugal pump

Total pump cost = $ 28500

Cost index in 1998 = 5920

Cost index today = 6342

Cost in 2021 = 28500 x 6342/5920

Cost in 2021 = $ 30000

8.8 Compressor

Actual Capacity: 1593 ft^3/min

- Less than 70 psi discharge
- Single stage Centrifugal Compressor

Total Compressor cost = $ 587300

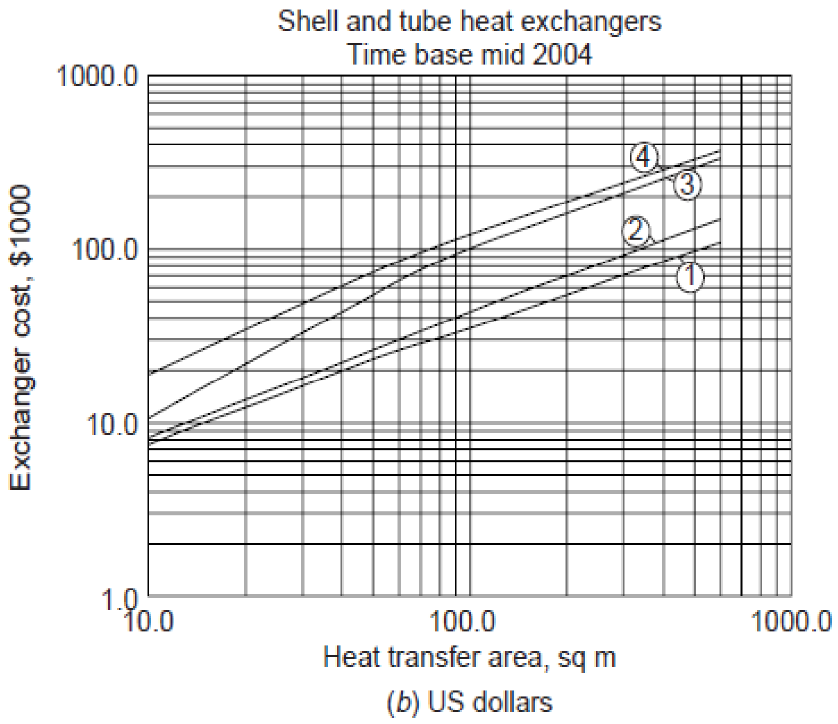
Cost index in 1998 = 5920

Cost index today = 6342

Cost in 2021 = 587300 x 6342/5920

Cost in 2021 = $ 699961

8.9 Heat Exchangers



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

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

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

Figure 29: Shell and Tube Heat Exchanger PCE

8.9.1 Heat Exchanger-1

Area = 33.6m2

Material Carbon Steel for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 0.85 for U tube

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (321000) x (1) x (0.85) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 344500

8.9.2 Heat Exchanger-2

Area = 34m2

Material Carbon Steel for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 0.85 for U tube

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (324000) x (1) x (0.85) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 346700

8.9.3 Heat Exchanger-3

Area = 23.9m2

Material Carbon Steel   for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 0.85 for U tube

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (223000) x (1) x (0.85) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 246500

8.9.4 Heat Exchanger-4

Area = 24.2m2

Material Carbon Steel   for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 0.85 for U tube

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (217000) x (1) x (0.85) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 248500

8.9.5 Heat Exchanger-5

Area = 23.9m2

Material Carbon Steel for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 0.85 for U tube

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (221000) x (1) x (0.85) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 236300

8.9.6 Heat Exchanger-6

Area = 26m2

Material Carbon Steel for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 0.85 for U tube

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (307000) x (1) x (0.85) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 256500

8.9.7 Heat Exchanger-7

Area = 34.4m2

Material Carbon Steel   for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 0.85 for U tube

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (327000) x (1) x (0.85) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 246500

8.9.8 Thermo Siphon Heat Exchanger

Area = 32 m2

Material Carbon Steel for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 0.85 for U tube

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (427000) x (1) x (0.85) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 546500

8.9.9 Stripper reboiler

Area = 49m2

Material Carbon Steel for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 1.1 for kettle type

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (227000) x (1) x (1.1) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 246500

8.9.10 Condenser

Area = 24 m2

Material Carbon Steel for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 0.85 for U-tube

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (27000) x (1) x (0.85) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 46500

8.9.11 Side Stripper 1- Heat Exchanger

Area = 14m2

Material Carbon Steel for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 1.1 for kettle type

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (127000) x (1) x (1.1) = $ 321000

Index in 2004 = 111

Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 146500

8.9.12 Side Stripper 2- Heat Exchanger

Area = 15 m2

Material Carbon Steel   for both shell and tubes

Material Factor = 1

Pressure factors = 1 for 1-10 bar pressure

Typer factor= 0.85 for U-tube

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Cost in 2004= (137000) x (1) x (0.85) = $ 321000

Index in 2004 = 111

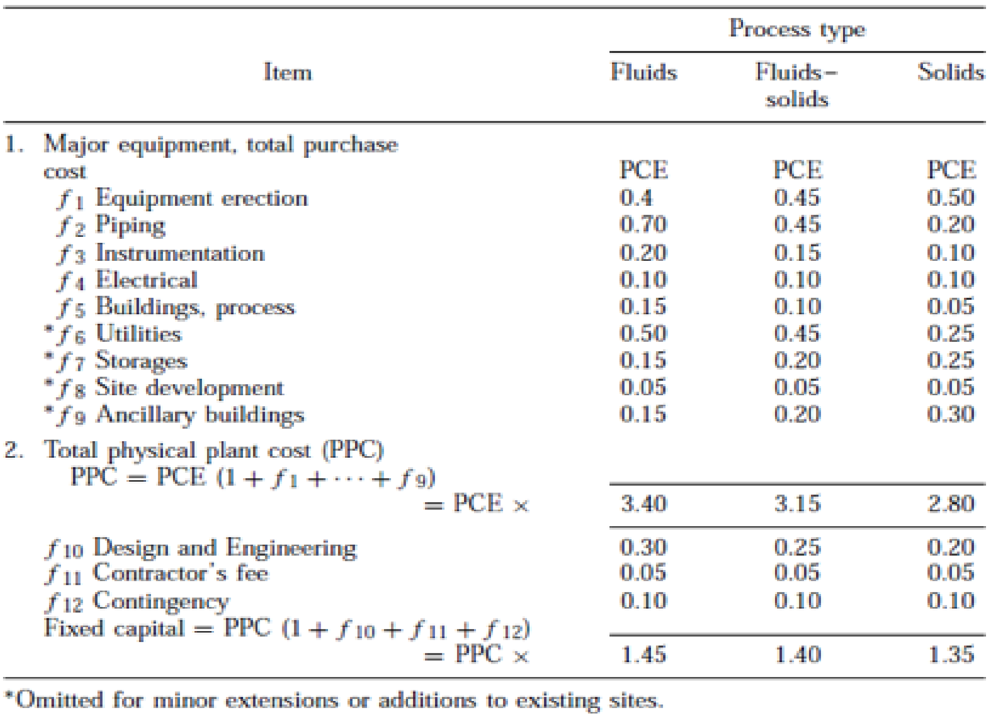
Index in 2021 = 172

$$Cost\ in\ 2020 = \frac{cost\ in\ 2004\ \times index\ in\ 2021}{index\ in\ 2004}$$

Cost in 2021 = $ 148000

8.10 Total Cost

Table 35: Fixed capital calculation:



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

*Omitted for minor extensions or additions to existing sites.

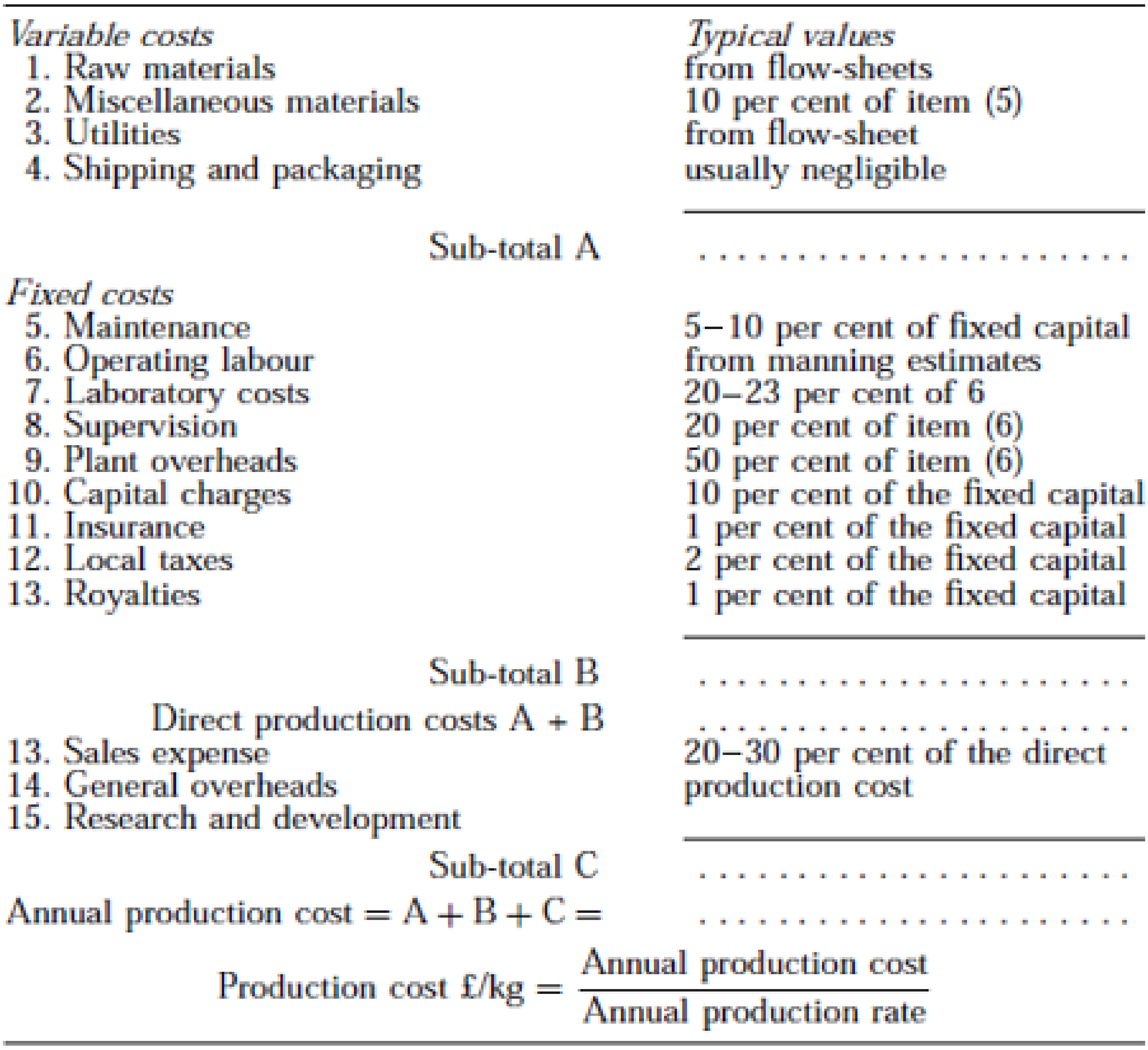
Table 36: physical plant cost

Physical Plant Cost	
Equipment Erection	0.4
Piping	0.7
Instrumentation	0.2
Electrical	0.1
Process	0.5
Storages	0.15
Physical Plant Cost	**$ 208 M**

Table 37: Total investment

Fixed capital	
Design and Engineering	0.3
Contingencies	0.05
Contractor's fees	0.1
Total	**$ 330 M**

Table 38: Variable cos



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

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

Fixed Operating Cost	
Maintenance	9.9M
Operating cost	5M
Labour	1M
Total	**15.9 M**
Variable Operating Cost	
Raw material	139 M
Miscellaneous	0.4 M
Utilities	0.5 M
Total	**140.9**

8.11 Economic Analysis Summary

Table 39: Purchased equipment cost summary

Purchased Equipment Costs	
Equipments	**Cost**
Desalter	**245000**
Flash Vessel	**102000**
Stripper	**10700000**
Rectifier	**17600000**
Side stripper 1	**400000**
Side stripper 2	**430000**
Pumps	**60000**
Compressor	**699961**
Heat Exchangers	
HX-1	**344500**
HX-2	**346700**
HX-3	**246500**
HX-4	**248500**
HX-5	**236300**
HX-6	**256500**
HX-7	**246500**
Thermo siphon HX	**546500**
Stripper reboiler	**246500**
SS-1 HX	**146500**
SS-2 HX	**495000**
Total	**$ 68 M**

Table 40: Economic analysis summary

Summary	
Purchased Equipment Cost (PCE)	68 M
Physical Plant Cost (PPC)	222 M
Fix Capital	330 M
Fix Operating Cost	15.9M
Variable Operating Cost	140.9M
Annual Operating Cost	156M
Annual Revenue	286M
Annual profit	129M
ROR	0.39
Payback Time	2.55 years

CHAPTER 9

Controls

9.1 Temperature Control Loop

- Control loops employed to keep process variables within specified range.
- Sensor picks up changes in parameter, sends signal to controller.
- Parameter compared with set point; controller acts accordingly.
- Temperature Control Loop has been installed to maintain Stream entering Pre-Flash at 150°C.
- Steam flow rate into heat exchanger varied to acquire desired output temperature.

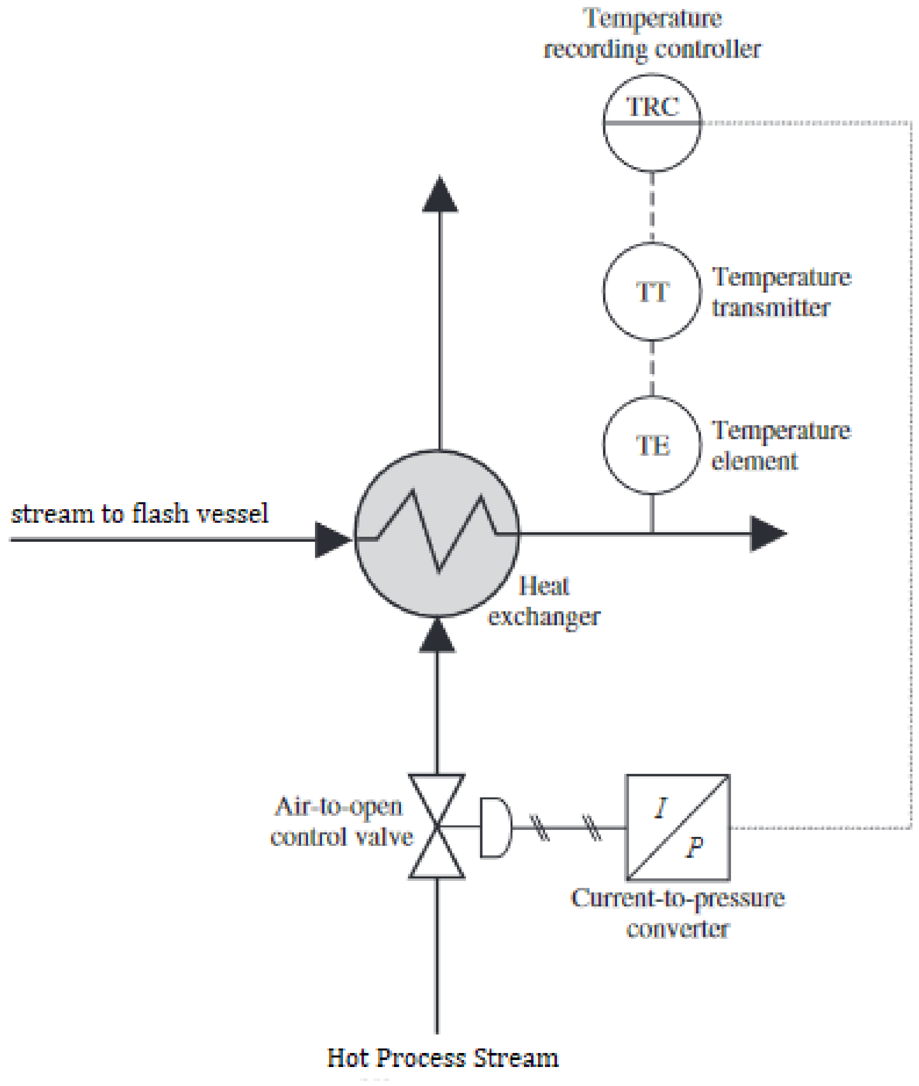


Figure 30: Temperature Control Loop

9.2 Flow Control Loop

- A ratio control loop is employed to maintain desired ratio of water flow rate to crude oil feed flow rate.
- Set point given to the controller.
- The flow rates are measured via flow sensors.
- Desired ratio achieved by varying flow rate of water.

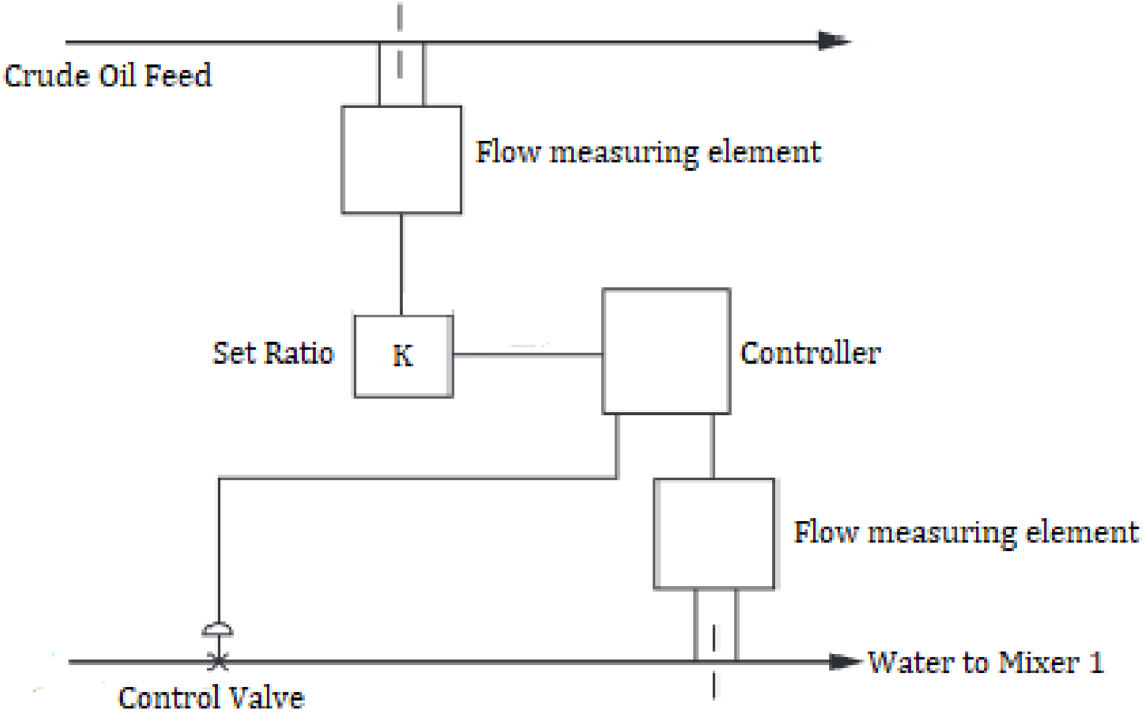


Figure 31: Flow Control Loop

9.3 Pressure Control Loop

- A pressure control loop is installed to monitor and control exit temperature of compressor.
- Control achieved by detecting deviation in pressure and then varying compressor energy input accordingly.

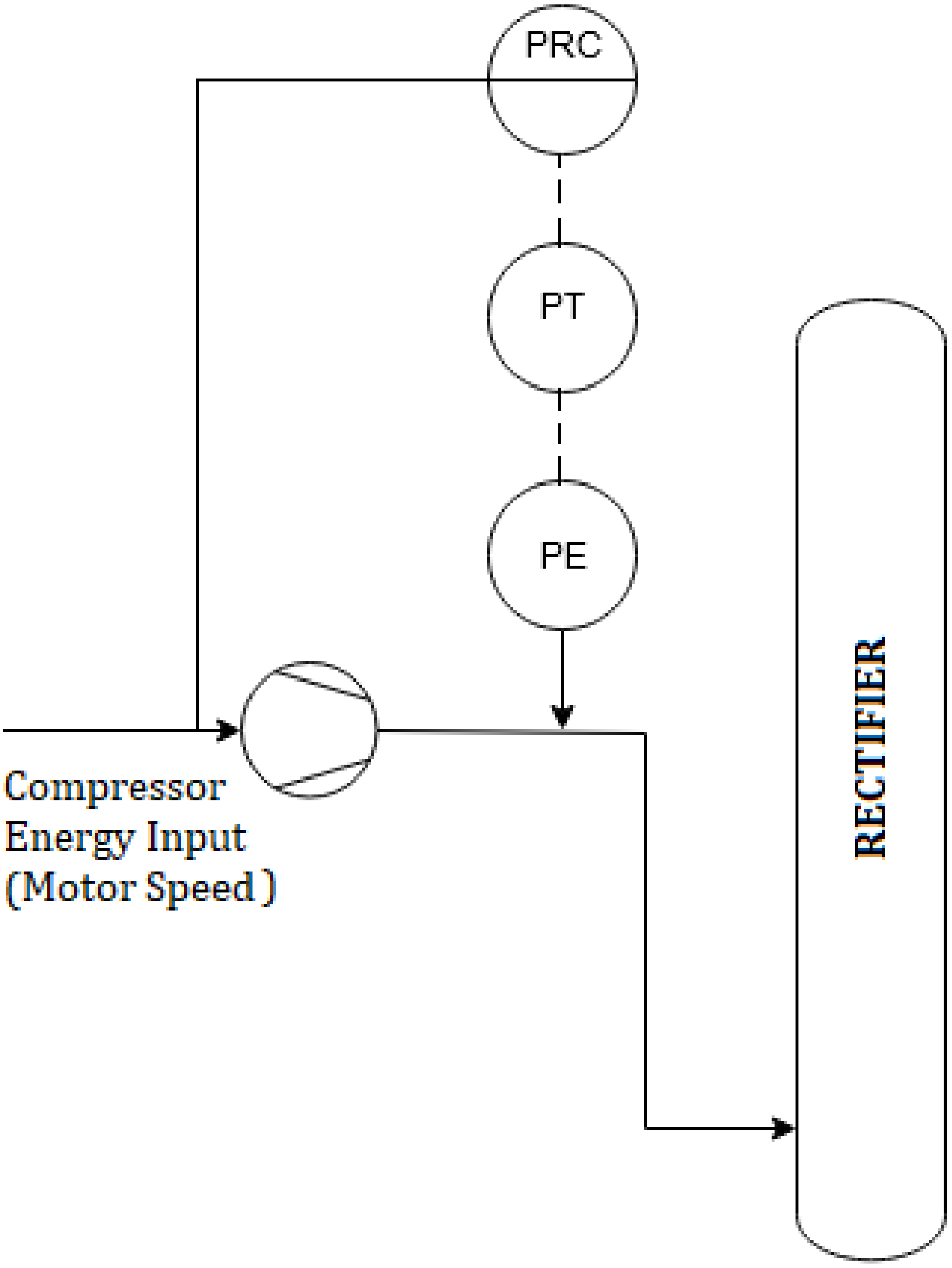


Figure 32: Pressure Control Loop

9.4 Level Control Loop

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

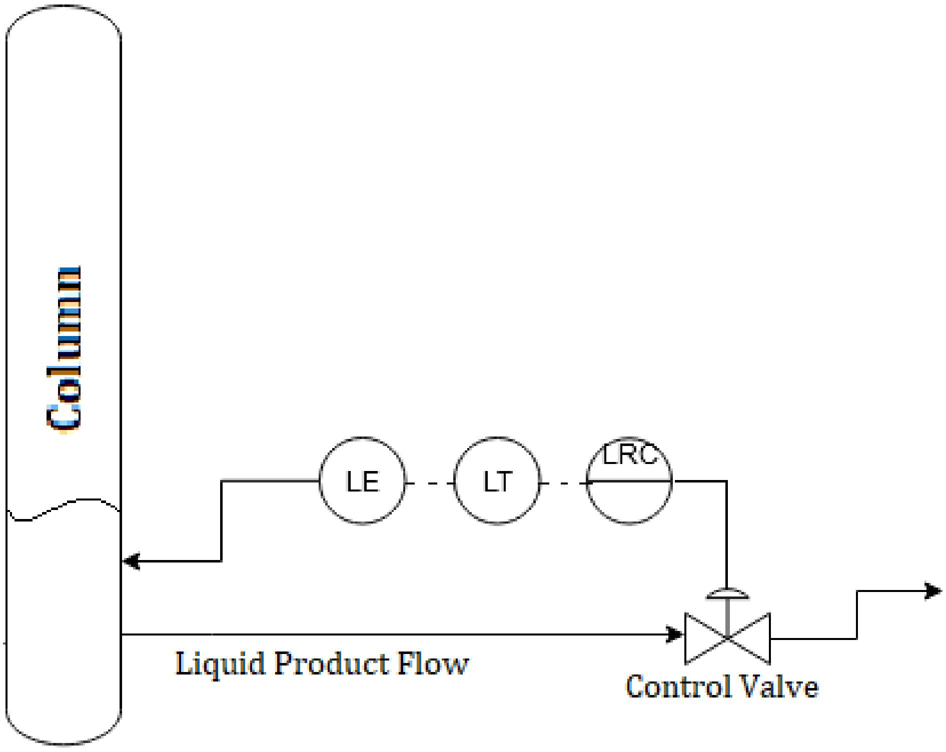


Figure 33: Level Control Loop

CHAPTER 10

HAZOP Analysis

A Hazard and Operability study of the proposed is fundamental to its technical feasibility from a safety point of view. Conventionally, such analyses usually involve using specific guidewords to assess the cause, effects, and preventative measures for the process. Examples of said guidewords may be “high temperature”, “low pressure” or something else of the sorts. These guidewords vary with the equipment under study within a system. Each component within a system must be considered for a thorough risk assessment [1]. The guidewords are essentially deviations that may potentially be encountered within the equipment and help to imagine several possible scenarios. HAZOP involves [2]:

1. Brainstorming what could possibly go wrong in the system. Listing down guidewords.
2. A qualitative risk assessment.
3. A “bottom-up” approach. Risk assessment relies on the observer’s predictive abilities and experience.

KEY WORDS [2]**:**

1. Nodes: Pipe sections or vessels containing where process chemicals may be present.
2. Super nodes: Several nodes considered together as one node for quicker studies.
3. Hazard: Potential source of harm.
4. Harm: Actual injury or damage to equipment, personnel, or environment.
5. Risk: Probability of harm occurring and its associated level of damage.

METHODOLOGY:

1. Identify all major nodes and super nodes.
2. Identify why the node was chosen for the HAZOP study.
3. Choose a key process parameter for said node.

4. Apply all relevant guidewords for chosen parameter. These would result in unwanted deviations within the process. List down the deviations.
5. Determine the potential causes of the deviations identified and add them to the table.
6. For the deviations brainstormed, identify the outcomes they would lead to. List these down as the consequences.
7. Once a node has been thoroughly studied, repeat steps 3 through 6 for the next node. Repeat this step till all chosen nodes have been accounted for.

10.1 PUMPS

Table 41: hazop analysis pumps

Guide Word	Deviations	Causes	Consequences	Actions Required
NO	No flow of crude oil	Outlet valve closed/blocked, pipeline rupturing, total pump failure	Overheating of pumps, bursting/explosion of pipeline, release of crude oil into the environment.	Adjustment of valves on the delivery lines, proper monitoring of pump and connected pipeline
LESS	Low flow of Crude Oil	Partial opening of the downstream valve, improper suction at inlet, pump cavitation.	Overheating of pump leading to failure, possible pressure buildup in pipelines.	Control system to vary valve opening according to flow requirements, priming of pump to avoid cavitation.
MORE	More flow of Crude oil than required	Fail open valve at delivery line.	Flooding of the following vessel.	Fail close valve at delivery with strict control of pump so that it closes. Level indicators on following vessel to identify overflow.

10.2 DESALTER

Table 42: hazop analysis Desalter

Parameter	Guide Word	Deviation	Causes	Consequences	Actions Required
Flow	NO	No incoming flow	Rupture of preceding pipeline.	Spillage of crude oil-desalting water mixture.	Strict monitoring of pipeline and turbomachinery leading up to (and away) from the vessel.
	LESS	Oil within the oil water mixture is less than required.	Disproportionate flow of water and oil into mixer (required ratio is not achieved)	Leads to inefficiencies in separation of salt from crude. More unwanted water within desalted oil stream.	Flow control on both water and crude oil streams so that the desired ratio for mixing may be maintained.
Level	MORE	Level of liquid in vessel higher than normal.	More than required inlet flows. Discharge pipelines are clogged.	Overflow of contents. Complete flooding of vessel	Cleaning of desalted crude oil exit pipes. Monitoring condition of succeeding pipelines

10.3 HEAT EXCHANGERS

Table 43: hazop analysis heat exchangers

Guide Word	Deviations	Causes	Consequences	Actions Required
LESS	Less flow of oil into Heat exchanger	Pipe blockages.	Higher temperature of exiting oil stream since steam flow rate remains unchanged.	Temperature sensors at outlet and implementation of control for steam and oil stream flow rates.
MORE	More than optimal oil flow to heat exchanger.	Complete opening of inlet valve upon failure.	Less than required temperature of exiting oil stream achieved since same amount of steam is being used to heat a bigger process stream. Higher	Temperature sensors at outlet and implementation of control for steam and oil flow rates. High pressure

			pressure on tube side. May lead to bursting of tubes.	alarms to halt operation if this happens.

10.3 PRE-FLASH SEPARATOR

Table 44:: hazop analysis Pre-Flash Separator

Parameter	Guide Word	Deviation	Causes	Consequences	Actions Required
Pressure	LESS	Low pressure within vessel	Low vessel temperature, Ruptured vessel (Leakage).	Lower vapor recovery since most of it would leak out.	Pressure sensors. Thorough vessel maintenance and repair.
	MORE	More than optimal pressure within the vessel.	High temperature, blockage of vapor outlet stream.	Bursting of vessel, damaging, nearby equipment, personnel, and environment	Pressure sensors, Pressure Relief, Thorough vessel maintenance and repair.
Level	LESS	Level of liquid in vessel less than optimal.	Less than required inlet flows. Discharge pipelines are clogged. Inlet stream too hot.	Improper separation of vapor and liquid phases.	Temperature & flow monitoring and control of inlet streams. Pipeline cleaning and maintenance.
	MORE	Level of liquid in vessel more than optimal.	More than required inlet flows. Blocked vapor outlet.	Improper separation of vapor and liquid phases. Risk of an explosion.	Temperature & flow monitoring and control of inlet streams. Pipeline cleaning and maintenance.

10.4 HIDC SYSTEM

Table 45: hazop analysis HIDIC

Parameter	Guide Word	Deviation	Causes	Consequences	Actions Required
Pressure	LESS	Pressure in Rectifier lower than required.	Improper function of compressor. Blockage at rectifier inlet.	Excessive heating of contents that remain in column (waste of heat), Inefficient separation	Scheduled service and maintenance of compressor. Checking for blockages at exit.
	MORE	Pressure in Rectifier is higher than required.	Blockage at exit Compressor speed increased.	Inefficient separation, Higher condenser duty.	Scheduled service and maintenance of compressor. Checking for blockages at exit.
Temperature	LESS	Low Temperature within Stripping Section.	Improper Reboiler operations. Improper preheating of feed.	Inefficiencies in separation. Flooding within column.	Check heat supplied by reboiler. Check heat exchangers before Stripper inlet.
	MORE	High Temperature within stripping Section.	Overheating by reboiler. Improper preheating of feed.	Inefficient separation. Thermosiphon system disturbed.	Check heat supplied by reboiler. Check heat exchangers before Stripper.

10.5 FLASH DRUM

Table 46: hazop analysis Falsh Drum

Parameter	Guide Word	Deviation	Causes	Consequences	Actions Required
Temperature	LESS	Temperature of separation is less than design value.	High cooling water flow in heat exchanger beforehand.	Improper separation of phases. Reduced Vapor flow. Flooding of vessel.	Adjusting cooling water supply to heat exchanger.
	MORE	Temperature of separation is more than design value.	Low cooling water flow in heat exchanger beforehand.	Leads to inefficiencies in separation. Lower Liquid flow rate.	Adjusting cooling water supply to heat exchanger.
Pressure	LESS	Pressure of separation is less than design value.	Low Temperature. Inlet pipelines are clogged.	Inefficient separation. Less than required product flow rates.	Checking preceding heat exchanger. Maintenance of inlet pipe.
	MORE	Pressure of separation is more than design value.	High Temperature. Outlet pipelines are clogged	Inefficient separation. Flooding and rupturing of vessel.	Checking preceding heat exchanger. Maintenance of inlet pipe.

CONCLUSIONS

1. **Increased naphtha yield:**

   The use of a light, sweet crude blend in a HIDC system has shown to give a naphtha yield of around 16,000 kg/h which is around 33%more than that which would have come out of a conventional column. This naphtha, when subjected to further downstream processing, can be used to increase gasoline production of the plant.

2. **Energy efficient operation:**

   Heat integration allows us to utilize the heat that would have been otherwise wasted in a conventional process. This leads to overall energy savings of 32%.

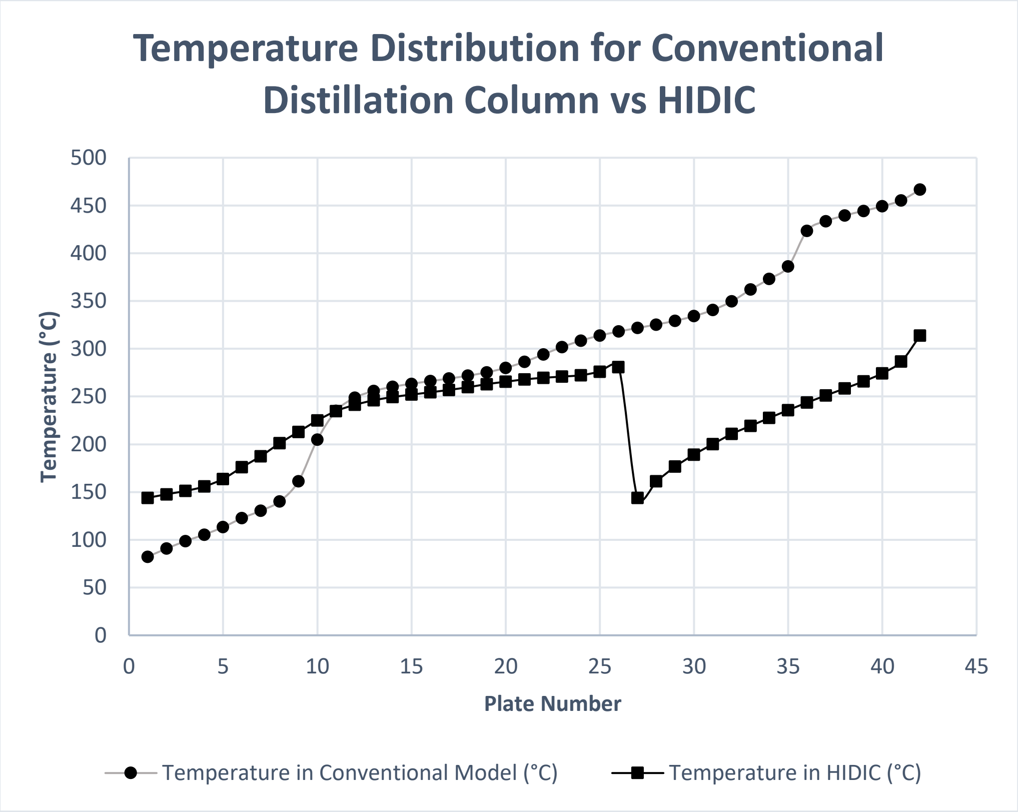


Figure 34: Temperature comparison of Conventional distillation column and HIDIC

3. **Economical, Sustainable Processing:**

   Because of better efficiency operating cost of HIDIC is lesser than conventional distillation column. Hence payback period for HIDIC is almost 5 months less than convectional system.

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