STUDY FOR TREATMENT & REMOVAL OF SALTS/SOLID DEPOSITION FROM E-300 TUBING/HEAT EXCHANGER

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CERTIFICATE

This is to certify that work in this thesis has been completed by **Mr. Mukhamad Zaman Chatkha** under the supervision of Dr. Erum Pervaiz at the School of Chemical and Materials Engineering (SCME), National University of Sciences andTechnology, H-12, Islamabad, Pakistan.

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DEDICATION

This FYP Thesis is dedicated to first of all, our parents, who had our back throughout our life and ourteachers, who have ever been the source of guidance for all of us.

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I want to acknowledge and thank the dedicated faculty at our School of Chemical and Materials engineering. This project would not have been possible without their constant support and guidance. In specific, **Dr. Erum Pervaiz** served as a source of guidance and encouragement for the entirety of this project. I want to express our gratitude to my industrial supervisor, **Mr. Aryab Bhayo**, for his continuous cooperation. I would also like to acknowledge Sir Taqi Mehran who offered a guiding hand. This thesis would not have been possible without Izza Afaq who has multiple times assisted me in many ways, Osama Shabbir and Muneeb Amjad.

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We are highly indebted to our parents for believing in us and for always providing us a shoulder to rely on.

ABSTRACT

Refining of crude oil is thorough and intricate process involving hundreds of steps in order to produce usable products. At the heart of every refining process on any level, heating of streams is one of the major unit operations involved. In most cases, this is done using utility streams through the use of heat exchangers. However, fouling of heat exchangers still persists as one of the biggest problems faced by industries. This issue is especially relevant to shell and tube type heat exchangers that have a higher chance of fouling as compared to some of its alternative designs like the plate and frame exchanger.

In high temperature applications with considerable number of salts and other fouling factors, scaling is justifiable with measures in place to limit this phenomenon. While this is the case for highly impure feed, other plants use a much cleaner type of crude oil which allows for a far less extensive pretreatment process. In such settings, when fouling develops over time, integration of harsher pretreatment equipment may turn out costly and time consuming. A simple yet effective method for such plants is the design and installation of desalters in order to separate salts and other foulants from the process stream which results in a considerably lower risk of fouling and reduces the maintenance cost of the heat exchangers used.

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CHAPTER 1

INTRODUCTION

1.1 Fossil Fuels

Over centuries of progress in human's history, energy has been a great necessity driving development survival and comfort of our race. While in older times, fire has been used as the prime source of heat and light, The development of methods for utilization of electricity have paved the way for discovering new sources of energy that could produce electricity on a mass scale. The discovery of fossil fuels and the development of process which allowed generation of energy from these carbon-based compounds in the 18th century, made them the prime drivers in the industrial revolution. Through many years of research and refinement, the method of extraction of energy from fossil fuels have evolved into several fields with individual processes at the core of each method suiting a different type of fossil fuel.

Fossil fuels are essentially the present in three main forms: coal, petroleum, and natural gas, all of which are resultants of millennia of alternating temperature and pressure on the remains of sea creatures and other animals to form carbon-based blends which we today extract and use as fuels.

Over the years the trend of fossil fuels varied with coal being the main fossil fuel of the 1700 to oil and gas being discovered in the 1800s and commercialized through use of cars in the 1900s. However even to date, fossil fuels still hold an important role in not only the energy sector but also in multiple chemical industries worldwide.

1.1.1 Oil Refineries

Oil is the main source of chemical products in the modern world; however, these goods cannot be fabricated from crude oil directly and are made from oil derivatives. These derivatives are produced in large oil processing plants known as oil refineries. These plants are generally present as the second stage in crude oil's transformation from a highly complex mixture of a multitude of components into usable products such as diesel, gasoline, petroleum jelly, etc. The main unit operation in these refineries is distillation or other separation techniques. These techniques require very specific conditions of temperature and pressure in order to achieve the desired separation. For the temperature constraint, many heating and cooling methods may be used; these include direct heating methods such as fire heaters and indirect ones such as heat exchangers which are the most common way of providing heat exchange for process streams.

1.1.2 Types of Oil Refineries

There exist a large array of different oil refineries mainly depending on the main processes and products that are present within. Some notable types include:

- Topping refinery
- Hydro Skimming refinery
- Conversion/cracking refinery

A topping refinery is the simplest refinery whose operation is based on basic refining usually resulting in two to three products such as light and heavy hydrocarbon blends. These products are by no means final and are usually sold to other local refineries that use more complex refining processes to make end-consumer goods. These refineries have a series of separators, distillation columns and heat exchangers in order to achieve a rough separation of the crude oil. The resultants of this refining end up as raw materials for other petrochemical plants or even other downstream refineries.

A step up from a basic topping refinery is a hydro skimming refinery which adds a few additional units to the basic set up. These refineries have additional reforming units that allow them to product some consumer end products such as gasoline, naphtha, etc. Unfortunately, due to limitations of the processes, the output of these plants is not categorized as high-quality petroleum products as they have relatively high Sulphur content which the consumer base is moving away from.

Conversion or cracking refineries have even further additions to hydro skimming refineries. As can be deducted from the name, cracking units or other gas oil conversion units in order to boost production of lighter hydrocarbon fuels with diesel and high-octane gasoline as products. These refineries also aim to limit residual fuels which are less profitable in comparison and more polluting.

1.1.3 Role of Heat Exchangers

As can be viewed above, regardless of the type of refinery, heat exchangers are required in order to bring the process streams to its required process temperature. Separation is usually a very temperature specific process and requires set temperatures to achieve the specified separation. Due to this, heat exchangers hold a major role in the refining process.

While many alternative methods are available for heating fluid streams, heat exchangers continue to be the widest spread and effective way of elevating temperatures. Over the past couple of decades, efficient energy management of energy streams have taken priority and the embodiment of that is the wide array of different designs provided for heat exchangers. As the applications of heat exchangers span from chemical and petrochemical industries to food and pharmaceutical industries, the feed varies accordingly. This means that the composition of the streams varies greatly from one industry to another.

1.1.4 Problem Identification

Irrespective of the industry of interest, suspended solids as well as dissolved solids are present in some quantity in all types of feed. For heat exchangers this brings about the most prominent and common problem across all industries: fouling. Regardless of the extensive research and understanding of heat transfer coefficients which allow us to transfer energy between fluids at the highest efficiency, fouling and scaling the same issues faced by the primary designs of heat exchangers plague the industry today. Although complete prevention of this phenomena is impossible limitation of it is an option through using pretreatment methods and other techniques.

1.2 Problem Statement

The objective of this paper is to examine methods for prevention or treatment of fouling within E-300 heat exchanger. As described above, even in the modern world fouling is still at the forefront of the hurdles faced by most refineries. This is especially true for topping refineries as they lack conventional pretreatment methods when the type of crude is sweet. Sweet crude is characterized by a very low Sulphur content as well as lower salt concentrations. In the case of topping refineries that primarily use this type of crude for processing, pretreatment operations such as CSTRs, liquid-liquid extraction columns and desalters are absent as removal of these compounds is not a requirement of their buyers. Thus, fouling does accumulate over time but simple cleaning of heat exchanger over time is sufficient.

However, due to change of feed composition over time which may arise from changing of the source well or the general deterioration of heat exchanger performance over the years of use may increase the frequency of these cleanings which lead to a massive spike in the maintenance costs. In order to counter this action, other preventive measures need to be undertaken to limit the extent of fouling in the heat exchanger.

1.3 Pakistan Petroleum Limited

The project objective given has been provided by Pakistan Petroleum limited. PPL is a topping refinery that operates in Pakistan on 13 crude oil and natural gas production sites. PPL contributes to roughly 20% of Pakistan total oil and natural gas supply and is involved in extraction operations as well. PPL is a topping refinery which means that it produces large amounts of roughly refined crude oil which are ultimately produced as two streams to downstream refineries like Attock Refinery Limited (ARL) and OGDCL as a raw material.

The operation/process of PPL consists of the following leading equipment, Distillation column, heat exchangers, separators, mixers, and gas recovery facilities. The plant given to me has all of the following equipment with a shell and tube heat exchanger responsible for the heating of the heavy hydrocarbon stream prior to the distillation column. The plant capacity given is small relative to other PPL sites with an almost Sulphur-less crude oil composition of API 45°.

1.4 Research On Fouling Mitigation

As mentioned previously, fouling is impossible to prevent completely but several mitigation techniques exist. Through proper understanding of the duty and design of the heat exchanger as well as the nature of the fouling, we can design an effective way to reduce this problem which subsequently minimizes our maintenance costs which potentially go up to 50000 USD annually. A thorough understanding of the stream composition and the presence of potential foulants allows us to make realistic assumptions about the type of fouling, and helps us reduce the undesirable phenomena considerably.

The main techniques that could be utilized are split into two major categories: on the line treatment and off the line treatment. Both are to be explored in the literature review section of this thesis.

CHAPTER 2

LITERATURE REVIEW

2.1 Heat Exchanger

Heat exchangers are devices made specifically to promote the exchange of heat between two fluid streams with different temperatures; the streams mentioned may be in direct contact or indirect contact with each other. The more commonly used heat exchanger on an industrial level is the indirect contact/surface type heat exchanger which allows miscible fluids to exchange energy with each other. This is not the case for direct heat exchangers which promote a far better exchange process due to lack of a physical barrier between the two fluids but requires the two streams to be immiscible in order toto prevent the contamination of one stream with the other.

The presence of a physical boundary between the two streams adds for an additional resistance to energy flow between the two fluids and also promotes fouling but the upside of keeping the two streams apart is immense.

The two main principles which drive heat exchange in these devices are conduction and convection. So, it isn't a surprise that these are the main parameters of design for any heat exchanger.

2.1.1 Types of Heat Exchangers

As described in section 2.1, heat exchangers are of two broad types, the direct contact and Surface type. Both are in use in the industries worldwide with the process streams of most refineries utilizing surface type.

2.1.2 Direct Contact Heat Exchangers

At the start of the industrial revolution, direct contact heat exchanger drove the progress of the petrochemical industry through its first major steps. However, the lack of complete understanding and modeling of the process our comprehension of direct contact heat exchangers is on a much lower level as compared to the surface type exchangers which are today used in every major plant for a variety of heating purposes. Regardless of this, direct contact heat exchangers have still been used historically and are still widely in use today.

2.1.3 Surface Type Heat Exchangers

Surface type heat exchangers are characterized by the fact that there is no direct contact between the two fluids which exchange heat. The limitations of direct contact heat exchangers were that contamination and mixing of the fluids to some degree is almost impossible to prevent, and this problem was tackled in the design of surface heat exchangers. There are quite a few numbers of surface heat exchanger types but the most important and widely used ones are presented below.

- Shell and Tube heat exchanger
- Plate type heat exchanger
- Double Pipe heat exchanger
- Condensers and boilers

2.1.4 Shell and Tube heat exchangers

These heat exchangers are the most common one used industrially. The basic construction of these devices is a bundle of parallel tube running through a pressure vessel known as a shell. The tubes offer a channel of passage for one of the fluids while the shell gives way for the flow of the other fluid.

Contact area for heat transfer in these designs is the surface area of the tubes which with a provided temperature gradient allows heat to flow from one fluid to the other. The basic configuration could be boosted using fins in order to promote a greater surface area allowing for a higher rate of energy transfer between the two streams.

This type of exchanger may be operated in many modes such as single pass or double pass, co-current or counter current, and even multiple phase heat transfer.

2.1.5 Plate type heat exchanger

This device consists of a series of corrugated plates place back-to-back with each other. The plates form channels between each other which allow fluids to pass through. In an alternating formation, the plates allow the fluids to pass through making the area of the plate the contact area for each heat transfer. Up to many hundreds of plates may be placed together which allows for an immense area to be present within a single module. While the upside of the high surface area seems great, this device cannot be operated under high pressure conditions which makes this a huge limitation. The pressure limit is set in order to prevent leaking of the fluids through the gaskets and losing the streams.

2.1.6 Double Pipe heat exchanger

A subcategory of the shell and tube heat exchanger, the double pipe exchanger is constructed based on two concentric pipes with one of the fluids flowing through the internal pipe and the other finding its way down the annulus. The design of these is extremely basic and effective with the only constraint is the limitation of contact area provided by the single internal pipe. The modules can be put in parallel and series configuration to allow for a larger contact area.

2.1.7 Boilers and Condensers

These are a special type of heat exchangers which are designed specifically for two phase heat transfer. The conditions of the streams require one of both streams to undergo phase change and thus the designs need to be considered accordingly. In condensers, the phase change is usually from gaseous to liquid while in boiler it’s the exact opposit

2.2 Fouling in Heat Exchangers

Fouling is the result of build up of a solid layer of foulants on the contact area of the heat exchanger. While many factors contribute to fouling from multiple sources, the main force that drives fouling is the presence of suspended solids or dissolved solids in one of the two fluid streams. This means that the buildup adds onto the already present surface of the exchanger retarding the rate of heat transfer across the contact area. In addition, the layer increases the pressure drop across the heat exchanger which requires higher pumping power to ensure the same pressure at the exit stream. Since the resistance to heat transfer is also increased, the stream entry conditions are to be changed as well in order to reach the required stream parameters.

There are several additional factors that influence the extent of fouling, and these can be analyzed in order to limit or mitigate fouling in our heat exchanger.

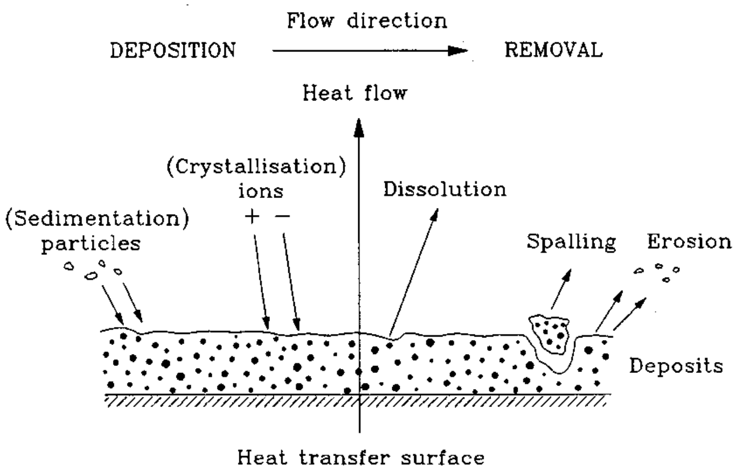


Figure 1: Fouling Mechanisms

Fouling types can be categorized into a few main types listed below. The major types include: particulate, precipitation, corrosion and biofouling. All of these will be explored further.

2.2.1 Types of Fouling

Important types of fouling.

- Particulate fouling is a type of fouling that is built up due to the suspended solid in the process stream. This type of fouling can usually be mitigated by pretreatment methods such as coagulation and flocculation.
- Precipitation fouling also known as crystallization-based fouling occurs due to dissolved salts present in the stream. This type of fouling is highly dependent on the temperature conditions as well as the flow rate of the stream which determine the shear stress on the deposition. Desalting of the stream will generally be sufficient to prevent this type of fouling.
- Corrosion fouling is the result of inorganic reactants in the process stream that react with the surface material in order to form an extra coat of deposition on the surface known as corrosion. These reactants are generally metal ions or Sulphur present within the process stream. Another type of fouling that could fall under this category is scaling which is formed due to the presence of carbonate compounds in the stream.
- Biofouling is the result of microscopic biological organisms that deposit on the surface and grow. The growth of this type of fouling is dependent upon the conditions of the stream and if it allows the survival of the bacteria and nutrition that allows further growth of the microorganisms. This type of fouling is common in food processing industries.

2.3 Mitigation of Fouling

Fouling mitigation can be achieved through multiple ways depending on the type of fouling that is present within the exchanger. The broad categorization of fouling prevention techniques is on the line prevention and off the line prevention.

2.3.1 On the Line Fouling Mitigation

On-line prevention techniques are those that do not require the removal of heat exchanger from its operation to execute. This type of prevention or cleaning is usually done periodically in order to limit fouling and prevent off-line cleaning which takes up considerable amount of time.

- The primary method for this category is change of operating conditions at the top of which is the temperature conditions. While most plants have processes designed to run under very specific temperature, if there is any room for change, this is the simplest way to influence the rate of fouling. Further techniques that could be used include changing up the flow of the exchanger to pulsating flow. This allows the shear stress applied on the walls of the exchanger to be increased allowing deposition build up to be removed and prevented over time.
- Mechanical methods can also be used in form of projectiles or brushes or even increase of the surface resistance to fouling using different materials during the design stage.
- Physical mitigation methods include surface coating, ultrasonic or electromagnetic. While ultrasonic and electromagnetic isn't as popular on an industrial scale, surface coating is a technique that has been utilized with excellent results. However, this method is held in check by the stream conditions in order to ensure no reaction between the fluids and the coating. Ultrasonic mitigation has been conducted on a lab scale and even in industrial tests with absurd results proving that it might be the best technique to date, but industrial applications are extremely limited in today's heat exchangers.

2.3.2 Off the Line Fouling Mitigation

Off the line cleaning is only done after the performance of the exchanger has shown a significant dip due to fouling and none on the line techniques work. Severe Chemical and mechanical cleaning can be done here or even replacement of elements such as tubes or plates.

2.3.3 Pre-treatment of Feed

The best way to combat fouling is to prevent it from forming in the first place. This is done through multiple processes that treat the feed and the utility stream in order to ensure the absence of any foulants in those fluids in the first place.

Several equipment and processes exist to this extent such as desalter, CSTRs to counter suspended solids and even LLE to extract liquid foulants.

2.4 Three Phase Separator

The three-phase separator is a cylindrical device that separates the components based on density. A stream containing a gaseous phase and two liquid phases can enter a three- phase separator and is given a particular retention time. The gaseous phase leaves from thetop. The two liquid phases are separated based on density. Lighter fluid leaves from the middle section while the heavy liquid leaves from the bottom. The three-phase separators are most commonly used in petroleum refineries to separate oil from gases and water.

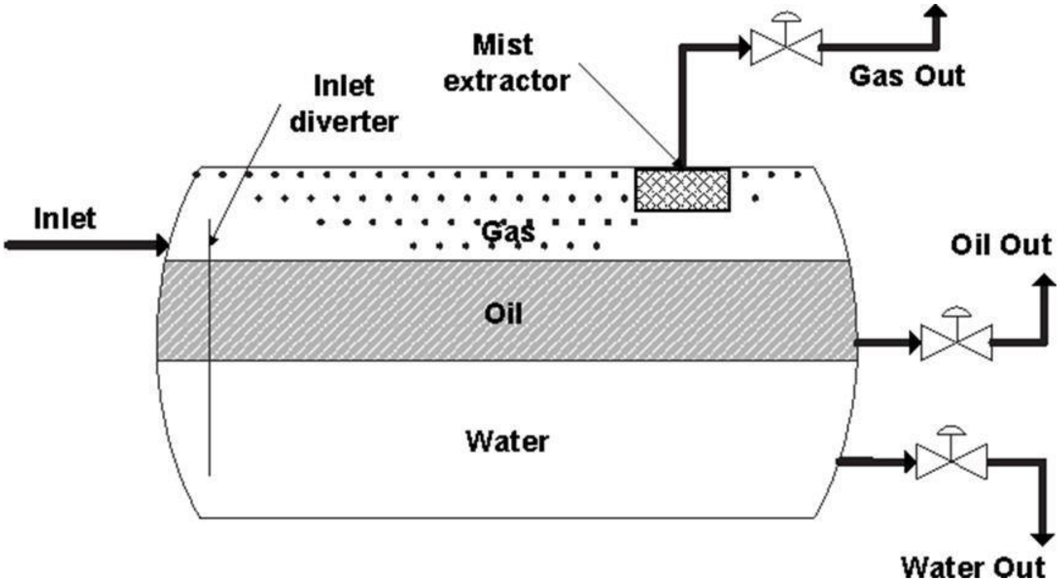


Figure 2:Three phase separator

The three-phase separator can also be designed to work as a desalter which saves the cost of additional equipment and provides and acceptable removal of dissolved solids. The dissolved solids are separated from the main process hydrocarbon stream and carried away with the water outlet in the bottom. This prevents the deposition of the dissolved solids altogether.

CHAPTER 3

PROCESS DESCRIPTION

The process is initiated from the crude oil feed which is pumped to a separator which flashes the feed to give us two different streams based on the density and volatility of the components. This equipment, however, is replaced by a 3-phase separator which is also used as a desalter. Since water is also present in the crude oil stream in minor quantities, this isn't enough for our desalter separation. Thus, an additional fresh water stream is added in order to achieve more effective separation of salts in the 3-phase separator. The separator also performs the duty of the previously installed two phase separator and gives us the primary separation of hydrocarbons into two streams.

This gives us the initial separation of the crude oil and ensures that minimum to no salt is present in the proceeding process streams. The streams are then taken to their respective processing equipment described below.

3.1 Cooler and KBR Cool Box

The lighter hydrocarbon stream is taken through a cooler, separator and a KBR cool box or purifier. The KBR purifier is a patented technology which allows the stream to reach temperatures of as low as -182°C while performing ongoing separation of the components.

3.2 Expander and Separator

After further splitting of the light hydrocarbon mixture, the main process stream is taken to the separator 3 while the top part is expanded in order to further lower the temperature of the stream. The two are mixed and enter the final separator 3 from which our top is pure light hydrocarbon product spanning to propane and the lower is taken to the distillation column for further separation.

3.3 Heat Exchangers and the Distillation column

The heavier hydrocarbon stream out of the 3-phase separator is introduced into the shell and tube heat exchanger which heats it to the required temperature of 54°C. It is then taken towards the distillation column where it is separated with the final heavy product stream sent towards the plate and frame heat exchanger for reheating of the bottom stream coming out of separator 3 and further being entered into the distillation column. The shell and tube heat exchanger uses a utility stream in order to heat the process stream.

3.4 Expander

After separator 2, one of the streams is directly introduced to the mixer, whereas the other stream passes through an expander before going to the mixer. The aim of this is to achieve a lower temperature in our process stream and make separation possible. This also allows some energy to be extracted out of the turbine.

3.5 Mixer

The two streams (top product stream and the distillate of the column) sent to the mixer are extracted as final light products. This final light product is then mixed with the final light products of the 13 fields operated by PPL and its partner organizations to produce an equivalent of 1 billion cubic feet equivalent (bfce) of natural gas.

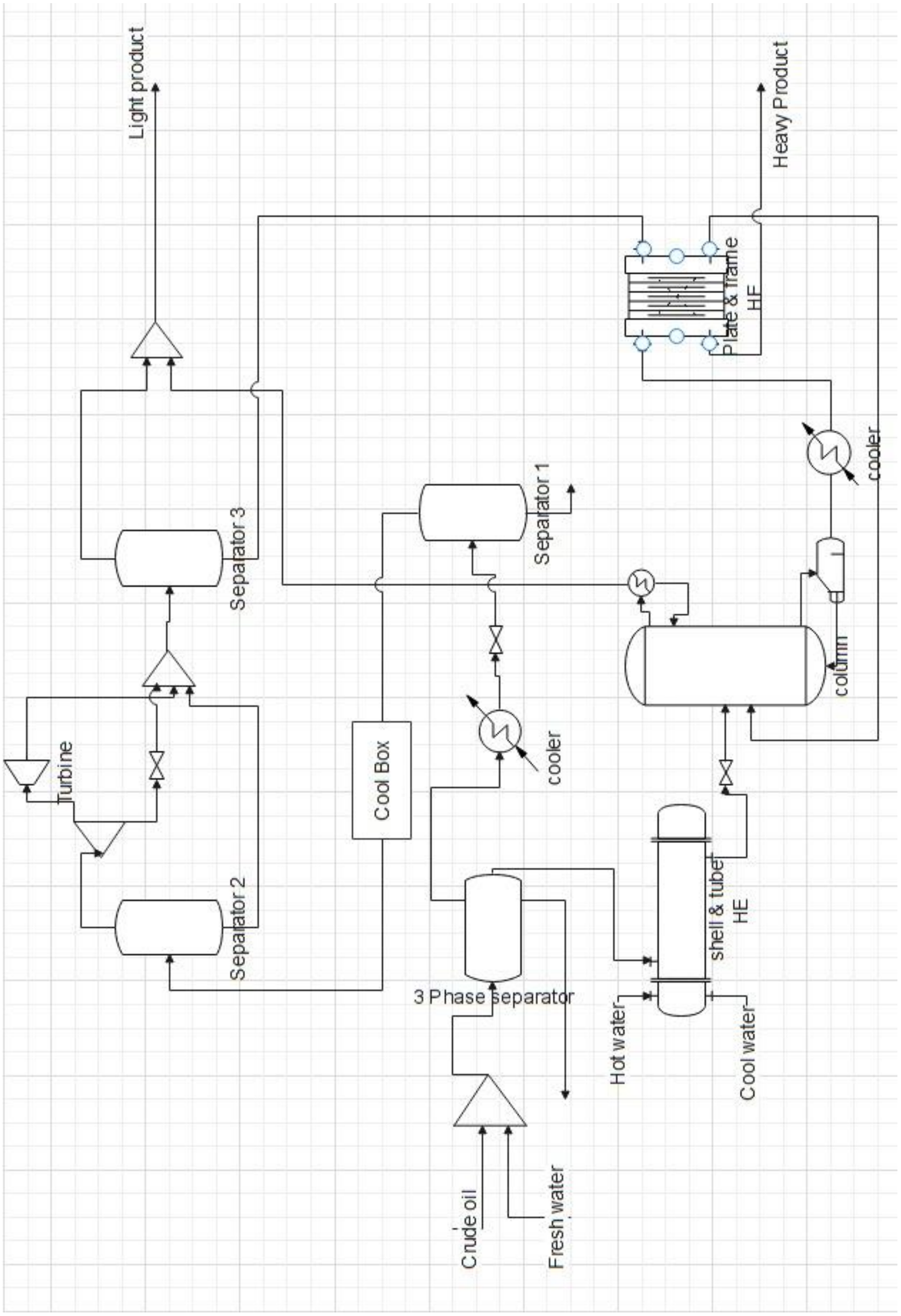


Figure 3: Process Flow Diagram

CHAPTER 4

MATERIAL BALANCE

The Law of conservation of mass states that mass can neither be created nor be destroyed. The modified version of this based for chemical engineers is the material balance, also known as mass balance. This balance studies the inlet and outlet materials in order to ensure that whatever amount of mass that enters is the same as the mass leaving. However, from a chemical engineering point of view, mass balance allows us to find out much more by taking a basis of time and studying the compositions and reactions to find out the flow rates of the outlet streams as well as their makeup.

The general mass balance equation can be seen below:

Accumulation within the system = (Mass in the system) – (Mass out of the system) + (Generation within the system) – (Consumption within the system)

As all of the equipment in the given plant are unit operations, no generation or consumption can be assumed which leaves us with

Mass in the system = Mass out of the system

The material balance of our process is given below(mole fractions of components are given):

4.1 3-Phase Separator/Desalter

Table 1: Conditions of 3-phase separator streams

	Inlet	**Light**	**Heavy**	**Water**
Flowrate (kg/hr)	41600	21086.7	20509.7	3.61

Temperature (^{0}C)	25	25	25	25
Pressure (kPa)	6250	6250	6250	6250

Table 2: Material Balance on 3-Phase Separator

Component	**Inlet**	**Light**	**Heavy**	**Water**
H_2O	**0.0175**	**0.0006**	**0.0005**	**1**
CO_2	**0.0051**	**0.0056**	**0.0036**	**0.0001**
N_2	**0.0138**	**0.0169**	**0.0019**	**0**
CH_4	**0.7215**	**0.8514**	**0.2471**	**0**
Ethane	**0.0798**	**0.0794**	**0.0883**	**0**
Propane	**0.0386**	**0.0283**	**0.0846**	**0**
Iso-butane	**0.0104**	**0.0053**	**0.0324**	**0**
n-butane	**0.0155**	**0.0067**	**0.0535**	**0**
Iso-pentane	**0.0076**	**0.0315**	**0.0210**	**0**
n-pentane	**0.0074**	**0.0016**	**0.0320**	**0**
Hexane	**0.0118**	**0.0011**	**0.0570**	**0**
Heptane	**0.0123**	**0.0005**	**0.0621**	**0**
Octane	**0.0141**	**0.0003**	**0.0730**	**0**
Nonane	**0.0098**	**0.0001**	**0.0511**	**0**
Decane	**0.0076**	**0**	**0.0395**	**0**
C_{11+}	**0.0271**	**0**	**0.1418**	**0**
Sulphur	**0.0002**	**0**	**0**	**0.0155**
Total	**1**	**1**	**1**	**1**

4.2 Separator 1

Table 3: Conditions of separator 1 streams

	Inlet	**Top**	**Bottom**
Flowrate (kg/hr)	19250	19190	68
Temperature (^{0}C)	12.04	12.04	12.04
Pressure (kPa)	6216	6216	6216

Table 4: Material Balance on Separator 1

Component	**Inlet**	**Top**	**Bottom**
H_2O	0.0006	0.0003	0.2509
CO_2	0.0056	0.0031	0.0056
N_2	0.0169	0.0169	0.0016
CH_4	0.8514	0.8523	0.2079
Ethane	0.0794	0.0794	0.0790
Propane	0.0283	0.0283	0.0785
Iso-butane	0.0053	0.0053	0.0305
n-butane	0.0067	0.0066	0.0510
Iso-pentane	0.0020	0.0019	0.0302
n-pentane	0.0016	0.0015	0.0310
Hexane	0.0011	0.0059	0.0545
Heptane	0.0005	0.0022	0.0557
Octane	0.0003	0.0007	0.0557
Nonane	0.0001	0.0001	0.0288
Decane	0	0	0.0142
C11+	0	0	0.0274
Total	1	1	1

4.3 Separator 2

Table 5: Conditions of separator 2 streams

	Inlet	Top	Bottom
Flowrate (kg/hr)	19190	15660	3053
Temperature (^{0}C)	-44	-44	-44
Pressure (kPa)	6166	6166	6166

Table 6: Material Balance on Separator 2

Component	***Inlet***	***Top***	***Bottom***
H_2O	*0.0003*	*0*	*0.0023*
CO_2	*0.0056*	*0.0053*	*0.0077*
N_2	*0.0169*	*0.0187*	*0.0049*
CH_4	*0.8523*	*0.8920*	*0.5842*
Ethane	*0.0794*	*0.0648*	*0.1784*
Propane	*0.0283*	*0.0151*	*0.1175*
Iso-butane	*0.0053*	*0.0018*	*0.0290*
n-butane	*0.0066*	*0.0018*	*0.0392*
Iso-pentane	*0.0019*	*0.0003*	*0.0129*
n-pentane	*0.0015*	*0.0002*	*0.0107*
Hexane	*0.0011*	*0.0001*	*0.0079*
Heptane	*0.0005*	*0*	*0.0034*
Octane	*0.0002*	*0*	*0.0014*
Nonane	*0*	*0*	*0.0003*
Decane	*0*	*0*	*0.0001*
C11+	*0*	*0*	*0.0001*
Total	*1*	*1*	*1*

4.4 Separator 3

Table 7: Conditions of separator 3 streams

	Inlet	Top	Bottom
Flowrate (kg/hr)	19190	14390	4791
Temperature (^{0}C)	-63.85	-63.85	-63.85
Pressure (kPa)	3700	3700	3700

Table 8: Material Balance on Separator 3

Component	*Inlet*	*Top*	*Bottom*
H2O	*0.0003*	*0*	*0.0018*
CO2	*0.0056*	*0.0048*	*0.0092*
N2	*0.0169*	*0.0198*	*0.0029*
CH4	*0.8523*	*0.9209*	*0.5150*
Ethane	*0.0794*	*0.0479*	*0.2347*
Propane	*0.0283*	*0.0058*	*0.1387*
Iso-butane	*0.0053*	*0.0004*	*0.0294*
n-butane	*0.0066*	*0.0003*	*0.0376*
Iso-pentane	*0.0019*	*0.0000*	*0.0112*
n-pentane	*0.0015*	*0.0000*	*0.0090*
Hexane	*0.0011*	*0.0000*	*0.0063*
Heptane	*0.0005*	*0.0000*	*0.0027*
Octane	*0.0002*	*0.0000*	*0.0011*
Nonane	*0.0000*	*0.0000*	*0.0002*
Decane	*0.0000*	*0.0000*	*0.0001*
C11+	*0*	*0*	*0*
Total	*1*	*1*	*1*

4.5 Distillation Column

Table 9: Conditions of Distillation column streams

	Inlet 1	Inlet 2	Top	Bottom
Flowrate (kg/hr)	4791	17890	5982	16700
Temperature (^{0}C)	20	54	2	266.1
Pressure (kPa)	1800	5122	3350	3450

Table 10: Material Balance on Distillation column

Component	*Inlet 1*	*Inlet 2*	*Top*	*Bottom*
H2O	9.23E-03	3.59E-03	9.76E-03	1.07E-13
CO2	2.88E-03	1.94E-03	3.84E-03	4.48E-22
N2	0.515035	0.247115	0.590339	7.22E-15
CH4	0.234718	8.83E-02	0.245401	4.03E-08
Ethane	0.138679	8.46E-02	0.147851	4.40E-02
Propane	2.94E-02	3.24E-02	7.77E-04	7.77E-02
Iso-butane	3.76E-02	5.35E-02	3.59E-04	0.118503
n-butane	1.12E-02	3.15E-02	8.06E-06	5.86E-02
Iso-pentane	9.01E-03	3.20E-02	3.02E-06	5.72E-02
n-pentane	6.31E-03	5.70E-02	9.18E-08	9.16E-02
Hexane	2.66E-03	6.21E-02	1.95E-09	9.54E-02
Heptane	1.10E-03	7.30E-02	4.34E-11	0.109998
Octane	2.47E-04	5.11E-02	6.05E-13	7.65E-02
Nonane	5.45E-05	3.95E-02	9.15E-15	5.90E-02
Decane	4.54E-05	0.141819	4.95E-16	0.211499
C11+	1.79E-03	4.54E-04	1.66E-03	1.66E-13
Total	*1*	*1*	*1*	*1*

4.6 Mixer

Table 11: Conditions of Mixer streams

	Inlet	Top	Bottom
Flowrate (kg/hr)	14390	5981	20370
Temperature (^{0}C)	-63.85	3.063	-43.92
Pressure (kPa)	3700	3350	3350

Table 12: Material Balance on Mixer

Component	*Inlet*	*Top*	*Bottom*
H2O	4.85E-03	9.76E-03	5.98E-03
CO2	1.98E-02	3.84E-03	1.61E-02
N2	0.920898	0.590339	0.844814
CH4	4.79E-02	0.245401	9.33E-02
Ethane	5.82E-03	0.147851	3.85E-02
Propane	4.13E-04	7.77E-04	4.97E-04
Iso-butane	3.36E-04	3.59E-04	3.41E-04
n-butane	3.44E-05	8.06E-06	2.83E-05
Iso-pentane	1.83E-05	3.02E-06	1.48E-05
n-pentane	3.14E-06	9.18E-08	2.44E-06
Hexane	3.39E-07	1.95E-09	2.62E-07
Heptane	3.64E-08	4.34E-11	2.80E-08
Octane	2.27E-09	6.05E-13	1.75E-09
Nonane	1.48E-10	9.15E-15	1.14E-10
Decane	3.34E-11	4.95E-16	2.57E-11
C11+	4.38E-07	1.66E-03	3.82E-04
Total	1	1	1

CHAPTER 5

ENERGY BALANCE

The Law of conservation of energy states that energy can neither be created nor be destroyed but can be converted from one form to another. Energy balance gives us the way to analyze systems and confirm that energy has been conserved. In a chemical system, this balance is usually done on the basis of heat flow or enthalpy. Enthalpy is defined as the total heat content of the stream. The enthalpies of inlet and outlet streams of the systems can be added and compared in order to see if all of the energy is conserved within the streams or if a part of that energy has been converted into either work or heat. In most chemical operations, the kinetic and potential energy is unchanged so they can be omitted from the balance.

Energy balances are also used to get an idea about the power required to run a plant. The General Energy balance equation is:

$$Q - W_s = \Delta H + \Delta K.E. + \Delta P.E.$$

At a steady-state the heat flow of a stream can be calculated as:

$$Q = mC_p \Delta T$$

The energy balance of our process is shown below neglecting any change in kinetic and potential energy:

$$Q - W_s = \Delta H$$

5.1 Shell and tube Heat exchanger

The energy balance on the shell and tube heat exchanger can simply be applied as this:

Energy in= energy out

$Q = \dot{m} C_p \Delta T$ heat received by stream

$\dot{m}$ = 17890 kg/hr
C_p = 2.323 kJ/kg^{0}C
T_{in} = 25^{0}C
T_{out} = 54.9^{0}C

Q = 17890 x 2.323 (54.9-25)
= 1242598 kJ/hr
= 345.1 kW

$Q = \dot{m}\, C_p\, \Delta T$ heat provided by stream

$\dot{m}$ = ?
C_p = 4.3 kJ/kg^{0}C
T_{in} = 90^{0}C
T_{out} = 50^{0}C

1242598 = $\dot{m}$ x 4.3 (90-50)
$\dot{m}$ = 7224 kg/hr

5.2 Distillation Column

5.2.1 Condenser

Energy in = energy out

$$Lh_L = Rh_r + Dh_D + Q_C$$

	Stream *kg/hr*	*Enthalpy of Stream* *kJ/kg*
(L) To condenser	19330	-2995.6
(D) Distillate	5982	-3640.2
(R) Reflux	13350	-30906

Q = 1423 kW

5.2.2 Reboiler

Energy in = energy out

$$Th_T + Q_R = Bh_B + Ih_I$$

	Stream *kg/hr*	*Enthalpy of Stream* *kJ/kg*
(T) To reboiler	71860	-1731.9
(B) Bottom	16700	-1536.9
(I) Boilup	55160	-1512.4

Q = 2191.7 kW

5.3 Plate and Frame Heat exchanger

Energy in = energy out

$Q = \dot{m}\, C_p\, \Delta T$
$\dot{m} = 16700$ kg/hr
$C_p = 2.492$ kJ/kg^{0}C
$T_{in} = 120^0$C
$T_{out} = 64.8^0$C

$Q = \dot{m}\, Cp\, \Delta T$
$= 16700 \times 2.492 \times 55.2$
$= 2297225$ kJ/hr
$= 638.1$ kW

5.4 Expander

Energy in = energy out

	Stream Flow kg/hr	h_1 (Kj/kg)	h_2(KJ/kg)
Through expander	15660	-4552	-457

$W_t = \dot{m}\,(h_1 – h_2)$
= 15660 x (-4552 – 4575)
= 360180 kg/hr
= 100.05 kW

CHAPTER 6

PROCESS DESIGN

6.1 Distillation Column/Crude Stabilizer

6.1.1 Feed Stream

Mass Flow = 16058.33 kg/hr = 3.906 MMSCFD

Table 13: Mole fractions of feed stream

	Feed 1	Feed 2	Resultant feed
Flow rate(kg/hr)	16058	5891	21949
Components			
H2S	9.23E-03	3.59E-03	0.000
CO2	2.88E-03	1.94E-03	0.006
Nitrogen	0.515035	0.247115	0.0023
Methane	0.234718	8.83E-02	0.3647
Ethane	0.138679	8.46E-02	0.1510
Propane	2.94E-02	3.24E-02	0.01052
i-Butane	3.76E-02	5.35E-02	0.03
n-Butane	1.12E-02	3.15E-02	0.044
i-Pentane	9.01E-03	3.20E-02	0.021
n-Pentane	6.31E-03	5.70E-02	0.021
n-Hexane	2.66E-03	6.21E-02	0.0340
n-Heptane	1.10E-03	7.30E-02	0.0363
n-Octane	2.47E-04	5.11E-02	0.0430
n-Nonane	5.45E-05	3.95E-02	0.03
n-Decane	4.54E-05	0.141819	0.023
n-C11+	1.79E-03	4.54E-04	0.085
H2O	9.23E-03	3.59E-03	0.00

After defining the characteristics of our feed stream, we select our key components, i-pentane, and Propane

Propane recovery can be estimated from the given, and we assume pentane recovery

For i-pentane, we assume that in the top product, 0.001 i-pentane is present, and in the bottom product, 0.999 of i-pentane is present

Table 14: Distribution of LKC and HKC

Component	**Top Fraction**	**Bottom Fraction**
Propane	0.9252	0.0747
I-pentane	0.001	0.999

6.1.2 Applying mass balance on individual components

- **Propane**

Feed = Top product + Bottom product

0.01052 x 21949 = 0.9252 x Top product + 0.0747 x Bottom Product

- **I-Pentane**

Feed = Top product + Bottom product

0.0217 x 21949 = 0.001 x Top product + 0.999x Bottom Product

0.01052 x 21949 = 0.9252 x Top product + 0.0747 x Bottom Product

925.2 x (0.0217 x 21949) = (0.001 x Top product + 0.999x Bottom Product) x 925.2

Solving the equations simultaneously gives us the following values:

Bottom product=8130.4 kg/hr
Top product=13678.3 kg/hr

We can assume the light and heavy non key components in order to model pseudo binary distillation.

Ethane, methane, nitrogen, and CO2 can be assumed to be light non key components.

i-Butane, n-Butane, n-Pentane, n-Hexane, n-heptane, n-Octane, n-Nonane, n-Decane, and n-C11 are assumed as heavy non key components.

6.1.3 Operating Temperatures

Depriester charts can be used in order to estimate the bubble and dew point of the distillation column. Since a large number of components are present, a one decimal point error is permitted.

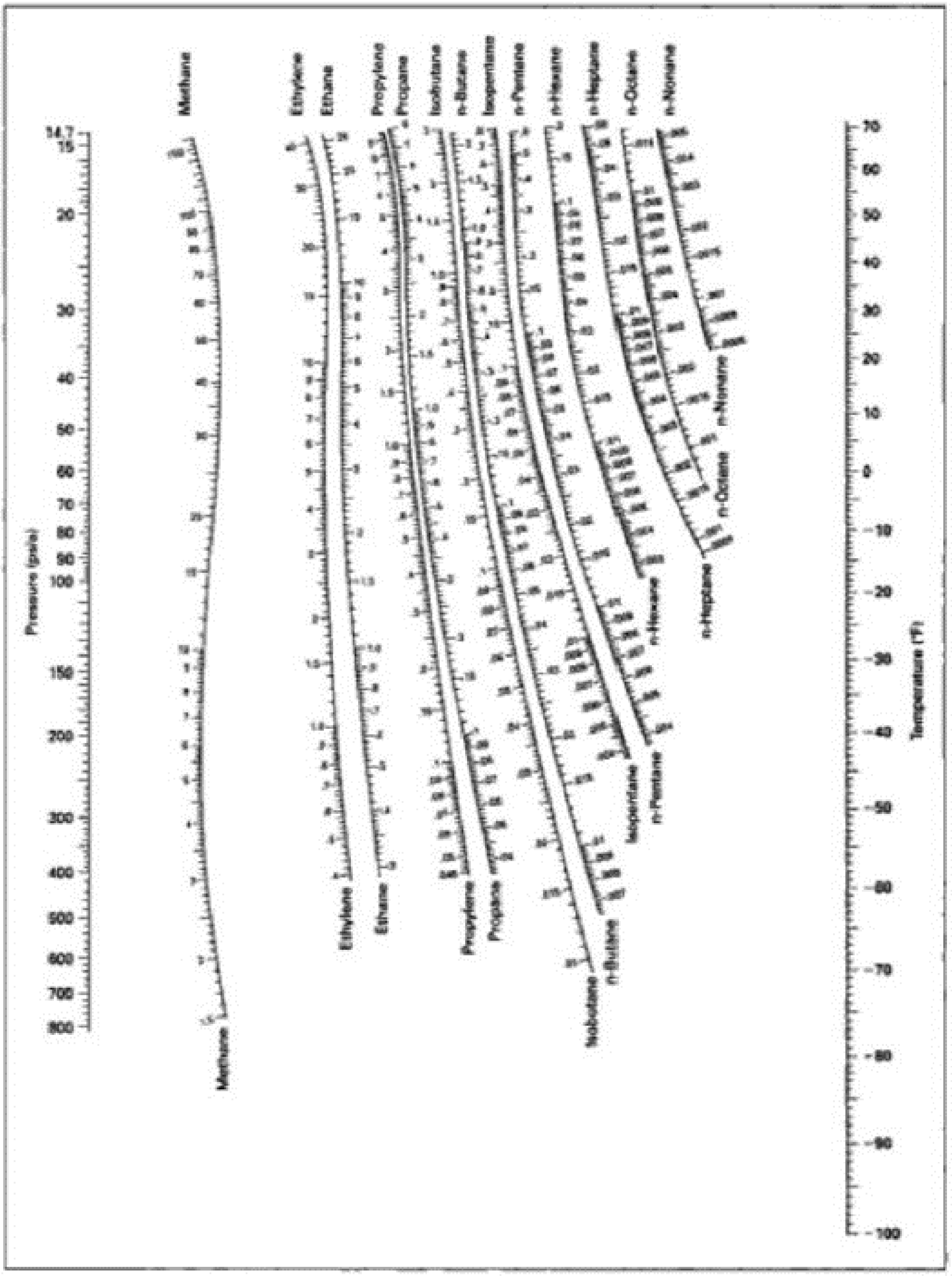


Figure 4: Depriester Chart

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

This relation could be used to calculate the bubble point

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

This relation can be used for dew point calculations

The bubble point of our feed stream comes out to be 2 ^{0}C, and the dew point of ourfeed stream comes out to be 241.9 ^{0}C

The calculations of underwood's equations via hit and trial would take far too long with so many components thus an alternative path has been taken.

6.1.4 Distribution coefficients

Now for multi-component distillation, we specify our distribution coefficients to define it

$M_i = \frac{P_i*}{p}$ (For any component I)

To devise a relationship between relative volatility and the distribution coefficient

$M_{ij} = \frac{Pvi}{Pvj}$

6.1.5 Relative volatility

To find the relative volatility (α) of our key components, we need to find their vapor pressure (P_v) or saturation pressure (P_s). We will find the P_s or P_v by using the Antione equation

$$lnPv = \frac{A}{C - T_k}$$

For Propane

P^v = 65.88 bar = 6500 Kpa

The constants A, B, and C are taken from Milo D, Korestky Engineering & Chemical Thermodynamics Book.

For i-pentane

P^v= 11.52 bar = 1152 Kpa

Relative Volatility=$\frac{65}{11.52}$ =5.7187 relative volatility of light key

6.1.6 Minimum Reflux Ratio

After calculating relative volatility, we will use its value to calculate the minimum refluxratio using the underwood and Fenske equation. This short cut distillation method is used due to the large number of components which would prove difficult in calculations.

$$Rm = \frac{1}{a-1}[\frac{X_{Da}}{X_{Fa}} - a\frac{X_{Db}}{X_{Fb}}]$$

a = propane

b = i=pentane

X_{Da} = fraction of propane in top product = 0.952

X_{Fa} = fraction of propane in feed = 0.1052

X_{Db} = fraction of i-pentane in top product = 0.001X_{Fa} = fraction of i-pentane in feed = 0.0217

Minimum Reflux ratio = R_m = 1.80

Since we have a multi-component distillation column, we cannot use the VLE diagram tofind out the number of plates as it becomes complex and inaccurate.

To calculate the theoretical number of plates required, we use Gilliland's empirical relation. For this relation, we need to know the values of the Minimum reflux ratio (R_m), Actual reflux ratio (R), and the Minimum number of plates (N_m). Using these, we can find the number of theoretical plates required.

Using the general rule of thumb, the value of K can be estimated between 1.3 and 1.8 for the value of R_m 1.80.

Assuming the value of K to be 1.3R= K x R_m

R = 1.3 x 1.8

R= 2.34

6.1.7 Minimum number of plates

Now to find the minimum number of plates at total reflux, we use Fenske's equation

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

propane X_{LK} = 0.92 $X_{LK(B)}$ = 0.0747

i-pentane X_{HK} = 0.001 $X_{HK(B)}$ = 0.999

$$\boldsymbol{N_m = \frac{log\left[\frac{0.92}{0.001}\right]_d log\left[\frac{0.999}{0.0747}\right]_b}{log\, 5.7}}$$

N_m = 5.4 = 6 (minimum number of plates)

For Gilliland's co-relation graph, we find the values below for R of 2.34

$$\frac{R - Rm}{R + 1} = 0.16$$

Using the Gilliland co-relation graphs

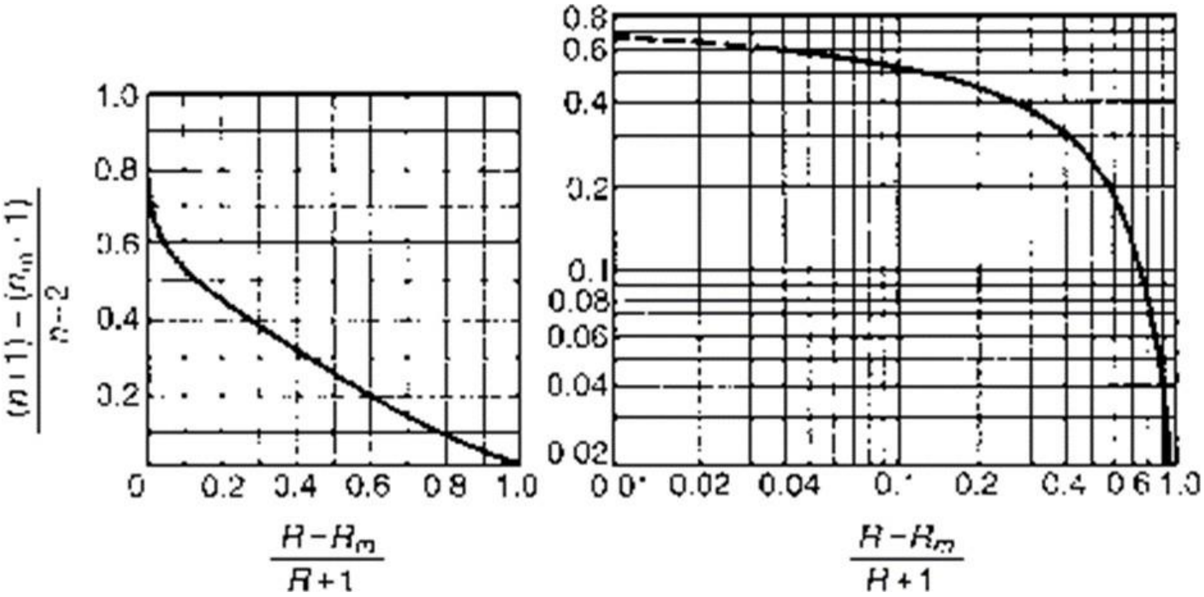


Figure 5: Gilliland co-relation Graph

From the curve we obtain the value of $\frac{(n+1)-(Nm+1)}{N+2}$ = 0.5

As N_m = 0.5N

N = 12 theoretical stages are required.

6.1.8 Feed entry location

The feed point of the column can be calculated using Kirkbird’s empirical relation

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

Inputting the values specified, we get the relation of

$$\mathbf{N_r = 1.1\ N_s}$$

We already know that the total number of plates is 12 so, another relationship can be shown.

$$N_r + N_s = 11$$

Solving the two simultaneously, we get:

Nr= 8

So, the feed should enter at the 8th plate.

6.1.9 Column Diameter

Column diameter is calculated using the relationship specified below

$$u_v = \left(-0.171 l_t^2 + 0.27 l_t - 0.047\right)\left[\frac{\rho_l - \rho_v}{\rho_v}\right]^{0.5}$$

L_t is the plate spacing which is taken to be 0.5 m

The densities are provided below:

ρ_l= 325.2 kg/m^3

ρ_v= 46 kg/m^3

$$u_v = (-0.171 \times 0.5^2 + 0.27 \times 0.5 - 0.047)\left[\frac{325.2 - 46}{46}\right]^{0.5}$$

$$u_v = 0.112 m/s$$

$$V_w = 2.3\ kg/s$$

Use the formula below to approximate the diameter.

$$D_C = \left(\frac{4V_w}{\pi \rho_v u_v}\right)^{0.5}$$

D_c= 1.6 m

6.1.10 Column Efficiency

The column efficiency can be calculated using O'Connell correlation

μ = 0.1939 cp

α = 5.7

Corresponding to this, E_o = 0.68

So, the number of plates is given by: 12/0.68= 24 plates

6.2 Shell and Tube Heat Exchanger

Table 15: Shell and tube heat exchanger conditions

Hot Fluid(Water)		Cold Fluid(Hydrocarbon)
Inlet Temp(T_1) ^{0}C	90	30
Exit Temp(T_2) ^{0}C	50	54
Mass Flow Rate(W) (kg/hr)	4.89×10^3	1.606×10^4
C_p (kJ/kg.^{0}C)	4.2	2.295
Specific gravity	1	0.63
Thermal Conductivity (W/m.K)	0.67	0.098
Viscosity (cP)	0.39	0.238

6.2.1 Heat Duty

$$Q = mc_p \Delta T$$

Q=353.5 KW

6.2.2 Operating Temperature

$$\mathrm{LMTD} = \frac{\Delta t_2 - \Delta t_1}{2.3\log\left(\frac{\Delta t_2}{\Delta t_1}\right)}$$

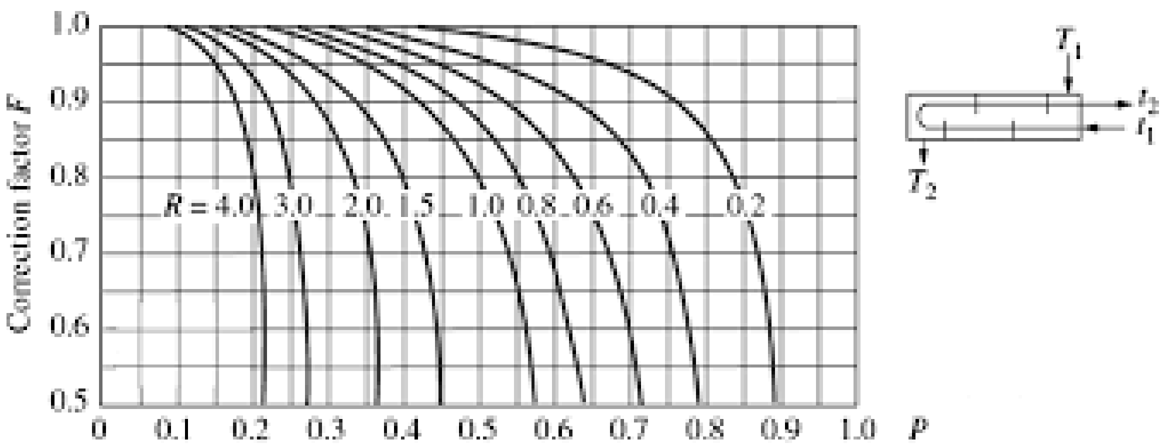
Co-current flow is used.

$$\mathbf{LMTD} = \frac{65-6}{2.3\log\left(\frac{65}{6}\right)} = 24.76°\mathrm{C}$$

$$R = \frac{T_h - T_c}{t_c - t_h} = 1.03$$

$$P = \frac{t_c - t_h}{T_h - t_c} = 0.44$$

For a single shell pass and 2 tube passes,



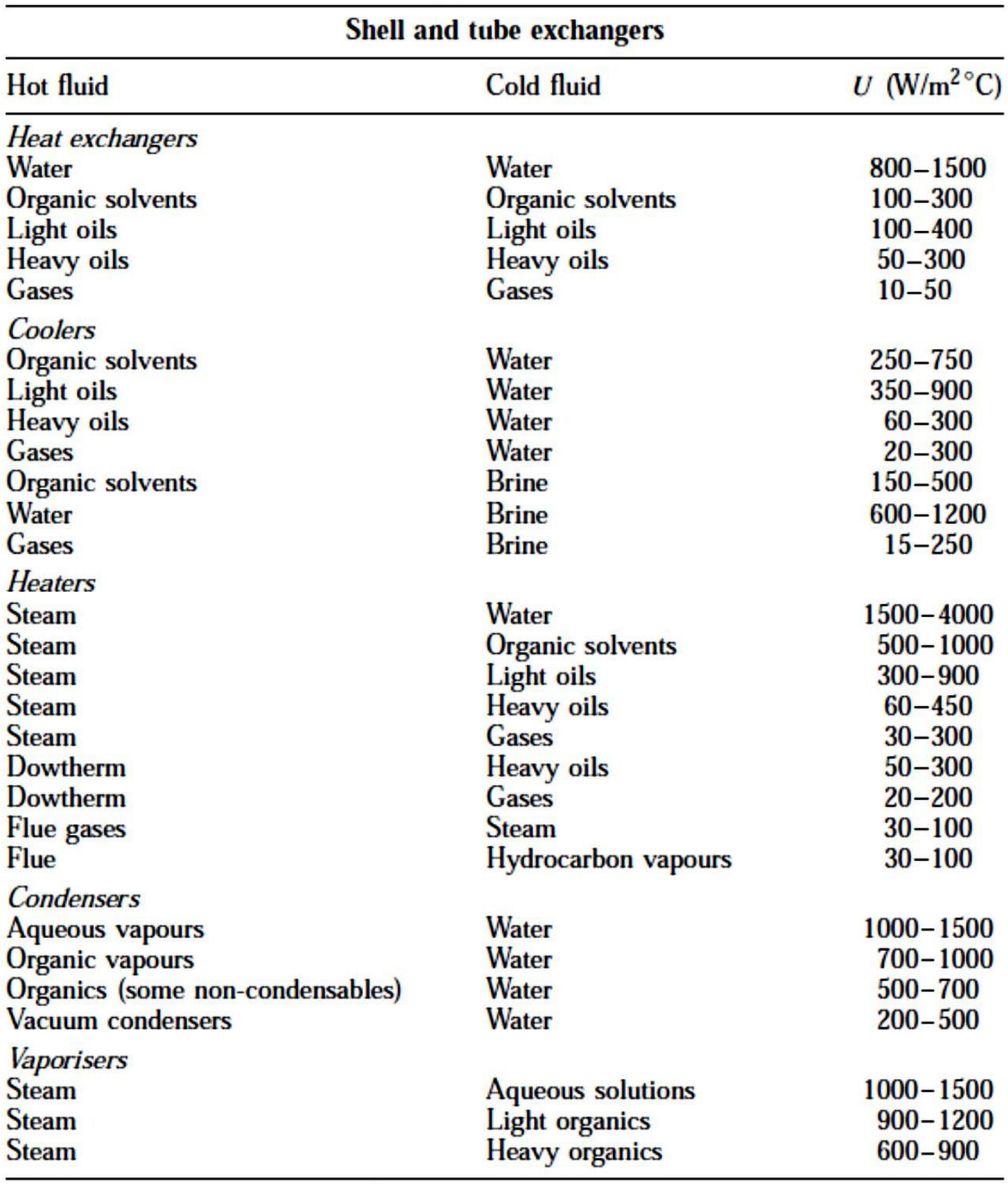
(a) One-shell pass and 2, 4, 6, etc. (any multiple of 2), tube passes

Figure 6:Ft factor for shell and tube

F=0.87

Δt_m =21.54 °C

6.2.3 Area Calculations



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

Figure 7: overall heat transfer coefficient

The coefficient can be estimated to be 600 W/m^2 C

Area can be calculated using:

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

A=46.425 m^2

So, the number of tubes can be calculated using:

Tube length=4.5m

Area of a single tube= πdl=0.359 m^2

Number of tubes= area/area of a single tube= 46.425/0.359=129 tubes

6.2.4 Shell Side Calculations

Shell area can be calculated using

$$As = \frac{(Pt - Po)Ds}{Pt}$$

Using this formula, the cross-sectional area of the shell is:

As=0.008m^2

The mass velocity thus comes out to be.

G=$\frac{\omega}{As}$= 1.24x10^7 kg/hr m^2

So, the Reynold's number turns out.

Re= 27582

$(Pr)^{1/3}$=1.773

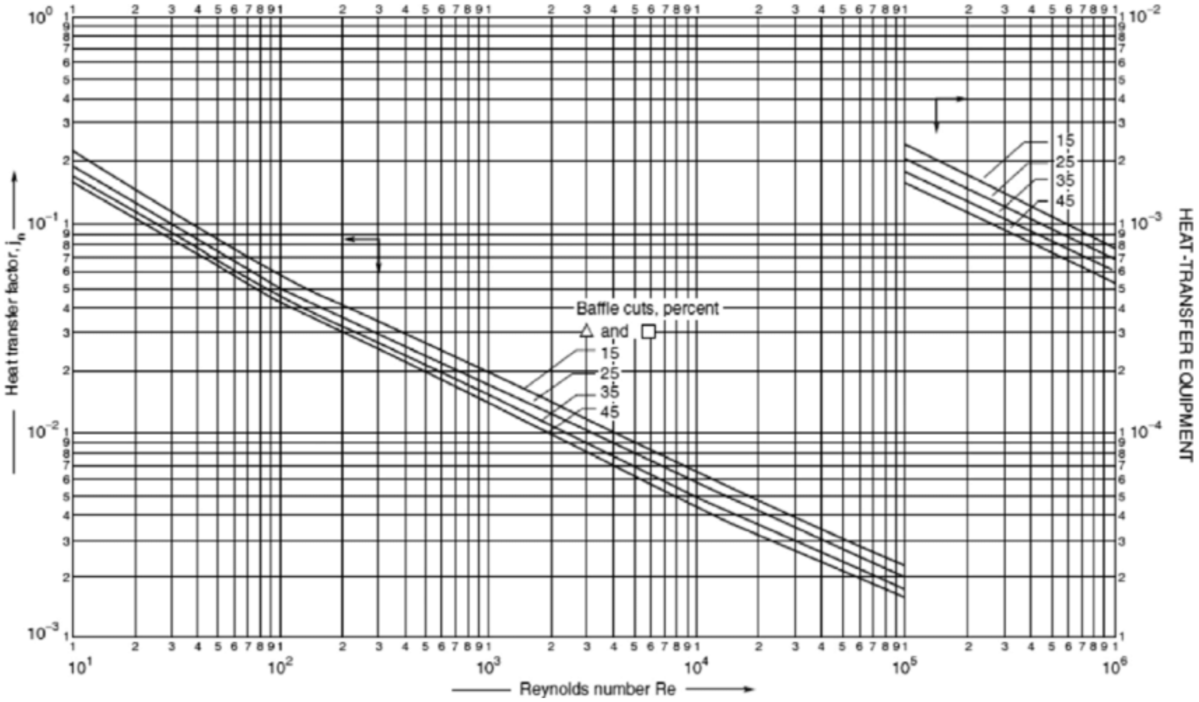


Figure 8: Heat transfer coefficient

From the graph jh=0.04

With the assumption of no variance in viscosity between the wall and the bulk flow, ho can be

calculated as:

$$ho = \frac{Re(Pr)^{1/3}k}{De}$$

De= 0.019m

ho= 412.7W/m^2K

6.2.5 Tube Side Calculations

Cross sectional area of tube= 0.034m^2

This are is considered for all the tubes combined.

Reynold's number=4554.8

$(Pr)^{1/3}$=1.34

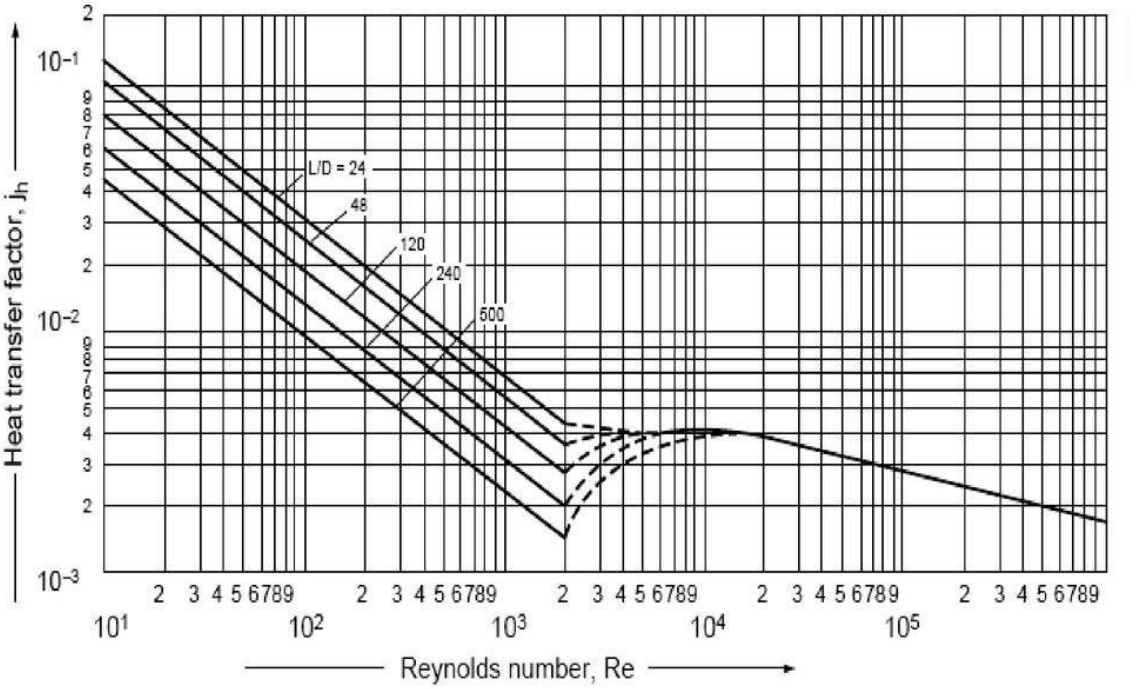


Figure 9: heat transfer factor graph

Using this graph, jh value is 0.05

hi calculations

$$hi = \frac{Re(Pr)^{1/3}k}{De} = 863.4W/m^2K$$

hio can be calculated using:

$hi = hio\frac{ID}{OD}$=$751.2W/m^2K$

6.2.6 Pressure drop Calculations

Pressure drop in shell:

$$\Delta Ps = \frac{fGs^2D_s(N+1)}{5.22\times10^{10}D_e s\,\varphi_s}$$

$$\Delta Ps = 4.76psi$$

Pressure drop in tubes:

$$\Delta P_t = \frac{f G t^2 L n}{5.22\times10^{10} D_s \varphi_t}$$

$$\Delta P_t = 9.2\text{psi}$$

6.2.7 Final Equipment specifications

Table 16: Shell and tube specs

Exchanger Data			
Shell Side		Tube Side	
ID (inch)	12	Number & Length (ft)	126, 15
Baffle Space (inch)	4	OD (inch), BWG, & Pitch (inch)	1, 16, & 1.25 (triangle)
Passes	1	Passes	2

6.3 Plate and Frame Heat Exchanger

Table 17: Plate and frame heat exchanger conditions

Hot Fluid		**Cold Fluid**
Inlet Temp(T_1) ^{0}C	120	-64.47
Exit Temp(T_2) ^{0}C	64.8	20
Mass Flow Rate(W) (kg/hr)	1.67 x 10^4	4791
C_p (kJ/kg.^{0}C)	2.491	2.295
Specific gravity	0.621	0.09
Thermal Conductivity (W/m.K)	0.9346	0.1027
Viscosity (cP)	0.25	0.06

6.3.1 Heat Duty

$Q = mc_p\Delta T$

Q=8.846 x 10^5 KJ/Kg

6.3.2 Operating Temperature

$$\mathrm{LMTD} = \frac{\Delta t_2 - \Delta t_1}{2.3\log\left(\frac{\Delta t_2}{\Delta t_1}\right)}$$

Counter current flow is used.

$$\mathbf{LMTD} = \frac{144.2 - 100}{2.3\log\left(\frac{144.2}{100}\right)} = 120.78°\mathrm{C}$$

$$\boldsymbol{NTU = 0.829}$$

For 1:1 pass,

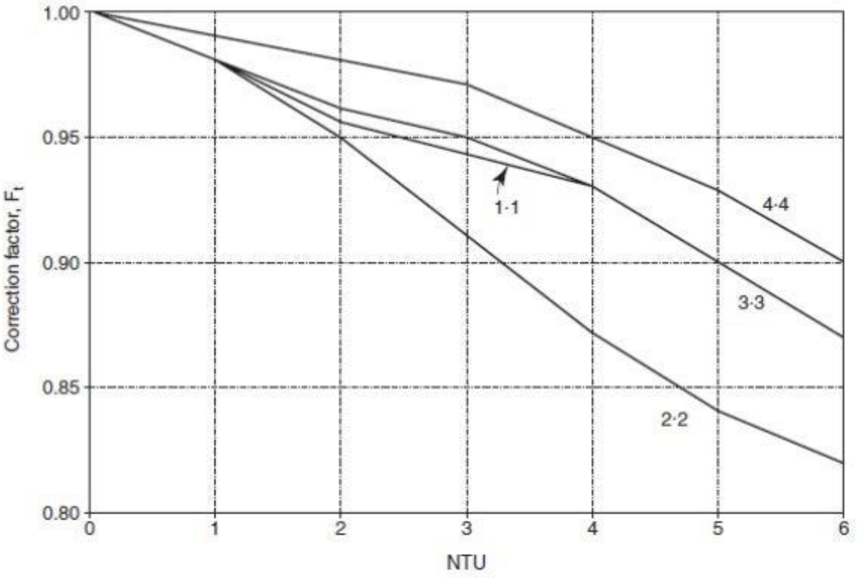


Figure 10: Ft for plate and frame

Δt_m =118.36 °C

6.3.3 Area Calculations

Assume U to be 700 W/m^2 C

Area can be calculated using:

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

A=9.8 m^2

Plate dimensions:

Length=1.5m

Width=0.5m

Area correction=1

Effective plate area= L x W= 0.75 m^2

Number of plates required can be calculated using= total area/ area per plate= 14 plates

So, the number of channels is 7

6.3.4 Hot fluid Side Calculations

Take the spacing of plates to be 3mm

Ac= 0.0015 m^2

De=2 x plate spacing= 6mm

Calculate the velocity requirement using:

$$channel\ velocity\ ut = \frac{\omega}{\rho \times Ac \times Nc} = 0.7 m/s$$

Then calculate Re, Pr and Nu using the relations below.

$$Re = \frac{ut \times \rho \times De}{\mu} = 10593$$

$$Pr = \frac{Cp \times \mu}{k} = 0.66$$

$$Nu = 0.26 Re^{0.65} Pr^{0.4} = 91$$

Use the Nusselt's number to find the value of hp

$$hp = \frac{Nu \times k}{De} = 14174 W/m^2 C$$

6.3.5 Cold Fluid Side Calculation

Calculate the velocity requirement using:

$$channel\ velocity\ ut = \frac{\omega}{\rho \times Ac \times Nc} = 1.43 m/s$$

Then calculate Re, Pr and Nu using the relations below.

$$Re = \frac{ut \times \rho \times De}{\mu} = 12641$$

$$Pr = \frac{Cp \times \mu}{k} = 1.39$$

$$Nu = 0.26Re^{0.65}Pr^{0.4} = 134.65$$

Use the Nusselt's number to find the value of hp

$$hp = \frac{Nu \times k}{De} = 2304.76W/m^2C$$

U can now be calculated using:

$$U = \frac{1}{\frac{1}{h_{ph}} + \frac{1}{h_{pc}} + \frac{t_p}{k_p} + Rc + Rh}$$

Assuming Rc and Rh to be 0.0003 each.

The thickness of plate is taken as 1.5mm.

The material used is stainless steel with k value of 15W/m^{2}C

U=830 W/m^{2}C

This shows a 18% overdesign. Given the large temperature difference, the difference is acceptable.

6.3.6 Pressure drop Calculations

Pressure drop in hot side:

$$\Delta Pp = \frac{8j_f(\frac{Lp}{De})\rho ut^2}{2}$$

$$jp = 0.6Re^{-0.3} = 0.037$$

$$\Delta Pp = 81.6 \text{ KPa}$$

Pressure drop in tubes:

$$\Delta Pp = \frac{8j_f(\frac{Lp}{De})\rho ut^2}{2}$$

$$jp = 0.6Re^{-0.3} = 0.035$$

$$\Delta Pp = 29.3 \text{ KPa}$$

6.4 Design of Desalter and Separators

Design of separators is subject to a variety of constraints with optimization carried out to get the most effective equipment.

6.4.1 Desalter

The desalter is taken a horizontal vessel and thus the design is subject to the constraints below.

$$d^2 Leff = 420\left[\frac{TZQg}{P}\right][\left(\frac{\rho g}{\rho l - \rho g}\right)\frac{Cd}{dm}]^{0.5}$$

This is based on droplet settling theory using the ballistic model. The constraint is for liquid drops in gas phase.

$$d^2 Leff = 1.42[Q_w t_{rw} + Q_o t_{ro}]$$

The equation above represents retention time constraint which plays a far more significant role in our separation.

Settling velocity is given by: u=0.07$\sqrt{\frac{\rho_l - \rho_v}{\rho_l}}$

A few assumptions were made in order to solve the above relations:

Vessel half filled with liquid

t_0 *= 1 minutes*

t_w *= 1 minutes*

Given

T = 536.2 R

Z = 0.8265

P = 906.5 psia

Qg = 20.15 MMsfd

$\rho \boldsymbol{l}$ = 39.19 lb/ft^3

$\rho \boldsymbol{v}$ = 3.657 lb/ft^3

μ = 0.2945 cP

ΔSg = 1 - 0.628 = 0.372

Qo = 4303 B/D

Qw = 58.67 B/D

Above are all of the required parameters for calculation of the equations. Plugging them in results in:

$\boldsymbol{d^2 Leff = 132.74}$ **Gas constraint**

$\boldsymbol{d^2 Leff = 13053.634}$ **Retention time constraint**

The thickness of the vessel is given by a simple pressure vessel formula.

$$e = \frac{Pi\ Di}{2f - Pi}$$

For, f=135 and Di=1.0668

e= 1.1 inch

The sizing which satisfies the equipment is given below.

EQUIPMENT SPECIFICATIONS	
Diameter	42 in
Effective length	11.25 ft
Seam to seam length	15 ft
Thickness	in

Table 18: Desalter equipment specification

6.4.2 Vertical Separators

The separators are vertical vessel and thus the design is subject to the constraints below.

$$d^2 = 5054 \left[\frac{TZQg}{P}\right] [\left(\frac{\rho g}{\rho l - \rho g}\right) \frac{Cd}{dm}]^{0.5}$$

This is based on droplet settling theory using the ballistic model. The constraint is for liquid drops in gas phase.

$$d^2(ho + hl) = \frac{1}{0.12} [Q_w t_{rw} + Q_o t_{ro}]$$

An additional Demister sizing constraint must also be added where kd is the demister coefficient.

$$d^2 = \frac{60\frac{TZQg}{P}}{kd\left(\frac{\rho g}{\rho l - \rho g}\right)^{0.5}}$$

The equation above represents retention time constraint which plays a far more significant role in our separation.

Settling velocity is given by: u=0.07$\sqrt{\frac{\rho_l-\rho_v}{\rho_l}}$

A few assumptions were made in order to solve the above relations:

Vessel half filled with liquid

t_0 = 1 minutes

t_w = 1 minutes

The data provided is depicted in the table below:

Table 19: Separator conditions

	Separator 1	Separator 2	Separator 3
T (R)	513.34	412.47	376.74
Z	0.798	0.5929	0.7045
Qg (MMscf)	20.12	17.53	16.72
P(psia)	901.5	894.3	536.6
Density gas(lb/ft^3)	3.426	6.11	3.257
Density liquid(lb/ft^3)	37	25.7	29.29

Carrying out the calculations, the constraints are organized in the table below.

Table 20: Constraints to design of separators

	Separator 1	Separator 2	Separator 3
Gas constraint	d=28 inch	d=26.4 inch	d=28 inch
Mesh sizing constraint	d=13.6 inch	d=35 inch	d=30 inch
Retention time	d^2hl=132	d^2hl=380333	d^2hl=499000

Through hit and trial, designing a vessel that supports all of the above constraints gives us resultant vessels in the table.

The thickness of the vessel is given by a simple pressure vessel formula.

$$e = \frac{Pi\ Di}{2f - Pi}$$

For, f=135 and various Di

Table 21: Equipment specification of separators 1-3

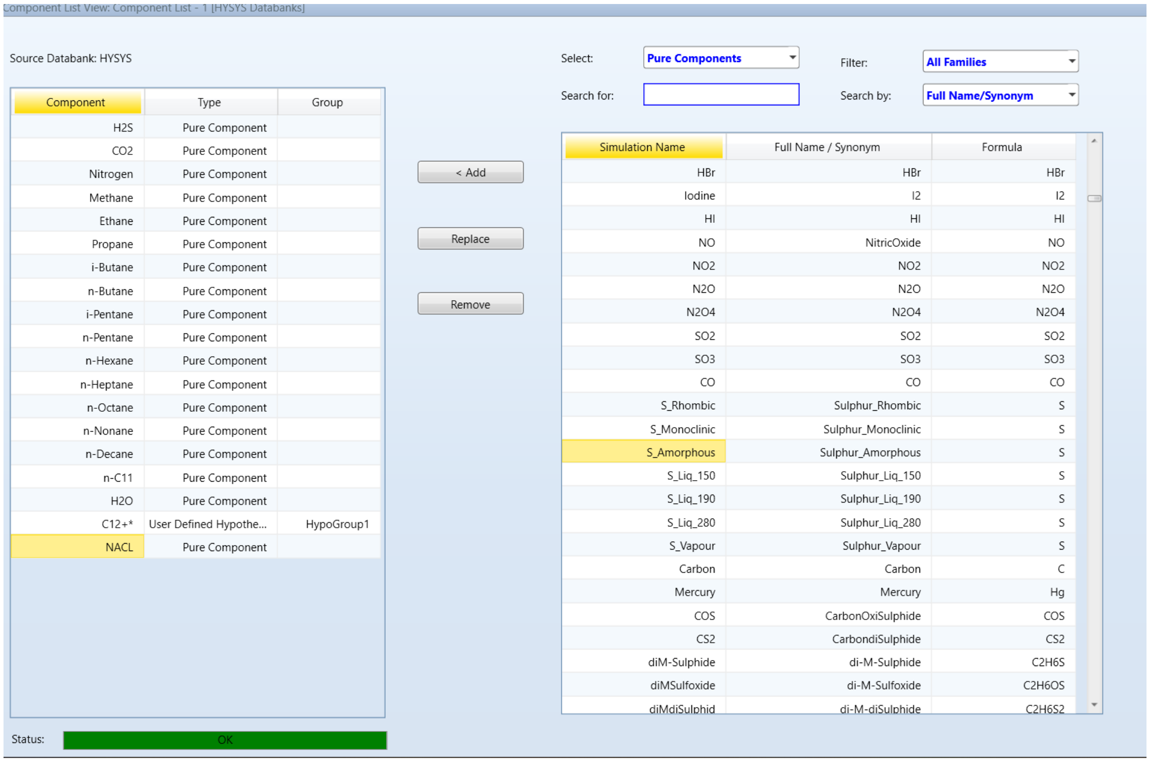
EQUIPMENT SPECIFICATIONS			
Separator	1	2	3
Diameter	28 in	35 in	30 in
Effective length	7 ft	8.5 ft	7.89 ft
Seam to seam length	9.5 ft	11.5 ft	15 ft
Thickness	1 in	1 in	1 in

CHAPTER 7

SIMULATION

7.1 Selection of Components

The components were selected from the HYSYS library. The C12+ component has been added manually. NaCl was used to simulate salts in the crude oil.



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

Source Databank: HYSYS

Select: Pure Components

Filter: All Families

Search for:

Search by: Full Name/Synonym

Component	Type	Group
H2S	Pure Component	
CO2	Pure Component	
Nitrogen	Pure Component	
Methane	Pure Component	
Ethane	Pure Component	
Propane	Pure Component	
i-Butane	Pure Component	
n-Butane	Pure Component	
i-Pentane	Pure Component	
n-Pentane	Pure Component	
n-Hexane	Pure Component	
n-Heptane	Pure Component	
n-Octane	Pure Component	
n-Nonane	Pure Component	
n-Decane	Pure Component	
n-C11	Pure Component	
H2O	Pure Component	
C12+*	User Defined Hypothe...	HypoGroup1
NACL	Pure Component	

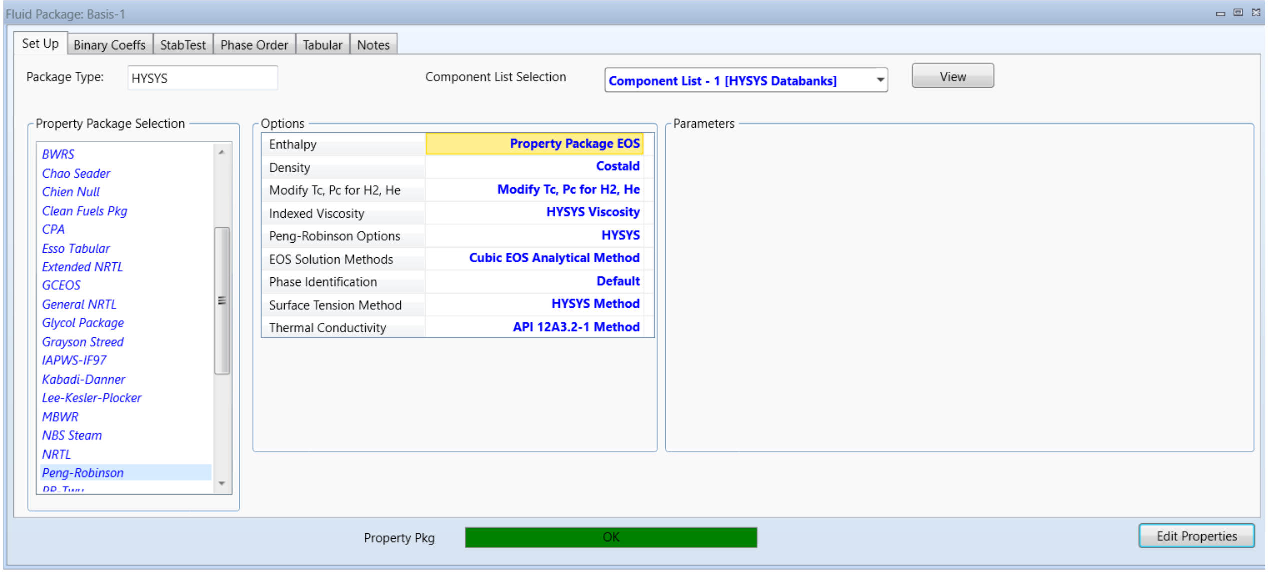
Simulation Name	Full Name / Synonym	Formula
HBr	HBr	HBr
Iodine	I2	I2
HI	HI	HI
NO	NitricOxide	NO
NO2	NO2	NO2
N2O	N2O	N2O
N2O4	N2O4	N2O4
SO2	SO2	SO2
SO3	SO3	SO3
CO	CO	CO
S_Rhombic	Sulphur_Rhombic	S
S_Monoclinic	Sulphur_Monoclinic	S
S_Amorphous	Sulphur_Amorphous	S
S_Liq_150	Sulphur_Liq_150	S
S_Liq_190	Sulphur_Liq_190	S
S_Liq_280	Sulphur_Liq_280	S
S_Vapour	Sulphur_Vapour	S
Carbon	Carbon	C
Mercury	Mercury	Hg
COS	CarbonOxiSulphide	COS
CS2	CarbondiSulphide	CS2
diM-Sulphide	di-M-Sulphide	C2H6S
diMSulfoxide	di-M-Sulfoxide	C2H6OS
diMdiSulphid	di-M-diSulphide	C2H6S2

Status: OK

Figure 11: Simulation of component selection

7.2 Selection of Fluid Package

The fluid package was selected to be Peng-Robinson on the basis of compatibility with the selected components. The HYSYS method assistant was also used, and the final selection was made.



Package Type: HYSYS

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

Property Package Selection:
BWRS
Chao Seader
Chien Null
Clean Fuels Pkg
CPA
Esso Tabular
Extended NRTL
GCEOS
General NRTL
Glycol Package
Grayson Streed
IAPWS-IF97
Kabadi-Danner
Lee-Kesler-Plocker
MBWR
NBS Steam
NRTL
Peng-Robinson

Options

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

Property Pkg: OK

Figure 12: Simulation of fluid package selection

7.3 Process Flowsheet Modelling and Simulation

The process flow diagram was replicated as the flowsheet on Aspen HYSYS, as shown bythe snippet. The molar flows were the same as that of our manual calculations. The temperature and pressure parameters were also defined according to the requirements. The final separation of the distillation column was checked as well as the U calculations for the heat exchangers.

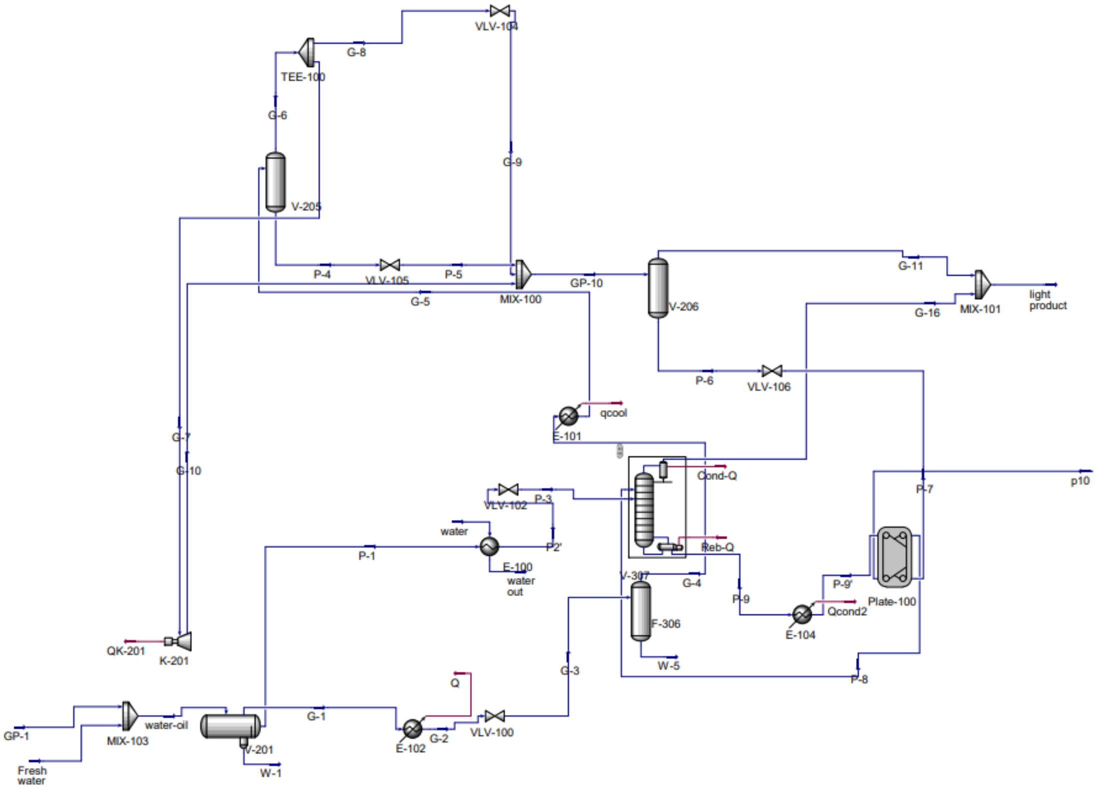
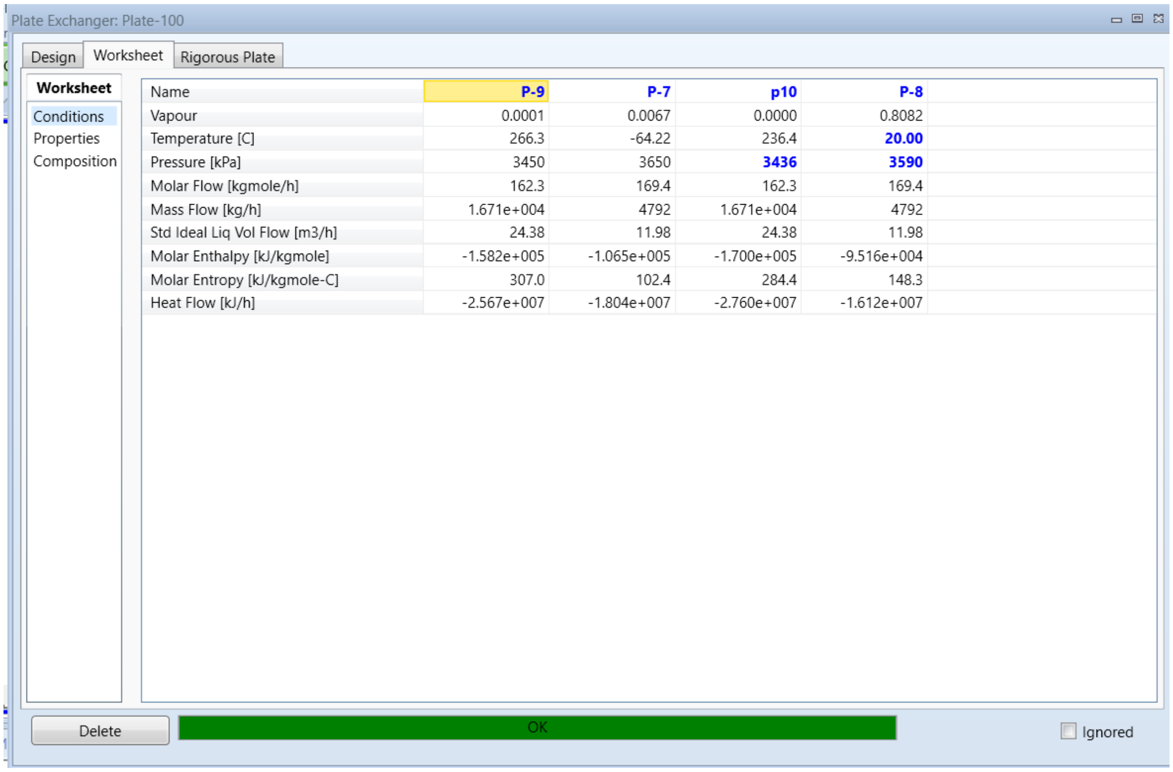


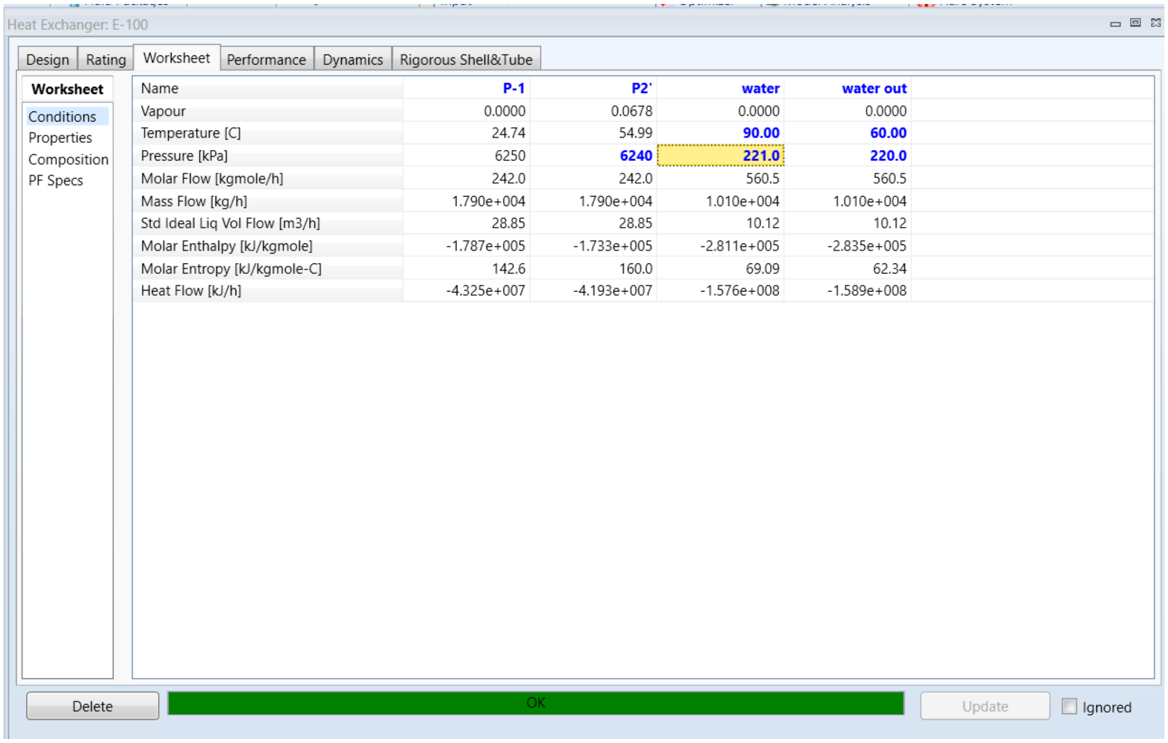
Figure 13: Simulation of flowsheet

7.4 Heat Exchanger worksheet



Name	P-9	P-7	p10	P-8
Vapour	0.0001	0.0067	0.0000	0.8082
Temperature [C]	266.3	-64.22	236.4	20.00
Pressure [kPa]	3450	3650	3436	3590
Molar Flow [kgmole/h]	162.3	169.4	162.3	169.4
Mass Flow [kg/h]	1.671e+004	4792	1.671e+004	4792
Std Ideal Liq Vol Flow [m3/h]	24.38	11.98	24.38	11.98
Molar Enthalpy [kJ/kgmole]	-1.582e+005	-1.065e+005	-1.700e+005	-9.516e+004
Molar Entropy [kJ/kgmole-C]	307.0	102.4	284.4	148.3
Heat Flow [kJ/h]	-2.567e+007	-1.804e+007	-2.760e+007	-1.612e+007

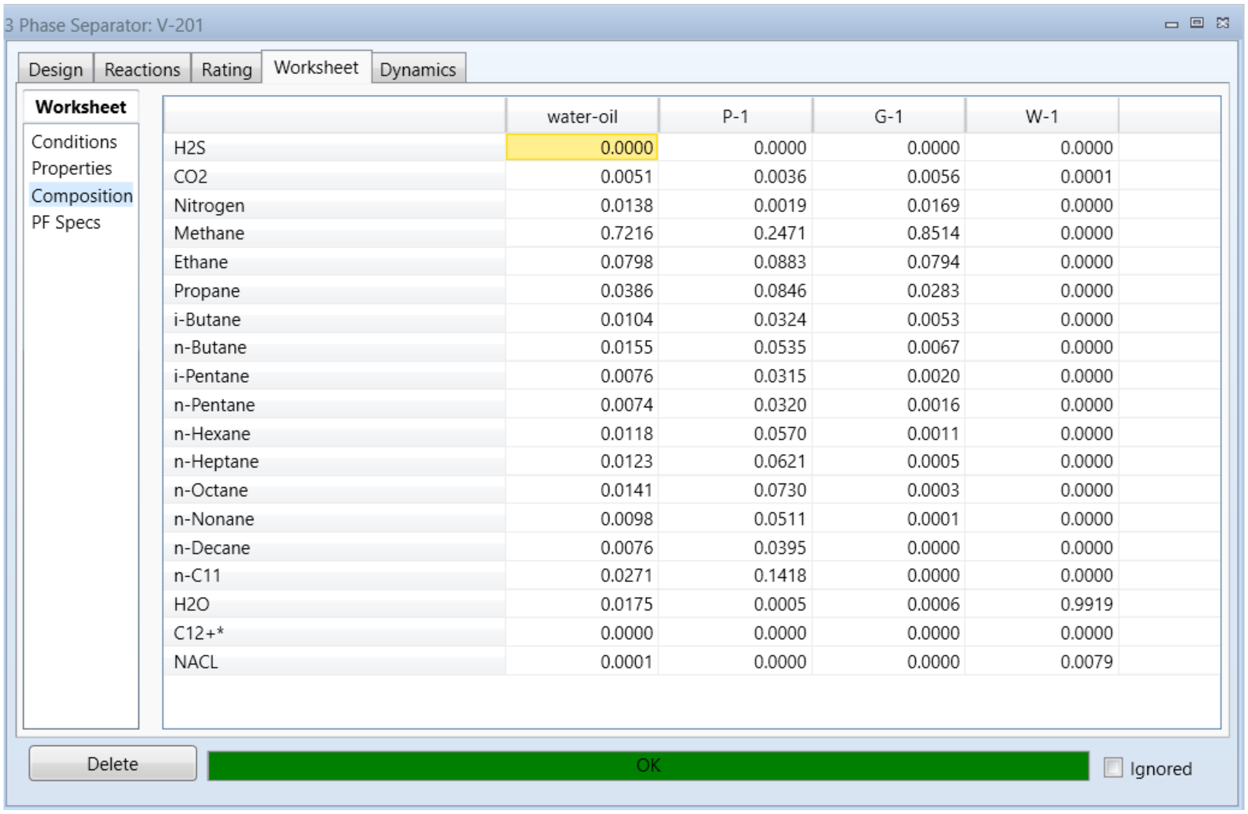
Figure 14:Simulation of Shell and tube



Name	P-1	P2'	water	water out
Vapour	0.0000	0.0678	0.0000	0.0000
Temperature [C]	24.74	54.99	90.00	60.00
Pressure [kPa]	6250	6240	221.0	220.0
Molar Flow [kgmole/h]	242.0	242.0	560.5	560.5
Mass Flow [kg/h]	1.790e+004	1.790e+004	1.010e+004	1.010e+004
Std Ideal Liq Vol Flow [m3/h]	28.85	28.85	10.12	10.12
Molar Enthalpy [kJ/kgmole]	-1.787e+005	-1.733e+005	-2.811e+005	-2.835e+005
Molar Entropy [kJ/kgmole-C]	142.6	160.0	69.09	62.34
Heat Flow [kJ/h]	-4.325e+007	-4.193e+007	-1.576e+008	-1.589e+008

Figure 15:Simulation of Plate and frame

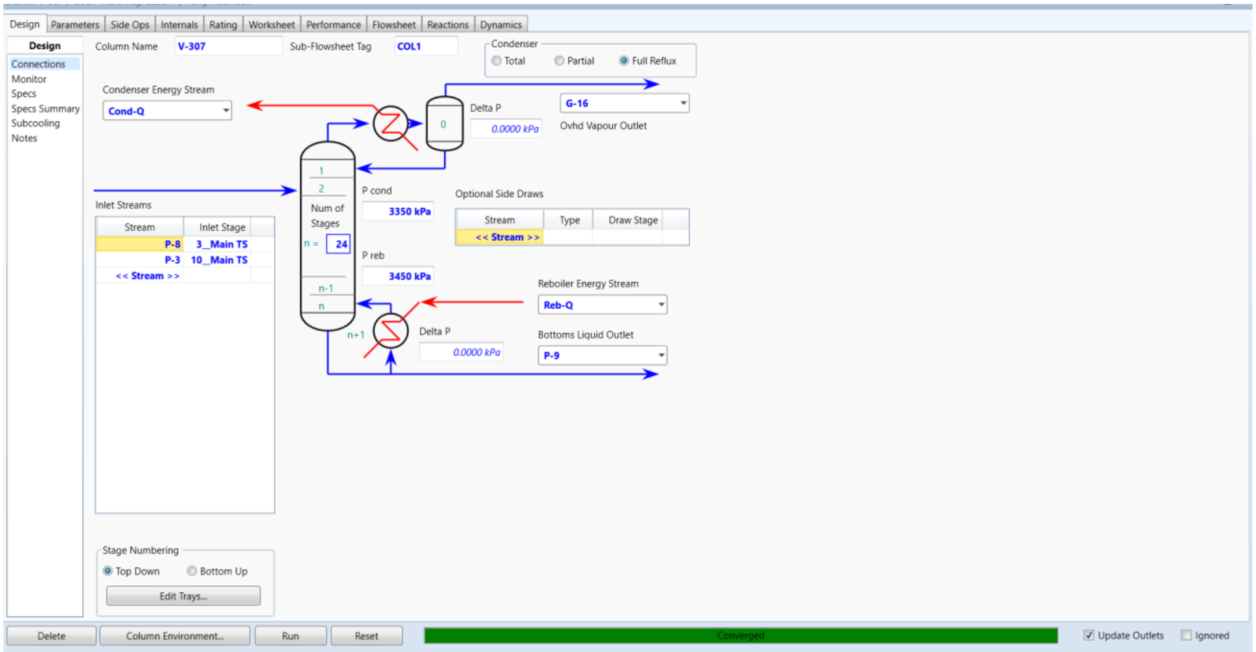
7.5 Three Phase Separator Worksheet



	water-oil	P-1	G-1	W-1
H2S	0.0000	0.0000	0.0000	0.0000
CO2	0.0051	0.0036	0.0056	0.0001
Nitrogen	0.0138	0.0019	0.0169	0.0000
Methane	0.7216	0.2471	0.8514	0.0000
Ethane	0.0798	0.0883	0.0794	0.0000
Propane	0.0386	0.0846	0.0283	0.0000
i-Butane	0.0104	0.0324	0.0053	0.0000
n-Butane	0.0155	0.0535	0.0067	0.0000
i-Pentane	0.0076	0.0315	0.0020	0.0000
n-Pentane	0.0074	0.0320	0.0016	0.0000
n-Hexane	0.0118	0.0570	0.0011	0.0000
n-Heptane	0.0123	0.0621	0.0005	0.0000
n-Octane	0.0141	0.0730	0.0003	0.0000
n-Nonane	0.0098	0.0511	0.0001	0.0000
n-Decane	0.0076	0.0395	0.0000	0.0000
n-C11	0.0271	0.1418	0.0000	0.0000
H2O	0.0175	0.0005	0.0006	0.9919
C12+*	0.0000	0.0000	0.0000	0.0000
NACL	0.0001	0.0000	0.0000	0.0079

Figure 16: Simulation of a three-phase separator

7.6 Distillation Column worksheet



Column Name: V-307

Sub-Flowsheet Tag: COL1

Condenser: ○ Total ○ Partial ● Full Reflux

Condenser Energy Stream: Cond-Q

Delta P: 0.0000 kPa

Ovhd Vapour Outlet: G-16

P cond: 3350 kPa

Num of Stages: n = 24

P reb: 3450 kPa

Reboiler Energy Stream: Reb-Q

Delta P: 0.0000 kPa

Bottoms Liquid Outlet: P-9

Inlet Streams

Stream	Inlet Stage
P-8	3__Main TS
P-3	10__Main TS
<< Stream >>	

Optional Side Draws

Stream	Type	Draw Stage
<< Stream >>		

Stage Numbering: ● Top Down ○ Bottom Up

Converged

☑ Update Outlets ☐ Ignored

Figure 17: Simulation of Distillation column

CHAPTER 8

ECONOMIC ANALYSIS

8.1 3-Phase Separator

The cost for the vertical separator made up of stainless steel with a height of 4.5 m is calculated by using Coulson and Richardson volume 6 and is then adjusted for inflation in 2022.

6000$x2x2.2=26400 $In 2022 = 43000 $

8.2 Separators

There are three separators installed, and each are calculated with carbon steel as the material. Cost for splitter is calculated by the graph in Coulson and Richardson volume 6and then adjusted for inflation in 2022.

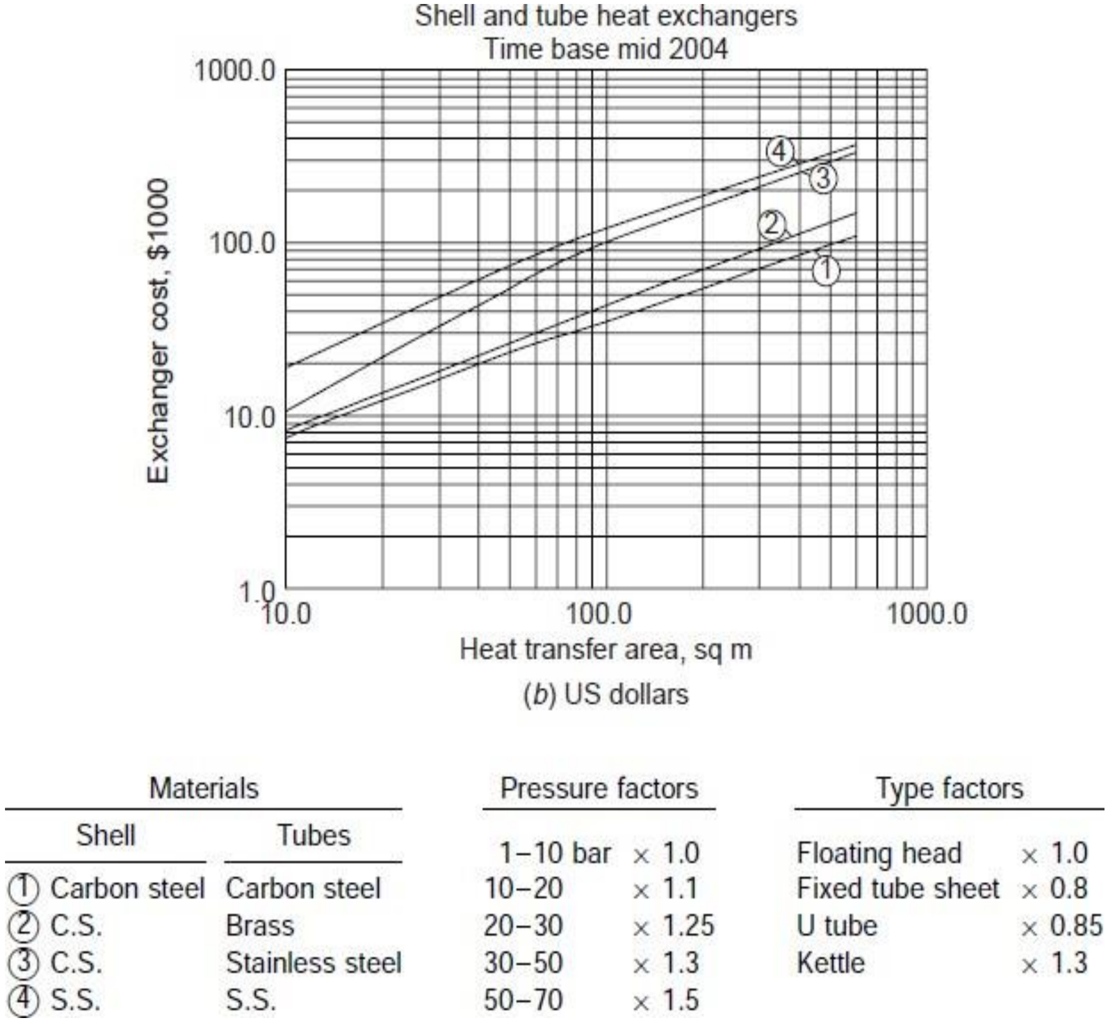
1. 6500$ x 1 x 2.2 x index2022=23700$
2. 8000$ x 1 x 2.2 x index2022=29200$
3. 7000$ x 1 x 2.2 x index2022=25500$

8.3 Coolers & Expander

The coolers employed excluding the cool box can be estimated using sources online. This price is roughly estimated at 11,400$ with the help of hysys. Similarly, the cost of the expander can be estimated at 4000$.

8.4 Shell and Tube heat exchanger

The heat exchanger is sized so the price can be estimated according to the area. The cost for the vessel made up of stainless steel is calculated byusing Coulson and Richardson volume 6 and is then adjusted for inflation in 2022.



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

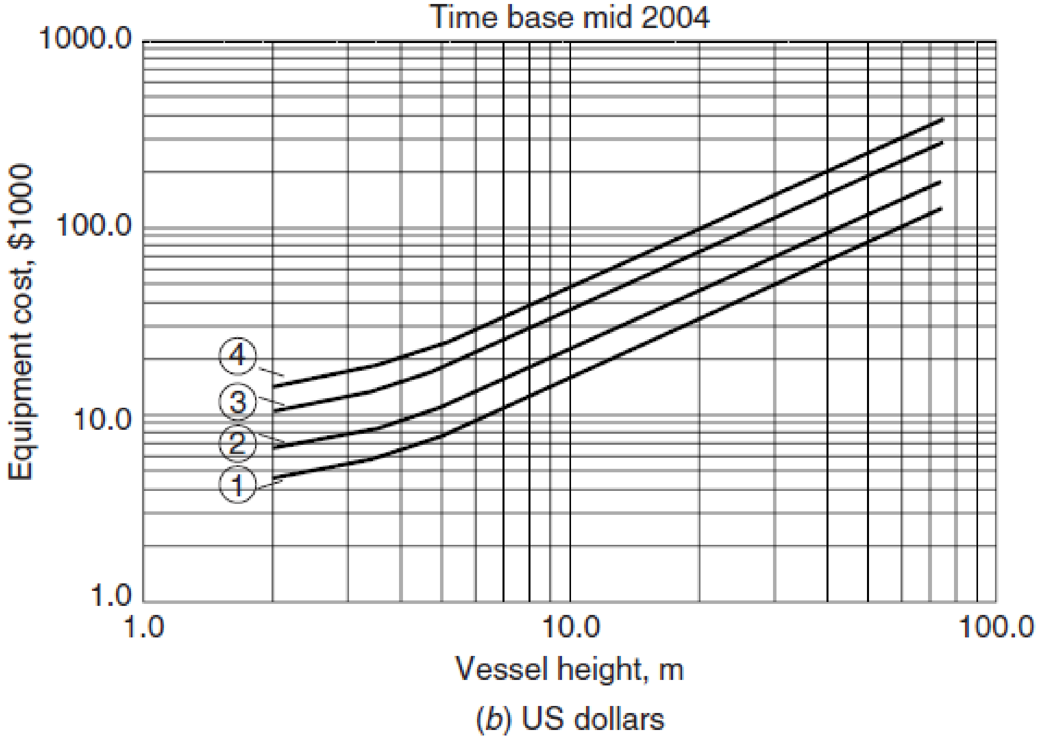
Figure 18: Design cost for Shell and tube heat exchanger

11000$x1x1.5=16500 $

In 2022 = 26900$

8.5 Crude stabilizer column

The distillation column used is 15.7 m in height and 1.6 m in diameter. It is fitted with 24 plates of the sieve type and operated at a high pressure. The Costing has been taken from the graph below and updated using the inflation index.



Diameter, m	
①	0.5
②	1.0
③	2.0
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 19:Graph for column cost

Column cost=40000$ x 1 x 2.2= 88000$

Plate costing= 24 x 200= 4800$

Reboiler costing= 97000$

Condenser costing= 49000$

Total column cost in 2022 =291000$

8.6 Plate and Frame heat exchanger

The heat exchangers designed is small in area however due to high temperature and pressure, the price of the heat exchanger is elevated. Using hysys, the price of the plate and frame heat exchanger is estimated to be 18000$.

8.7 Plant equipment total

Table 22: Cost of major equipment

Equipment	Cost/USD
3 phase separator/desalter	43000
Separator 1	23700
Separator 2	29200
Separator 3	25500
Shell and Tube heat exchanger	26900
Plate and Frame heat exchanger	18000
Cool box	56600
Distillation column	291000
Subtotal	513900

8.8 Plant Cost Calculations

Table 23: PPC

Total purchase cost	**factor**
Equipment erection	0.4
Piping	0.7
Instrumentation	0.2
Electrical	0.1
Buildings	0.15
Utilities	0.5
Storage	0.15
Site development	0.05
Ancillary buildings	0.15

Physical Plant Cost= 3.4 x Equipment cost = 1747600$

Table 24: PCE

Expenses	**factor**
Design and engineering	0.3
Contractor’s fee	0.05
Contingency	0.1

Fixed Capital Cost= PPC x 1.45=2534000$

Variable cost for operation of the plant annually:

Table 25: Variable cost

Expenses	**Cost/USD**
Raw materials	850000
Miscellaneous materials	12800
Shipping and packaging	negligible

Fixed cost for operating the plant annually

Table 26: Fixed cost

Expenses	**Cost/USD**
Maintenance	128000
Operating labor	540000
Lab costs	108000
Supervision	108000
Plant overheads	270000
Capital charges	253300
Insurance	25330
Local taxes	50660
Royalties	25330

Variable cost subtotal: 862800 USD

Fixed cost= 1508620 USD

Additionally, other costs such as sales, R&D and general overhead are also added:
Additional cost= 593000 USD

8.9 Summary of Economic Analysis

Table 27;summary of the costs

Summary of costs in USD	
Fixed Capital	2534000
Variable operating cost	862800
Fixed operating cost	1508620
Annual operating cost	2964300
Annual revenue	3761000
Annual profit	796700
ROR	0.211

The change delivered by installation of the desalter is quantified by a 15% reduction in maintenance and miscellaneous cost:

Payback= investment (purchase of equipment)/ reduction in cost=

43000(1.4)/0.15 x 140800=

2.85 years

CHAPTER 9

INSTRUMENTATION

Instrumentation and Process Control are critical to the efficient operation of a process plant. Process control is important for multiple reasons such as equipment safety or product specification, but generally the purpose of equipment control is to keep the operation within set boundaries.

The main mode of process control is through feedback process loops. These loops have sensors that control individual aspects of the process. The sensors measure what is known as a process variable and forwards it to the controller. The controller subtracts the process variable from the set point in order to get the value of the error. Following this, the error is amplified using a certain amount of gain in the controller which could be of three types: Integral, proportional or differential. The signal is then sent towards an actuator which tweaks the conditions of the process variable. This may be applied to many parameters and help in having a certain level of safety within the process.

9.1 Flow Control Loop

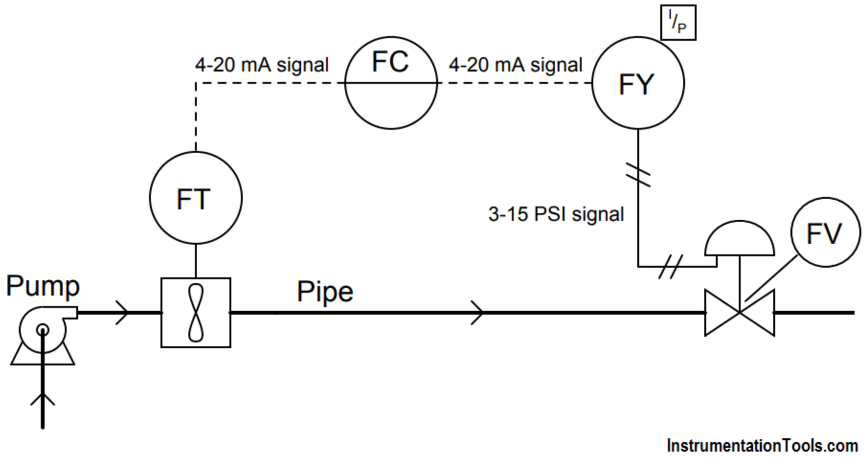


Figure 20: Flow control loop

In the loop design depicted above, a flow transmitter is used in order to capture the current value of the process variable. This is then forwarded to the flow controller which calculates the error and uses the valve opening as a control variable. Such a set up could provide a beneficial effect through achieving pulsating flow which would increase the shear stress for the walls of the shell and tube heat exchanger and allow less fouling to take place.

9.2 Level Control Loop

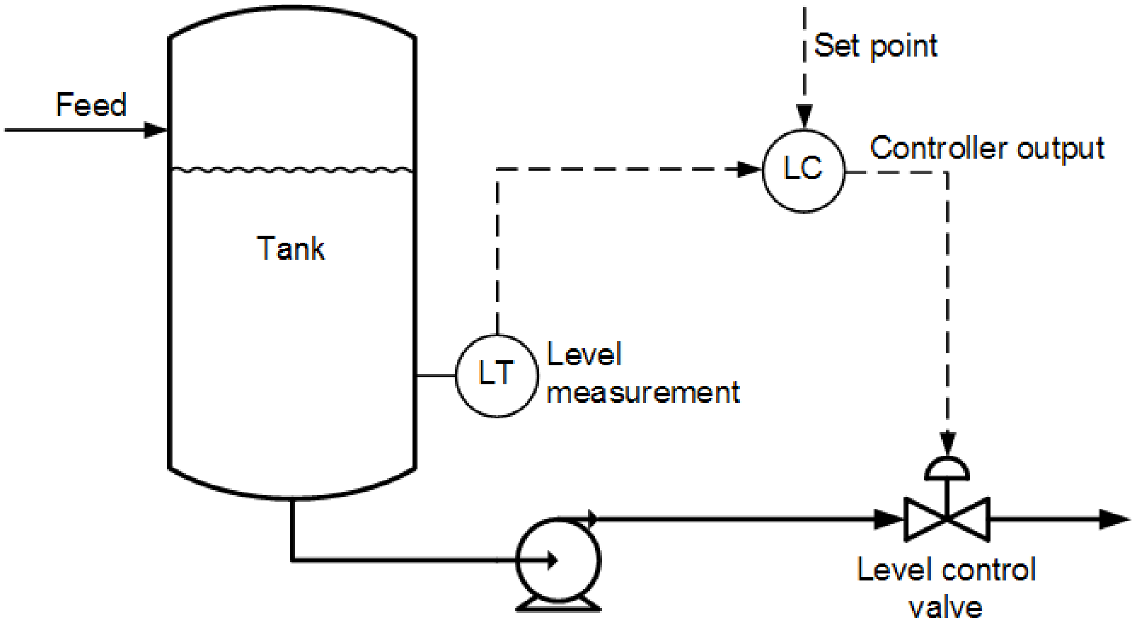


Figure 21: Level control loop

In this set up there is a level transmitter which allows for the measurement of the liquid or even vapor level within a tank or a vessel. The recorded value is noted and compared against the level set point in the controller. Level sensors generally work on the principle of buoyancy but newer technology has also been utilized. Finally, the controller sends out a signal to either the drain valve or the feed valve in order to ensure a proper level of fluid in the vessel.

9.3 Temperature Control Loop

Temperature control loops are one of the most significant control loops available.

This control loop can be actuated through many pathways including valves which allow a greater fraction of heated feed in, and heaters and coolers which simply increase the energy transferred to the stream. The working principle of these loops is the same.

9.4 Pressure Control Loop

Pressure control loops are extremely important for plant safety. These loops allow for pressure to be tracked and controlled in the case of fluctuations in the feed. Highly pressurized vessels have the capacity to burst which could cause injuries and even death to the field workers. Actuators for this loop may include compressors and pressure release valves allowing for deadly accidents to be mitigated.

CHAPTER 10

HAZOP ANALYSIS

HAZOP which stands for Hazard and Operability study of the proposed is a structured method of analyzing the system for any hazards that may arise. It is one of the fundamental techniques used in chemical plants in order to reduce the risk of accidents.

This sort of study is usually done using one of two methods: qualitative or quantitative risk analysis. Most companies use the quantitative method to analyze for a better and more detailed examination of any scenario, whereas qualitative is essentially a prerequisite form of study. Because this is a hazardous study, both of these techniques of analysis demand various methodologies and particular sorts of individuals who are fit for this profession to accomplish such duties. There are several guiding words for the many types of deviations that might occur, as well as what causes them, the immediate repercussions of the deviation, and what action to take if one occurs.

10.1 HAZOP on the Separators

Table 28: HAZOP analysis on Separators

Separators					
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.

10.2 HAZOP on the Distillation Column

Table 29: HAZOP analysis on Distillation column

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.

10.3 Hazop on Desalter

Table 30: HAZOP analysis on Desalter

Flow	NO	No incoming flow	Rupture of preceding pipeline.	Spillage of crude oil-desalting water mixture.	Strict monitoring of pipeline and pumps 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
	LESS	Level of liquid in vessel lower than normal.	Inlet Flow clogged. Mixer beforehand is clogged	Leads to inefficient working of the desalter. Complete dry out of the	Cleaning inlet nozzle. Monitoring the operation of the mixer.

				desalter is possible.	

CHAPTER 11

CONCLUSIONS

In conclusion, the issue of fouling has been mitigated successfully. The salt content of the feed stream has been reduced to untraceable amounts which reduces the likelihood of solid deposition within the heat exchanger, but also reduces the load of the distillation column as well as the rest of the process equipment. The installed desalter, was designed at minimum cost and will be able to produce tangible cost reductions in the process. Maintenance costs which are the largest for a heat exchanger have been reduced by roughly 15% which will be of far greater benefit in the future.

In the scenario of the utility fluid being the culprit of the fouling, an alternative method would be to switch the fluid to a hot oil such as Thermanole 66 or 55. The cost will sky-rocket but the life of the heat exchanger will be extended and possible energy conservation would also bring another perk.

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