Recovery of Ammonia and Hydrogen from Purge Gas of Ammonia Synthesis Loop of FFC

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Certificate

This is to certify that work in this thesis has been carried out by **Miss Zarqoon Mumtaz, Mr. Muhammad Ehtisham Khattak** and **Mr. Usman Ali Khan** and completed under my supervision in School of Chemical and Materials Engineering, National University of Sciences and Technology, H-12, Islamabad, Pakistan.

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Dedication

Our work is dedicated to SCME and our beloved parents.

Acknowledgement

All thanks to Allah Almighty for His countless blessings and rewards, Who gave us the strength and ability to complete this thesis.

We would like to thank our supervisor Dr. Wasif Farooq for his support and guidance during our project. He has been very generous and supportive and helped a lot to complete this thesis. We are thankful to other faculty of SCME as well especially Sir Muhammad Ahsan for his assistance.

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We are also thankful to our parents, without their support and encouragement, it would not have been possible to complete the project successfully.

Abstract

In the past few years, air pollution has become one of the major threats to the environment and it is harmful for every living being on the planet. The emission of harmful waste gases from the industries pose a major threat to life on the planet. The Purge and Off gas from the ammonia synthesis plant were also directly burnt off in the boilers which results in NOx formation. There was a need to develop an economical and environmentally friendly process to coup up with the increasing industrial challenges.

The project focuses on the recovery of Ammonia from the Purge Gas coming from the synthesis Loop and Off Gas from the Letdown Vessel of the Ammonia synthesis Plant-1 of FFC. The ammonia free Purge gas is further subjected to membrane separation for the recovery of Hydrogen Gas which is recycled back to the main synthesis loop as a raw material for Ammonia production, while the ammonia free Off Gas is sent to the Boilers for burning. Ammonia is finally recovered from the distillation tower and sent to storage. This project does not only provide an environmental friendly solution for the problem at hand, but also produces two very useful products, i.e. Ammonia and Hydrogen, which can further be used for various applications, thus making it an economical process.

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Nomenclature

C_p	Heat Capacity
ΔT_m	Log Mean Temperature Difference
m	Meter
mm	Millimeter
cm	Centimeter
m^2	Meter square
m^3	Meter cube
cm^2	Centimeter square
cm^3	Centimeter cube
KJ	Kilo Joule
kg	Kilogram
h	Hour
C_v	Heat Capacity of Vapor
C_l	Heat Capacity of Liquid
gmol	gram mole
L	Length
do	Outer diameter of tubes
di	Internal diameter of tubes
G	Mass Velocity
α	Selectivity
μ	Viscosity
ƛ	Latent heat
v	velocity
A	Area
y_1	Mole Fraction
y_2	Mole Fraction
T	Temperature

N	Number of tubes
ΔT	Temperature difference
m	mass flow rate
Q	Energy
ρ_f	fluid density
°C	Centigrade
°K	Kelvin
P	Pressure
ρ	Density
KJ/kg	Kilo Joule per kilogram
U	Overall Heat Transfer Coefficient
P_t	Tube Pitch
D_s	Shell Diameter
I_b	Baffle Spacing
G_s	Mean Velocity
F_t	Friction Factor
W	Watt
H_2	Hydrogen
N_2	Nitrogen
Re	Reynolds Number
ΔP	Pressure Difference
PCE	Purchase Cost of Equipment
h_i	Internal Pipe Heat Transfer Coefficient
O.D	Outer diameter
A_s	Cross Sectional Area
Pr	Prandtl Number
Nu	Nusselt Number
d_e	Equivalent diameter

Np	Number of Tube Passes
u_t	Tube Side Velocity
jh	Heat Transfer Factor
s	Second
do	Tube Outside Diameter
Dv	Diffusivity of vapor
Z	Height of Column
HRU	Hydrogen Recovery Unit
ARU	Ammonia Recovery Unit

Chapter 1: Introduction

1.1 Background

Modern Engineering and technological advancement have made processes very efficient. Processes with low yields are not preferred now a days and efforts are being made continuously to make a process more and more efficient. A convenient way for increasing efficiency of a process is to reduce waste emission and reuse the wastes still emitted. This not only increases the yield but also protects the environment from hazardous wastes of chemical processes.

In chemical plants, some of the wastes are inevitable. One such example is Ammonia production in Fertilizer industry. Purge-gas is extracted from the ammonia synthesis loop to maintain the inert level in the loop and maximize conversion in the ammonia synthesis reactor. Purge gas contains Ammonia, Hydrogen and other trace gases. Ammonia can be recovered from purge gas to increase total Ammonia gain while the recovered Hydrogen can be re-entered as feed. In this way, the process can be made more efficient.

Ammonia Section of Plant-I, FFC Goth Machi employs Ammonia recovery Unit (ARU) and Hydrogen Recovery Unit (HRU) for treating purge gas. ARU recovers ammonia from purge and off gas. HRU serves for Hydrogen separation either by Cryogenic process or by using membranes. Hydrogen recovered is sent back to the Ammonia Synthesis Loop at conditions required.

1.2 Problem Statement

Plant-I of FFC Goth Machi utilizes purge gas in auxiliary and waste heat boilers. Uncondensed ammonia vapors in purge-gas contribute to the high NOx formation in the boilers causing an environmental issue apart from being a direct ammonia production loss. A solution to overcome Ammonia loss and high NOx generation at boilers must be found. Besides, selection of an efficient process for Hydrogen separation is also required.

1.3 Purpose of the Study

The purpose of this project is to address the environmental issues because of high NOx generation and to favor ammonia production gain. Selection of an efficient and economic technique for Hydrogen separation was also the focus of study.

The purpose mentioned above was met by selecting proficient processes through meticulous research. Various options were evaluated to overcome the high NOx generation at utilities boilers, including low-NOx burners in the synthesis loop apart from ARU installation. Based on detailed technical and economic analysis, ARU installation was found to be the most suitable option, which not only addresses the environmental issue but also results in ammonia production gain.

Separation technologies available for Hydrogen recovery were Membrane Separation, Cryogenic Separation and Pressure Swing Adsorption. Choice had to be made on the basis of recovery required, operating conditions and economic analysis. Pressure Swing Adsorption gave less recovery as compared to other two processes. Economic analysis showed Cryogenic Separation technique to be costly so Membrane were used for hydrogen separation.

1.4 Contributions of the Study

The processes described in this thesis report are a proposal to some of the problems faced by FFC in their Plant-I at Goth Machhi. Special considerations have been taken as to define all the processes in detail so that the industry can employ the advised techniques without many hindrances. The authors hope that the project will play a part, no matter how small, in resolving some of the issues identified in the report.

1.5 Brief History of Products:

1.5.1 Ammonia

Ammonia is a chemical compound made up of Nitrogen (N_2) and Hydrogen (H_2). The chemical formula of Ammonia is (NH_3). It has no color and is characterized by its pungent smell. It is of significant importance in the fields of food and fertilizers. Liquid Ammonia boils at −33.3 °C and has a freezing point of −77.7 °C. Its density is about one half the density of air.

Uses of Ammonia

i. Agricultural industries account for major use of Ammonia, utilizing almost 80% of the Ammonia.
ii. Liquid fertilizer solutions are made from ammonia which consists of aqua ammonia, ammonium nitrate and urea etc.
iii. Nitric acid is widely manufactured using Ammonia.
iv. Alkalies and dyes production also account for major use of Ammonia.
v. Ammonia finds its use in petroleum industry for neutralizing the acid constituents of crude oil.
vi. Variety of pharmaceuticals are also prepared by using Ammonia as building block.
vii. Ammonia is used as a refrigerant on a large scale in food, beverages and petro chemical industries.
viii. Dilute solutions of Ammonia are extensively used as detergents and household cleaners.

1.5.2 Hydrogen

Hydrogen (H_2) is a colorless and odorless gas. Its density is the lowest of all gases. It is a commercially important element. Melting point of Hydrogen is - 259.16°C while boiling point is - 252.879°C

Uses of Hydrogen

i. The largest use of hydrogen is in Haber's process to produce ammonia (NH_3), where nitrogen from air is combined with hydrogen to produce Ammonia.
ii. Liquid hydrogen finds its use in the study of superconductors and hydrogen blended with liquid oxygen, serves as an excellent fuel used in rockets etc.
iii. Hydrogenation of oils uses large quantities of hydrogen to convert them into fats, Margarine production is one such example.
iv. Hydrogen is also coming as an option for clean burning fuel in the near future. Hydrogen powered fuel cells are increasingly being used now a days.

v. One common use of hydrogen is as coolant due to its properties like lower density, lower viscosity and also because it surpasses other gases by having highest specific heat and thermal conductivity value.

Chapter 2: Literature Review

2.1 Ammonia Separation

2.1.1 Gas purification

There can be many possible ways to purify a gas but most commonly used method for gas purification is scrubbing or absorption. In absorption, gas is passed through a suitable solvent in an absorption column where the one or more components which needs to be separated are absorbed into the solvent depending upon their solubility. Basically mass transfer takes place during absorption.

There are two types of absorption:

- Physical Absorption
- Chemical Absorption

Physical absorption is simply a unit operation in which solute only dissolves into the solvent and form a solution. No new products are formed. In case of water and hydrocarbon solvents, physical absorption takes place.

In chemical absorption, chemical reaction takes place when a solute is dissolved in a solvent and new products are formed. When absorption of acid gases in basic solvent occurs, mostly chemical absorption occurs. Chemical absorption is also called reactive absorption.

For absorption, high pressure and low temperature conditions are required. Absorption depends on the gas velocity. If the gas velocity is too high, it will not allow liquid to come down the column and liquid will accumulate on the trays. This phenomena is called *flooding*. If the gas velocity is very low, more liquid will flow down the column, vapor-liquid contact time will be reduced and liquid accumulates at the bottom of column. This process is called *weeping*. Both these processes are undesirable and reduce the absorption efficiency. In case of packing, channeling occurs which is the phenomena in which liquid forms a particular path or channel and moves along that channel only.

Absorption column can be packed or tray type depending upon the column diameter and the gas and liquid load. Packing can be random or structured. There are different types of packing available like Pall ring, Raschig Ring, Berl Saddle etc.

Similarly there are different types of trays i.e. Sieve Trays, Bubble Cap Trays, Valve trays. In our process, two absorption columns are used. For purge gas, tray column is used where trays are sieve type. Sieve trays are simpler, cheaper, have high capacity and efficiency and needs low maintenance as compared to other trays. For off gas, packed column is used.

There are different solvents available for the absorption but we have chosen Water as solvent because it is easily available, cheap, universal solvent, Ammonia has high solubility in water as compared to other components in purge gas and only physical absorption takes place due to which aqueous solution of ammonia is formed.

2.1.2 Solvent Recovery

Water is used for ammonia absorption from Purge and Off Gas, as a result of which rich solution of ammonia and water is formed. Solvent should be recovered from the solution so that it can be recycled to make the process economical.

There are different ways to separate a solute and recover a solvent from it. Most common are:

- Stripping
- Distillation

Stripping:

Stripping is also separation technique similar to absorption but in stripping gas is separated from solvent. A liquid solution (gas absorbed in solvent), is passed through a gas and as a result of this, gas from solution is absorbed by the gas used for separation. It is also mass transfer process. Gas and liquid moves in opposite direction. Gas is introduced from the bottom of stripper while the liquid solution enters the column from the top. Gas takes the desired component from the solution and leaves the column from the stop. High temperature and low pressure is required for stripping. The gas used for separation depends on the solubility of solute and its availability. Different fluids can be used which includes steam, hydrocarbons as well as other inert gases.

There are various design parameters for stripper. Amount of solute in exit gas, fluid used for separation and the operating conditions are few of them.

Stripping column can be packed or tray type depending upon the column diameter and the gas and liquid load. Packing can be random or structured. There are different types of packing available like Pall ring, Raschig Ring, Berl Saddle etc.

Similarly there are different types of trays i.e. Sieve Trays, Bubble Cap Trays, Valve trays.

Distillation

Distillation is a process of separation of components from a mixture on the basis of volatility difference. The mixture is heated and then sent to a distillation column where lighter components move upward and heavier components move downwards. At each tray vapor – liquid equilibrium exits. Lighter components from the liquid are vaporized and heavier components from the vapor phase are condensed thus enriching the vapor phase with lighter components and liquid phase with the heavier components. Temperature decreases from bottom to top of column. At the top condenser is used for condensing the top product. Some amount of product is refluxed back to increase the product purity. At the bottom of column reboiler is used for heating the bottoms of column. Feed is entered at the plate into the column called feed plate. The section of the column above feed plate is called rectification section while the bottom section is called stripping section.

There are different types of trays i.e. Sieve Trays, Bubble Cap Trays, Valve trays used in distillation column.

In our process, distillation is used for recovery of solute from the solution. We used sieve trays because they are cheap, simple, easy to maintain, have high capacity and high efficiency. As our system is binary, no complex separation occurs. Water and ammonia have distinguished boiling points so they are easily separated.

We have chosen distillation over stripping because both top and bottom products are desirable. If we use any third component to separate the mixture, then that stripping agent will form a mixture with the solute and we have to design another unit to recover solute from that otherwise we have to compromise for one product as it will be lost.

2.2 Hydrogen Separation

There are various methods of hydrogen separation and purification, a few of them are given below:

2.2.1 Pressure Swing Adsorption (PSA)

PSA is a kind of gas separation technology which is used to separate gas or gases from a blend of gases under the pressure as indicated by the gas’s sub-atomic qualities and its affinity for the adsorbent. It works at close encompassing temperature ranges and varies essentially from the cryogenic gas separation systems. Particular adsorptive materials are utilized as a trap, specially adsorbing the objective gas type at high pressures. Then the pressure is lowered in the system to desorb the adsorbed material.

Pressure swing adsorption process relies upon the path that under the influence of high pressure, gases gets adsorbed on surfaces of solid substances. Higher the pressure will be, more will be the gas which is adsorbed. And by decreasing the pressure, the adsorbed gas is released. PSA systems can be used to disengage gases in a mix in light of the fact that unmistakable gases tend to be pulled into different solid surfaces basically unequivocally. If a gas mix, for instance, air is applied on with pressure through a vessel which contains a bed of zeolite as an adsorbent that pulls-in Nitrogen more firmly as compared to Oxygen, a little portion or most of Nitrogen will stay in the zeolite bed, while the gas leaving the tank will be progressed in oxygen. Exactly when the bed reaches its maximum capacity to adsorb the nitrogen gas, it is recouped by decreasing the pressure, consequently desorbing the adsorbed nitrogen. It is then arranged for another cycle of making oxygen-improved air.

2.2.2 Vacuum Swing Adsorption

Utilizing unique solids, adsorbents, VSA isolates certain gasses from a vaporous blend under insignificant pressure as indicated by the species' atomic attributes and partiality for the adsorbents. These adsorbents (e.g., zeolites) frame a sub-atomic strainer and specially adsorb the objective gas species at close encompassing pressure. The procedure then moves to a vacuum to recover the adsorbent.

VSA contrasts from the cryogenic distillation system and also pressure swing adsorption (PSA) methods since it works at close encompassing temperature and pressure ranges. VSA may really be best portrayed as a subset of the bigger class of PSA. It varies basically from Pressure Swing Adsorption in that PSA regularly vents to barometrical pressure, and uses a pressurized gas encourage into the division procedure. VSA normally draws the gas through the partition procedure with a vacuum. For Oxygen & Nitrogen VSA frameworks, the vacuum is commonly produced by a blower. VPSA frameworks apply pressurized gas to the partition procedure furthermore apply a vacuum to the purge gas. Usually, higher recuperation prompts a small compressor, a small blower, or other compacted gas. Higher profitability prompts smaller sieve beds. The consumer will no doubt consider indies which have an all the more straightforwardly quantifiable distinction in the general framework, similar to the measure of item gas isolated by the framework weight and size, the framework beginning and upkeep costs, the framework power utilization or other operational expenses, and dependability.

2.2.3 Cryogenic Distillation

An air separation facility isolates the atmospheric air into its essential components, ordinarily Nitrogen, Oxygen, Argon and other uncommon dormant gasses.

The most well-known strategy for the air separation and purification is cryogenic distillation system. Cryogenic Air Separation Units are worked to give Nitrogen, Oxygen and Argon. Different techniques, for example, Membrane technology, the Pressure Swing Adsorption and the Vacuum Pressure Swing Adsorption, are industrially used to separate a solitary segment from standard air. High immaculateness Oxygen, Nitrogen and Argon utilized for the semiconductor gadget creation requires the cryogenic distillation process.

Purge gases can be isolated from the air by lowering its temperature until the gas liquefies, then specifically refining the segments at their different bubbling temperatures. The procedure can deliver high purity gasses however is energy concentrated.

The cryogenic separation process need a tight coordination of heat exchangers and partition segments to acquire a decent proficiency and all the energy for refrigeration is given by the pressure of the air at the gulf of the unit.

To accomplish the low refining temperatures, the air separation unit needs the refrigeration cycle that works by method for the Joule–Thomson effect, and the cool hardware must be kept inside a protected chilly box. The cooling of the gases requires a lot of energy to work the refrigeration cycle and is done by the air compressor. Cutting edge ASUs use development turbines for the cooling purpose; the yield of the expander drives the air compressor, for enhanced proficiency. The procedure comprises of the following fundamental steps:

Air is packed where the last conveyance weight is dictated by recuperations and the liquid state (gas or fluid) of the items. Common pressures ranges somewhere around five and ten bar. The air stream may likewise be packed to various pressures to improve the proficiency of the ASU. Amid pressure water is consolidated out in between stage coolers.

The procedure air is by and large gone through a sub-atomic strainer bed, which expels the remaining water vapors, and carbon dioxide, which would stop and attachment the cryogenic hardware. Atomic sifters are frequently intended to expel any vaporous hydrocarbon materials from the air following these can be an issue in the ensuing air refining that could prompt explosions. The sub-atomic strainers bed must be recovered. This is finished by introducing various units working in substituting mode and utilizing the dry co-delivered waste gas to release the adsorbed water.

Process air is gone through a coordinated heat exchanger and then cooled against the cryogenic streams. Part of the air melts to shape a fluid that is enhanced in oxygen. The remaining gas is wealthier in nitrogen and is refined to verging on immaculate Nitrogen in High Pressure (HP) refining segment. The condenser of this segment needs refrigeration which is gotten from extending oxygen rich stream, further through an Expander.

On the other hand, the condenser might be cooled by trading the heat with the reboiler in LP refining segment when the ASU is delivering immaculate oxygen. To minimize the pressure cost the consolidated condenser and reboiler of the HP and LP sections must work with a temperature distinction of just 1-2 K, requiring plate balance brazed aluminum heat exchangers. Commonplace oxygen purities range from 97.5% to 99.5% and impacts the most extreme recuperation of oxygen. The refrigeration needed for delivering fluid items is gotten utilizing the Joule–Thomson effect as a part of an expander which sustains compacted air straightforwardly to the low pressure segment. Thus, a specific portion of air is not to be isolated and should abandon the low pressure segment as waste stream from upper area.

Since breaking point of Argon lies between oxygen and nitrogen, so Argon develops in the lower area of the low pressure section. At the point when argon is created, a vapor side with drawl is taken from the LP segment where Argon fixation is most astounding. It is sent to another section amending the argon to the sought immaculateness from which fluid is come back to the same area in the LP segment. Utilization of advanced organized packings with low pressure drops empower argon with under 1 ppm polluting influences. In spite of the fact that argon is available in less to 1% of the approaching, the air argon segment requires a lot of energy because of high reflux proportion required in the argon segment. Cooling of Argon section can be supplied from chilly extended rich fluid or by fluid nitrogen.

At last the items delivered in gas structure are warmed against the approaching air to encompassing temperatures. This requires a painstakingly made heat combination that must take into consideration strength against unsettling influences (because of switch over of the sub-atomic strainer beds). It might likewise require extra outer refrigeration amid start-up.

The isolated items are some of the time supplied by pipeline to expansive mechanical clients close to the creation plant. Long separation transportation of items is by delivery fluid item for extensive amounts or gas chambers for little amounts.

2.2.4 Membranes

Membrane separation processes work without heating and therefore utilize less energy than customary thermal partition procedures, e.g. refining, sublimation or crystallization. The division procedure is absolutely physical and both portions (penetrate and retentate) can be utilized. It is difficult to isolate the constituents of azeotropic fluids or solutes by refining or recrystallization however such partitions can be accomplished utilizing membrane technology.

Membranes for Gases:

For separation of gases, membranes can compete with other separation processes economically, environmental friendly, from safety and technical point of view. Various improvements in membranes have been made since its first time usage. Membranes can be integrated with the existing industrial processes.

History:

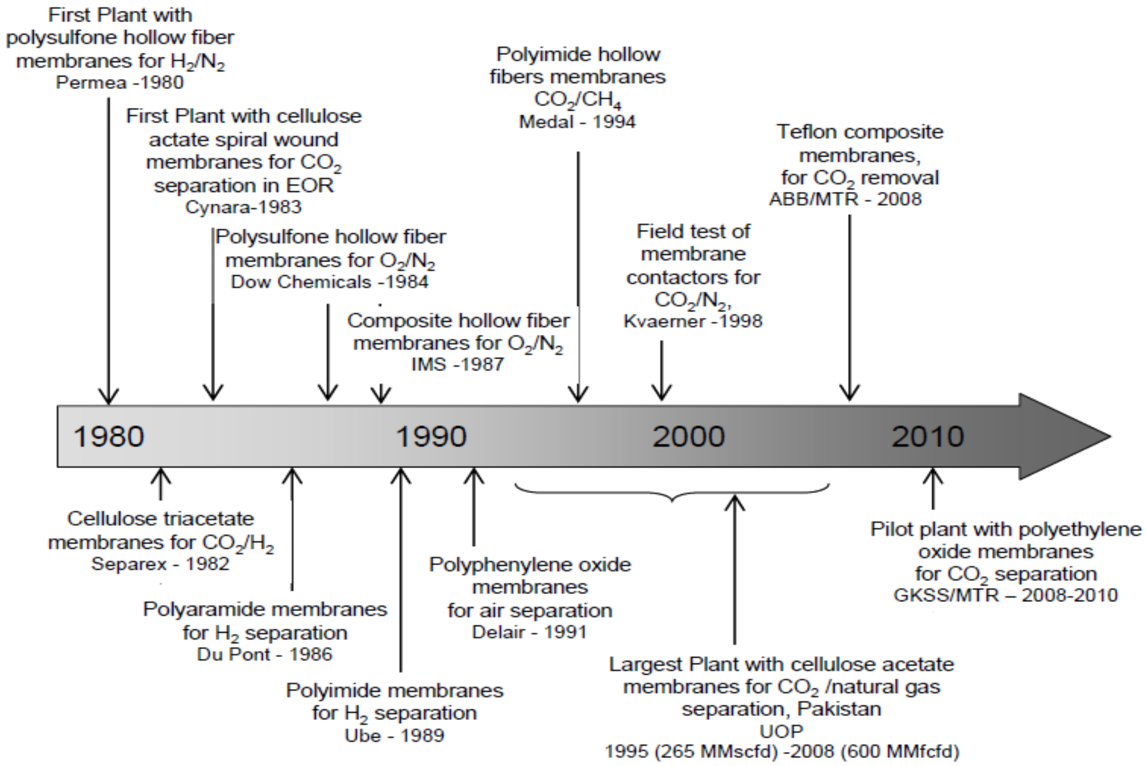


Figure 2.1: History of membranes (From Chemical Engineering Transactions Paper)

Different types of membranes are used for separation of different molecules. In our case, we are separating gas mixture so membranes for gas separation are used.

There are different categories of membranes and one particular membrane for our case is chosen on the basis of various factors.

Generally there are two major types of membranes:

- Porous Membranes
- Dense Membranes

Porous membranes has pores of various sizes through which particles pass. If the molecules have diameter smaller than the pore size then they will pass but if they have larger diameter than the pore size of membranes, they will remain behind. These membranes are used for separation of particles with large atomic diameter. In these membranes separation occurs on the basis of Knudsen diffusion, surface diffusion, capillary condensation and molecular sieving.

Dense membranes has no pores and used for separation of particles with very small particle diameter. These membranes work on the principle of Solution Diffusion theory. According to this theory, molecules are firstly absorbed to the surface of membrane, then they diffuse through the membranes, transported from one side to other and finally desorbed on the other side.

Molecules are separated on the basis of solubility of different molecules for particular membrane.

Another classification of membranes is on the basis of glass transition temperature.

- Glassy membranes
- Rubbery membranes

If material of membrane is below its glass transition temperature, such membrane is called glassy membrane while if the material of membrane is below its glass transition temperature, such membrane is called rubbery membrane. For smaller molecules as in our case, glassy membranes are used.

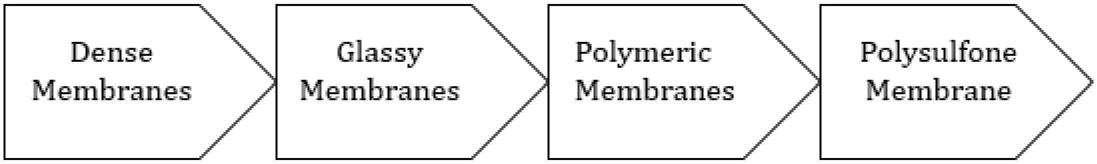
After above selection, material of membrane is selected. Membranes can be made from:

- Metals
- Ceramic
- Polymers

For gases, mostly polymeric membranes are used. For hydrogen separation, three different polymeric membranes are available.

- Polysulfone
- Polyimide
- Ethyl acetate

Membranes are selected on the basis of permeability, selectivity, mechanical strength and cost. Our main focus was selectivity and also considering other factors, we have chosen **Polysulfone** membrane.



Membranes performance depends on:

1- Stage Cut
2- Pressure ratio

Stage Cut: It is the ratio of permeate flowrate and feed flowrate

$$\Theta = \frac{Permeate\ flow}{Feed\ flow}$$

Higher is the value of stage cut, higher will be the recovery. It value should be 0.5 - 0.6.

Pressure ratio: It is the ratio of feed pressure and permeate pressure.

$$\Phi = \frac{Feed\ Pressure}{Permeate\ Pressure}$$

2.3 Equipment

2.3.1 Heat Exchanger

It is a device which is used to transfer heat between one or more than one fluids, which are either in direct contact or separated by a wall so that the liquids do not mix with each other. Heat Exchangers are widely used in various fields like natural gas processing, petroleum industry etc.

There are different types of Heat Exchangers, a few of them are given below:

Plate Heat Exchanger:

These are the kind of exchangers which have very large surface areas but also have small fluid passages and they are basically composed of thin plates which are slightly separated from each other. In comparison with the shell and tube heat exchanger, the plate heat exchanger have less volume & cost and these exchangers serve low to medium pressure fluids. Large sized Plate Heat Exchangers in HVAC are also known as plate and fin exchangers.

Plate and Shell Heat Exchangers:

The plate and shell type exchangers use passages which contain fins, to improve the efficiency of the unit. These are basically made up of alloys of aluminum as they provide better efficiency. This aluminum alloy tends to operate the system or machine at lower temperature difference.

These kind of exchangers are able to withstand high pressures and have high transfer efficiency. They also have a large surface area but the problem is that these are not easy to clean because of the narrow pathways.

Adiabatic wheel heat exchanger:

This is a type of heat exchanger which has a solid or liquid store which is used to store the heat and then allow it to be released from the other side of the exchanger.

Shell and Tube Heat Exchangers:

The heat exchangers used in the project are the shell and tube heat exchangers. This is the most commonly used type of heat exchanger and is used for high pressure systems.

This heat exchanger is composed of two main parts, the Shell and the Tube-Bundle. A fluid moves through the tube bundle while another fluid runs through the shell side, in turn providing a heat transfer between the two. Heat travels either from tubes to the shell side or from shell side fluid to the tube side fluid. The fluids used in the exchanger can be both liquid and gas. When operating a shell and tube heat exchanger, large heat transfer area must be used, because in this way, the waste heat can also be used up in the system.

Single phase and two phase heat exchangers are also used, i.e. same phases on either side and different phases on either sided respectively. The two phase exchanger can also be called a boiler as it is used to boil the liquid to its gas or vapor phase.

2.3.2 Reboiler

A reboiler is a type of heat exchanger used at the bottom of distillation column to produce vapors of the bottom product to improve separation. The exchanger is mostly of shell and tube type. The bottom liquid is partially vaporized and vapors are sent to column while remaining liquid is collected as a product. The stream used for heating is mostly steam but other fluids can also be used for heat exchange.

There are different types of reboilers and is chosen on the basis of following factors:

- Nature of the fluid e.g. viscosity, fouling tendency
- Operating conditions particularly operating pressure
- Layout of equipment

Types of Reboiler

There are different types of reboilers depending upon the orientation and circulation type.

1. *Kettle Reboiler:*

This type of reboiler is also called submerged bundle reboiler. Here tubes are immersed in the pool of liquid. Steam moves through the tube. These are used for the processes where heat exchange surface should be large, vapor quality should be high and high turndown ratio is required. Mostly they are used for vacuum applications and where high rate of vaporization is required.

2. *Thermosiphon:*

Thermosiphon is most common type of reboiler. These are mostly used in refining operations. The fluid to be vaporized is in the shell side while the heating fluid is kept in tube side. The boiling occurs in the shell side. A liquid vapor mixture is introduced into the column. It is of horizontal and vertical type. Maintenance of horizontal reboiler is simpler as compared to vertical but their fouling tendency is high. Vertical type are mostly found in chemical and petrochemical industries.

Density difference between vapor liquid mixture from reboiler and liquid from column to reboiler is the driving force for liquid circulation.

3. *Forced-Circulation Reboilers:*

In this type of reboiler, pump is used for circulating liquid. These are mostly used for highly viscous liquids and fluids with high fouling tendency. These are very expensive and uneconomical for common services.

2.3.3 Condenser

Condensation is a process where a gaseous state is converted into liquid state. It is of two types:

- Drop wise condensation
- Film wise condensation

Heat transfer coefficients are higher for dropwise condensation. In some condensers only phase change occurs while in some condensers, desuperheating and subcooling occurs as well. If the temperature of vapors is higher than the boiling point, firstly this super heat is removed to decrease its temperature till boiling point. Then phase change occurs at constant temperature. Sometimes the product is required at temperature below its boiling point, then after phase change cooling is done to remove its energy and decrease its temperature. This is called subcooling.

Condensers are similar to shell and tube heat exchangers where one fluid is inside the tubes and the other fluid is inside the shell. There can be different orientation of tubes. Either they can be horizontal or vertical.

Condensation can occur in following ways:

I. Condensation outside horizontal tubes i.e. on shell side

II. Condensation inside horizontal tubes

III. Condensation inside or outside vertical tubes

Every type of condenser has different formulae for calculation of heat transfer coefficient. Condenser is chosen on the basis of type of fluid being condensed and the cooling fluid used for condensation.

In our process, condensation occurs outside the horizontal tubes. Coolant is water which is inside the tubes. Our process requirement is to subcool the product, so firstly vapors are converted into liquid state and then extra heat is removed before collection of final product.

2.3.4 Pumps

Multi-Stage Centrifugal Pumps:

Multi-stage Centrifugal pumps have essentially the same principal as all multi-stage machinery such as heat exchangers and crystallizers or distillation columns. The multiple stages amplify the effect of one and build on it to create a much more powerful compounded effect than a single stage. For instance, it is very expensive and tedious to design a pump that can account for a pressure drop of 150 psig, but a Multi-Stage pump can do the job for a lower price. For a Multi-Stage Centrifugal Pump, there is more than one impeller either mounted on one or on different shafts. If impellers are connected in series, a greater pressure will be registered at the outlet but a greater flowrate will result when the impellers are connected in parallel. Multi-Stage Centrifugal Pumps are very common for the propulsion of Boiler Feed Water (BFW).

Diaphragm Pumps:

Diaphragm pumps are positive displacement pumps that employ some sort of rubber, Teflon or thermoplastic diaphragm that is used to pump fluids.

Peristaltic Pumps:

Peristaltic pumps are positive displacement pumps that work with the help of a rotor fitted with ‘shoes’ or ‘rollers’ that enable a forward pushing motion. Basically, the fluid is housed inside tubes that are then rolled or pushed forward by rotary motion. These pumps are useful for chemical dosing and prevent backflow due to controlled flow action.

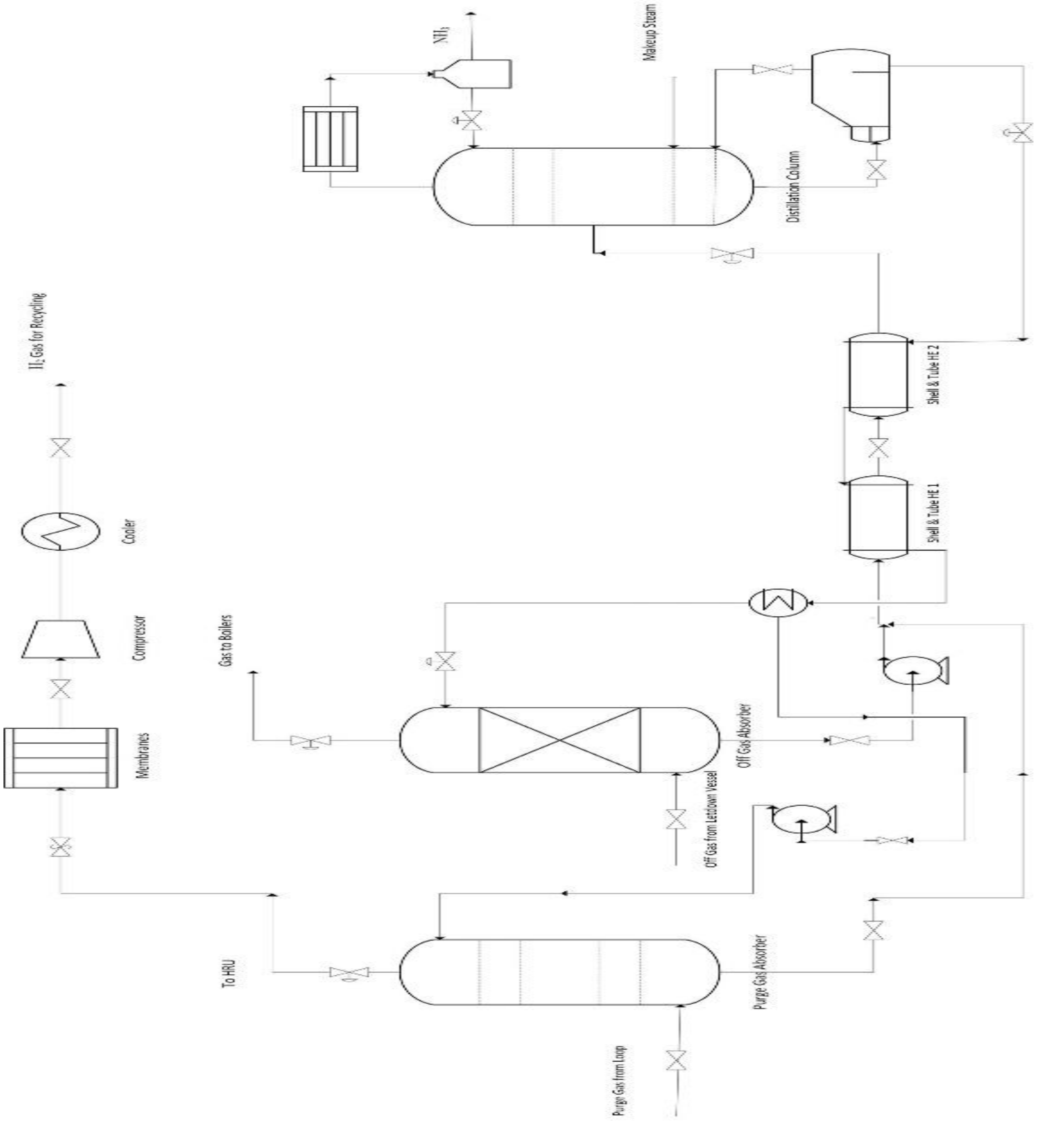
Hydraulic Pumps:

This is a kind of pump which converts mechanical energy into hydraulic energy to operate the hydraulic systems. The pump operate in a way that it creates a vacuum at the pump inlet so that the fluid is sucked into the inlet line of the pump, and then it is sent off towards hydraulic drive systems. They can further be divided into Hydrostatic and Hydrodynamic pumps which are positive displacement pumps and fixed displacement pumps respectively.

Chapter 3: Process Description

The processes described in this thesis are based on recovery. Ammonia is recovered from purge and off gas, which are coming from the main synthesis loop. After recovering ammonia from these gases, further treatment of the gases recovers Hydrogen from them. The whole process is described in detail below:

3.1 Process Flow Diagram



3.2 Process Description of ARU

Ammonia Absorption:

3.2.1 Purge Gas absorber

Purge gas from the synthesis loop enters the purge gas absorber from bottom at 12^0C and 231 bar. Flow of purge gas is 22,499 Nm^3/hr. Water (the liquid phase for absorption) enter from the top of the column. This water contains little amount of ammonia. Ammonia gets absorbed into the water (because of its affinity for water) forming aqueous solution andt leaves from the bottom as rich solution. Rest of the gas mixture i.e. purified purge gas leaves from the top of the column for recovery of hydrogen to Hydrogen recovery Unit.

3.2.2 Off-Gas Absorber

Off gas from the letdown vessel (vessel in main process) enters the second absorber at 40^0C and 24.1 bar. Flow of off gas is 1066 Nm^3/hr. Water enters from top and off Gas from bottom and absorption takes place as explained above. Rich Solution of Ammonia leaves from the bottom while unabsorbed gas mixture go to boilers from the top of the column.

The bottom product i.e. rich solution of both absorbers are combined. These are sent to distillation column so that water used can be recovered and recycled. Also ammonia is the desired product so it is also collected and stored as product.

Before distillation column, this solution is heated by a series of shell and tube heat exchangers so that distillation temperature is achieved.

3.2.3 Distillation Column

Rich solutions from both the absorbers are entered as feed to the feed plate in distillation column. The temperature and pressure of both the streams are adjusted before both of these are mixed and entered as single feed to the column. The temperature of feed is 163^0C. Separation of ammonia and water occurs because of their difference in volatility. Ammonia vapors leave from the top of the column, some are refluxed while remaining vapors are taken out. Lean solution (which is almost pure water) leaves from the bottom and is used as solvent in both absorbers after cooling.

Ammonia vapors from the top of column are sent to condenser where condensation takes place outside the horizontal tubes and cooling water used for condensation is passed through the tubes. Ammonia vapors are its boiling point. In the condenser not only condensation takes place but sub cooling is also done where its temperature is reduced below its condensation temperature. Some of liquid ammonia is refluxed back to column while remaining is collected as a product.

3.2.4 Rich Solution Exchanger and LP Circulation Pump

For distillation of Rich solution, it must enter the column at high temperature (163^oC) and appropriate pressure. So rich solution from both the absorbers pass through heat exchanger for temperature gain. The hot fluid in exchanger is the high temperature lean solution (water) of distillation column. Pressure of rich solution coming out of purge gas absorber is enough, so it enters heat exchanger without pumping. Rich solution coming out of Off-gas absorber is at low pressure hence it is passed through LP circulation pump for bringing it up to feed plate pressure. The hot lean solution is in the shell side while the cold rich solution is passed through the tubes where heat transfer takes place.

3.2.5 Lean Solution Cooler

This is a heat exchanger which further cools the lean solution (water) coming out from the bottom of the distillation column. Lean solution after transferring heat to rich solution in the two heat exchangers is still at high temperature. So further heat is removed by cooling with water. Temperature of exchanger is 38^0C at the exit of this heat exchanger. The recovered water is then used in both absorbers as a solvent.

3.3 Process Description of HRU

3.3.1 Expansion Valve

Gas mixture coming from the top of absorber 1 enters the Hydrogen Recovery Unit (HRU). This gas is at very high pressure i.e. 120 kg/cm^2_g. HRU starts with expansion valve to release the high pressure as membranes are very thin. They cannot withstand very high pressure and can get damaged by high pressure. Gas mixture after passing through expansion valve is ready to be treated further. Pressure is reduced to 60 kg/cm^2_g. As the gas expands, its temperature is also decreased.

3.3.2 Membranes

After expansion valve, the gas stream is passed through the membranes so that hydrogen can be recovered. The membrane used for separation is polysulfone which is dense, glassy type membrane. The arrangement of membranes is hollow fibers. Separation takes place by the phenomenon of solution diffusion. As hydrogen has high permeability as compared to other components of stream for polysulfone membranes, hydrogen will pass through membrane as permeate while the retentate contains other components. As the feed and permeate pressure ratio is 10 so the permeate leaves at the pressure of 6 kg/cm^2_g. Permeate contains 98% hydrogen. By membranes, 83.7% recovery of hydrogen is achieved. Permeate is then sent to compressors for further processing.

3.3.3 Compressor

Hydrogen is used as a raw material for the synthesis of ammonia so the Hydrogen recovered from the gas mixture has to be sent off to the main synthesis loop. This recycled hydrogen is mixed with the fresh hydrogen stream so it should be at the operating conditions of fresh hydrogen stream. For bringing it up to the pressure required at the loop, the gas is passed through multistage compressor. Its pressure is increased from 6 kg/cm^2_g to 28 kg/cm^2_g. Due to compression, gas temperature increases which is higher than the required for mixing. After compressor, it heads on to the cooler to lower its temperature.

3.3.4 Cooler/ Heat Exchanger

To bring the gas to the loop temperature and also to release the heat of compression, it is cooled in a cooler. After heat exchange in this heat exchanger, the gas is ready to be entered in the main synthesis loop for its reuse.

Calculations:

Chapter 4: Material Balance

4.1 Absorber 1

Purge Gas = P_1

Lean Solution (Water) = W_1

Purified Purge Gas = G_1

Rich Solution (water after absorbing ammonia from purge gas) = R_1

Purge Gas Components: Hydrogen, Nitrogen, Argon, Methane, Ammonia

Conditions: Molar Flow of Purge Gas = 1004 kgmol/hr

Inlet Temperature = 12 ^oC

Inlet Pressure = 231 kg/cm^2_g

Table 4. 1: Inlet Gas Composition (Absorber 1)

Components	Compositions
H_2	59.5 %
N_2	20.3 %
Ar	2.1 %
CH_4	12.6 %
NH_3	5.5 %

Solvent for Absorption: Water

Conditions: Molar Flow of lean sloution = W_1 (kgmol/hr)

Inlet Temperature = 40 ^oC

Inlet Pressure = 120 kg/cm^2_g

Composition of lean solution is assumed as shown below:

Table 4. 2: Inlet Water Composition (Absorber 1)

Components	**Compositions**
H_2O	99 %
NH_3	1 %

H_2, N_2, CH_4 and Ar acts as inerts and they don’t dissolve in water. These components comes in same amount in exit gas.

It is assumed that the exit gas contains 0.2% ammonia.

Apply component balance on these inerts.

H_2:

H_2 in P_1 = H_2 in G_1

0.595 * 1004 = H_2 in G_1 = 597.38 kgmol/ hr

N_2:

N_2 in P_1 = N_2 in G_1

0.203 * 1004 = N_2 in G_1 = 203.812 kgmol/ hr

Ar:

Ar in P_1 = Ar in G_1

0.021 * 1004 = Ar in G_1 = 21.084 kgmol/ hr

CH_4:

CH_4 in P_1 = CH_4 in G_1

0.126 * 1004 = CH_4 in G_1 = 126.504 kgmol/ hr

$N_2 + H_2 + Ar + CH_4 + NH_3 = G_1$

$597.38 + 203.812 + 21.084 + 126.504 + 0.002G_1 = G_1$

$948.78 + 0.002G_1 = G_1$

$948.78 = 0.998G_1$

G_1 = 950.68 kgmol/ hr

Let in rich solution R_1,

NH_3 = 15%, H_2O = 85% and in water 1 stream W_1,

H_2O = 99%, NH_3 = 1%

Apply ammonia balance,

Ammonia in purge gas + Ammonia in Water = Ammonia in exit gas + Ammonia in Rich Solution

$(0.055 * 1004) + (0.01 * W_1) = (0.002 * G_1) + (0.15 * R_1)$

$55.22 + 0.01 W_1 = 1.90136 + 0.15 R_1$

$53.31864 = 0.15 R_1 - 0.01 W_1$ ------------ (4.1.1)

Apply water balance,

Water in Lean Solution = Water in Rich Solution

$0.99 W_1 = 0.85 R_1$

$W_1 = 0.859 R_1$ ------------- (4.1.2)

Put (4.1.2) in (4.1.1)

$53.31864 = 0.15 R_1 - 0.01 (0.859 R_1)$

$53.31864 = 0.15 R_1 - 0.00859 R_1$

$53.31864 = 0.14141\ R_1$

$R_1 = 377$ kgmol/ hr

From (4.1.2), $W_1 = 0.859 * 377$

$W_1 = 323.843$ kgmol/ hr

Also by applying overall molar flow balance on the absorber,

$P_1 + W_1 = R_1 + G_1$

$W_1 = 377 + 950.68 – 1004$

$W_1 = 323.68$ kgmol/ hr

Table 4. 3: Flowrates (Absorber 1)

Streams	Flowrates (kgmol/hr)
Purge Gas P_1	1004
Purified Purge Gas G_1	950.68
Lean Solution (water) W_1	323.843
Rich Solution R_1	377

Table 4. 4: Gas Compositions (Absorber 1)

Components	Inlet Compositions	Outlet Compositions
H_2	59.5 %	63 %
N_2	20.3 %	21.5 %
Ar	2.1 %	2 %
CH_4	12.6 %	13.3 %
NH_3	5.5 %	0.2 %

4.2 Absorber 2

Off Gas = P_2

Lean Solution (Water) = W_2

Purified Off Gas = G_2

Rich Solution (water after absorbing ammonia from Off gas) = R_2

Off Gas Components: Hydrogen, Nitrogen, Argon, Methane, Ammonia

Conditions: Molar Flow of Off Gas = 48 kgmol/hr

Inlet Temperature = 40 ^{o}C

Inlet Pressure = 24.1 kg/cm^2_g

Table 4. 5: Inlet Gas Composition (Absorber 2)

Components	**Compositions**
H_2	43 %
N_2	17.6 %
Ar	2.0 %
CH_4	20.2 %
NH_3	17.2 %

Solvent for Absorption: Water

Conditions: Molar Flow of lean sloution = W_2 (kgmol/hr)

Inlet Temperature = 38 ^{o}C

Inlet Pressure = 20 kg/cm^2_g

Composition of lean solution is assumed as shown below:

Table 4. 6: Inlet Water Composition (Absorber 2)

Components	Compositions
H_2O	99 %
NH_3	1 %

H_2, N_2, CH_4 and Ar acts as inerts and they don't dissolve in water. These components comes in same amount in exit gas. It is assumed that the exit gas contains 0.1% ammonia.

Apply component balance on these inerts.

H_2:

H_2 in P_2 = H_2 in G_2

0.43 * 48 = H_2 in G_2 = 20.64 kgmol/ hr

N_2:

N_2 in P_2 = N_2 in G_2

0.176 * 48 = N_2 in G_2 = 8.448 kgmol/ hr

Ar:

Ar in P_2 = Ar in G_2

0.02 * 48 = Ar in G_2 = 0.96 kgmol/ hr

CH_4:

CH_4 in P_2 = CH_4 in G_2

0.202 * 48 = CH_4 in G_2 = 9.696 kgmol/ hr

NH_3 in off gas = 0.1 %

$G_2 = H_2 + N_2 + Ar + CH_4 + NH_3$

$G_2 = 20.64 + 8.448 + 0.96 + 9.696 + 0.001\ G_2$

$0.999\ G_2 = 39.744$

$G_2 = 39.784$ kgmol/ hr

In rich solution R_2, $NH_3 = 10\%$, $H_2O = 90\%$

In water 2, $H_2O = 99\%$, $NH_3 = 1\%$

Apply Ammonia balance,

NH_3 in Off gas + NH_3 in Water = NH_3 in exit gas + NH_3 in Rich Solution

$0.172 * 48 + 0.01\ W_2 = 0.001\ G_2 + 0.1\ R_2$

$8.256 + 0.01\ W_2 = 0.039784 + 0.1\ R_2$

$8.216216 = 0.1R_2 - 0.01\ W_2$ ----------- (4.2.1)

Apply Water balance,

$0.99\ W_2 = 0.9\ R_2$

$R_2 = 1.1\ W_2$ ------------- (4.2.2)

Put (4.2.2) in (4.2.1)

$8.216216 = 0.1\ (1.1\ W_2) – 0.01\ W_2$

$8.216216 = 0.11\ W_2 - 0.01\ W_2$

$W_2 = 82.16216$ kgmol/ hr

By applying overall molar flow balance on the absorber, we get

$P_2 + W_2 = R_2 + G_2$

$48 + 82.16216 = R_2 + 39.78$

$R_2 = 90.38216$ kgmol/ hr

Streams	**Flowrates (kgmol/hr)**
Off Gas P_1	1004
Purified Off Gas G_1	950.68
Lean Solution (water) W_1	323.843
Rich Solution R_1	377

Table 4. 7: Gas Compositions (Absorber 2)

Components	**Inlet Compositions**	**Outlet Compositions**
H_2	43 %	52.3 %
N_2	17.6 %	21.2 %
Ar	2.0 %	2.3 %
CH_4	20.2 %	24.1 %
NH_3	17.2 %	0.1 %

4.3 Mixer

Rich Solution from absorber 1 + Rich Solution from absorber 2 = Total Rich Solution

$R_1 + R_2 = R_3$

$377 + 90.38216 = R_3$

$R_3 = 467.38216$ kgmol/ hr

H_2O in $R_3 = x$

Apply water balance,

$0.85 * R_1 + 0.9 R_2 = x R_3$

$0.85 * 377 + 0.9 * 90.38216 = x * 467.38216$

H_2O x = 0.86

NH_3 1- x = 1- 0.86 = 0.14

4.4 Distillation Column

Feed of distillation column = F_1

Top product = A

Bottom Product = L

Vapors coming from top of column = V

Reflux amount = R

Now, Feed of Distillation:

R_3 = 467.38216 kgmol/hr = F_1

H_2O = 86%, NH_3 = 14%

Assumption:

Ammonia achieved at top is 99.5% and lean solution at the bottom contains 99% water and 1% ammonia.

Apply Ammonia balance:

Ammonia in feed = Ammonia in top product + Ammonia in bottom product

$0.14 * 467.38216 = 0.995 * A + 0.01L$

$65.434 = 0.995A + 0.01L \dashrightarrow \quad (4.4.1)$

Apply Water balance:

Water in feed + Make up Steam = Water in top product + Water in bottom product

$0.86 * 467.38216 + M_1 = 0.0005A + 0.99L \dashrightarrow \quad (4.4.2)$

As we know,

$L = W_1 + W_2$

L = 323.68 + 82.16216 **= 405.84216 kgmol/ hr**

From (4.4.2),

65.434 = 0.995A + (0.01 * 405.84216)

A = 61.684 kgmol/ hr

By applying overall balance around distillation column,

Feed + Make up Steam = Top Product + Bottom Product

$F_1 + M_1 = A + L$

$M_1 = 61.684 + 405.84216 – 467.38216$

M_1 = 0.144 kgmol/ hr

Reflux ratio: 3

R = reflux amount = 3 * A = 3 * 61.684 = 185.052 kgmol/hr

Vapors coming out from distillation column = Reflux + Top Product = A + R = 61.684 + 185.052

V = 246.736 kgmol/ hr

Table 4. 8: Flowrates (Distillation Column)

Streams	Flow rates (kgmol/hr)
Feed F_1	467.382
Top Product A	61.684
Bottom Product L	405.842
Reflux Amount R	185.052
Vapors from top V	246.736
Make up Steam M	0.144

Table 4. 9: Product Compositions (Distillation)

Components	Feed	Top Product	Bottom Product
NH_3	0.14	0.995	0.99
H_2O	0.86	0.005	0.10

4.5 Hydrogen Recovery Unit (HRU)

Membranes:

Table 4. 10: Inlet and Outlet Gas Composition (HRU)

Components	Inlet Compositions	Outlet Compositions
H_2	63 %	97.92 %
N_2	21.5 %	1.27 %
Ar	2 %	0.02 %
CH_4	13.3 %	0.788 %
NH_3	0.2 %	0.002 %

Table 4. 11: Operating Conditions (HRU)

Properties	Feed	Permeate	Retentate
Flowrate (kgmol/hr)	950	511.98	438.02
Temperature (^oC)	4.93	4	4.9
Pressure (kg/cm^2_g)	60	6	59

Chapter 5: Energy Balance

5.1 Absorber 1

$Q_{abs} = Q_g + Q_L$

$Q = m\ C_p\ \Delta T$

For gas,

M_{gas} = 948.78 Kgmol/hr

$C_{p\ =}$ 33 KJ/kgmol. ^oC

Inlet temperature of gas = 12 ^oC

Outlet temperature of gas = 52 ^oC

$Q_{gas} = m\ C_p\ \Delta T$

= 948.78*33*(52-12)

= 1252389.6 KJ/hr

For Lean Solution,

M_L = 320.45 Kgmol/hr

$C_{p\ =}$ 77.86 KJ/kgmol. ^oC

Inlet temperature of lean solution = 40 ^oC

Outlet temperature of lean solution = 60 ^oC

$Q_L = m\ C_p\ \Delta T$

= 320.45*77.86*(60-40) = 499004.74 KJ/hr

$Q_{abs} = Q_g + Q_L$

Q_{abs} = 1252389.6 + 499004.74

= 1751394.34 KJ/hr

Ammonia absorbed= 56.55 kgmol/hr

Heat of ammonia solution = 34000 KJ/hr

$Q_{ammonia}$ = 1922700 KJ/hr

5.2 Absorber 2

$Q_{abs} = Q_g + Q_L$

$Q = m\ C_p\ \Delta T$

For gas,

M_{gas} = 39.744 Kgmol/hr

$C_{p\ =}$ 31 KJ/kgmol. ^oC

Inlet temperature of gas = 40 ^oC

Outlet temperature of gas = 45 ^oC

$Q_{gas} = m\ C_p\ \Delta T$

= 39.744*31*(45-40)

= 6160.32 KJ/hr

For Lean Solution,

M_L = 81.344 Kgmol/hr

C_p = 77.86 KJ/kgmol. ^oC

Inlet temperature of lean solution = 38 ^oC

Outlet temperature of lean solution = 85 ^oC

$Q_L = m\ C_p\ \Delta T$

= 81.344*77.86*(85-38) = 297671.9 KJ/hr

$Q_{abs} = Q_g + Q_L$

Q_{abs} = 6160.32 + 297671.9 = 303832.2 KJ/hr

Ammonia absorbed= 9.038 kgmol/hr

Heat of ammonia solution = 34000 KJ/hr

$Q_{ammonia}$ = 307292 KJ/hr

5.3 Lean Solution Heat Exchanger 1

Tube Side: Rich solution

Shell Side: Lean Solution

Lean solution inlet temperature = T_1 = 166.5 ^oC

Lean solution outlet temperature = T_2 = 104 ^oC

Rich solution inlet temperature = t_1 = 60 ^oC

Rich solution outlet temperature = t_2

C_p of Lean Solution = 75 KJ/kgmol. ^oC

C_p of Rich Solution = 79 KJ/kgmol. ^oC

Molar flow rate of lean solution = 405.8825 kgmol/hr

Molar flow rate of rich solution = 467.4173 kgmol/hr

$Q_{shell} = m\ C_p\ \Delta T$

$= 405.8825 * 75 * (104-166.5) = -1902574$ KJ/hr

Energy given by shell side = Energy gained by tube side

$Q_{shell} = -Q_{tube}$

$Q_{tube} = m\ C_p\ \Delta T$

$1902574 = 467.4173 * 79 * (t_2 - 60)$

t_2 = 111.5 ^oC

5.4 Lean Solution Heat Exchanger 2

Tube Side: Rich solution

Shell Side: Lean Solution

Lean solution inlet temperature = T_1 = 229 ^oC

Lean solution outlet temperature = T_2 = 166.5 ^oC

Rich solution inlet temperature = t_1 = 111.5 ^oC

Rich solution outlet temperature = t_2

C_p of Lean Solution = 75 KJ/kgmol. ^oC

C_p of Rich Solution = 79 KJ/kgmol. ^oC

Molar flow rate of lean solution = 405.8825 kgmol/hr

Molar flow rate of rich solution = 467.4173 kgmol/hr

$Q_{shell} = m\ C_p\ \Delta T$

$= 405.8825 * 75 * (166.5-229) = - 1902574$ KJ/hr

Energy given by shell side = Energy gained by tube side

$Q_{shell} = - Q_{tube}$

$Q_{tube} = m\ C_p\ \Delta T$

$1902574 = 467.4173 * 79 * (t_2 - 111.5)$

t_2 = 163 ^oC

5.5 Lean Solution Cooler

Tube Side: Cooling Water

Shell Side: Lean Solution

Lean solution inlet temperature = T_1 = 104 ^oC

Lean solution outlet temperature = T_2 = 38 ^oC

Cooling water inlet temperature = t_1 = 30 ^oC

Cooling water outlet temperature = t_2 = 60 ^oC

C_p of Lean Solution = 75 KJ/kgmol. ^oC

C_p of cooling water = 77.5 KJ/kgmol. ^oC

Molar flow rate of lean solution = 405.8825 kgmol/hr

Molar flow rate of cooling water = m_{water}

$Q_{shell} = m\ C_p\ \Delta T$

$= 405.8825 * 75 * (38-104)$

$= -\ 2116271$ KJ/hr

Energy given by shell side = Energy gained by tube side

$$Q_{shell} = -\ Q_{tube}$$

By applying energy balance, amount of cooling water is calculated.

$Q_{ube} = m_{water}\ C_p\ \Delta T$

$2116271 = m_{water} * 77.5 * (60-30)$

m_{water} = 910.224 kgmol/hr

5.6 Condenser

Apply energy balance around condenser

Energy of the condenser	+ Energy of the distillate	+ Energy of the reflux	= Energy of the vapors coming into condensers

$$Q_c + H_D + H_L = H_V$$

Energy of the condenser = Q_c

Energy of the distillate = H_D

Energy of the reflux = H_L

Energy of the vapors coming into condensers = H_V

C_p of distillate = 89.17 kJ/kgmol. ^oC

C_p of reflux= 89.17 kJ/kgmol. ^oC

Temperature of vapors = 61 ^oC

Temperature of distillate = 40 ^oC

Temperature of reflux = 40 ^oC

Distillate flow rate = 61.684 kgmol/hr

Reflux flow rate = 185.53 kgmol/hr

Vapor flow rate = 246.73 kgmol/hr

Energy of the distillate = $H_D = m * C_p * \Delta T$

$= 82719.54$ KJ/hr

Energy of the reflux = $H_L = m * C_p * \Delta T$

$= 248158.6$ KJ/hr

Heat of vaporization = 18650 KJ/kgmol

Energy of vapors = $H_V = V * \lambda = 246.73 * 18650 = 4601515$ KJ/hr

$Q_c = H_V - H_D – H_L$

$Q_c = 4601515 - 82719.54 - 248158.6$

$Q_c = 4270636$ KJ/hr

Water required for condensation and cooling of product is calculated as below:

C_p of water = 77.72 KJ/kgmol. ^oC

Temperature rise for cooling water is limited to 30 ^oC

$Q = m\ C_p\ \Delta T$

$M_{water} = \frac{Q}{\mathrm{Cp}\ \Delta \mathrm{T}} \quad = \frac{4270636}{77.72*30}$ **= 1831.6 kgmol/hr**

5.7 Reboiler

Reboiler energy is calculated by applying overall energy balance on the

Energy given to reboiler + Energy of the feed = Energy of condenser + Energy of distillate + Energy of bottom product

$$Q_B + H_F = Q_c + H_D + H_B$$

Energy of the condenser = Q_c

Energy of the distillate = H_D

Energy of the bottom product = H_B

Energy of the reboiler = Q_B

Energy of the feed = H_F

C_p of distillate = 89.17 kJ/kgmol. ^{o}C

C_p of feed = 85 kJ/kgmol. ^{o}C

C_p of bottom product = 41 kJ/kgmol. ^{o}C

Temperature of feed = 163 ^{o}C

Temperature of distillate = 40 ^{o}C

Temperature of bottom product = 229 ^{o}C

Distillate flow rate = 61.684 kgmol/hr

Bottom flow rate = 405.84 kgmol/hr

Feed flow rate = 467.38 kgmol/hr

Energy of the distillate = $H_D = m * C_p * \Delta T$

$= 82719.54$ KJ/hr

Energy of the bottom product = $H_B = m * C_p * \Delta T$

$= 3394801$ KJ/hr

Energy of the feed = H_F = m * C_p * ΔT = 5482805 KJ/hr

$Q_B = Q_c + H_D + H_B - H_F$

= 4270636 + 82719.54 + 3394801 – 5482805

= 2265352 KJ/hr

Amount of steam required for reboiler is calculated as below:

Steam is available at 248 ^oC,

Latent heat of steam = 31670 KJ/kgmol

Steam required = $\frac{\mathrm{Q}\ boiler}{\lambda}$ = 2265352 / 31670

Steam required **= 71.529 kgmol/hr**

5.8 Hydrogen stream Cooler

Tube Side: Cooling Water

Shell Side: Hydrogen stream

Hydrogen stream inlet temperature = T_1 = 147 ^oC

Hydrogen stream outlet temperature = T_2 = 100 ^oC

Cooling water inlet temperature = t_1 = 25 ^oC

Cooling water outlet temperature = t_2 = 55 ^oC

C_p of Hydrogen stream = 28.94 KJ/kgmol ^oC

C_p of cooling water = 77.5 KJ/kgmol. ^oC

Molar flow rate of Hydrogen stream = 505.3 kgmol/hr

Molar flow rate of cooling water = m_{water}

$Q_{shell} = m\ C_p\ \Delta T$

= 505.3 * 28.94 * (100-147) = - 687299 KJ/hr

Energy given by shell side = Energy gained by tube side

$$Q_{shell} = - Q_{tube}$$

By applying energy balance, amount of cooling water is calculated.

$Q_{ube} = m_{water}\ C_p\ \Delta T$

$687299 = m_{water} * 77.5 * (55\text{-}25)$

m_{water} = 295.6 kgmol/hr

Chapter 6: Designing

6.1 Heat Exchanger 1

Exchanger type: Shell and tube Heat Exchanger

Shell Passes = 1

Tube Passes = $N_p = 2$

Shell Side Fluid: Lean Solution

Tube Side Fluid: Rich Solution

Lean solution inlet temperature = $T_1 = 166.5\ ^oC$

Lean solution outlet temperature = $T_2 = 104\ ^oC$

Rich solution inlet temperature = $t_1 = 60\ ^oC$

Rich solution outlet temperature = $t_2 = 111.5\ ^oC$

C_p of Lean Solution = 79 KJ/kgmol. ^oC

C_p of Rich Solution = 75 KJ/kgmol. ^oC

Molar flow rate of lean solution = 405.8825 kgmol/hr

Molar flow rate of rich solution = 467.4173 kgmol/hr

$Q_{shell} = m\ C_p\ \Delta T$

$= 405.8825 * 79 * (104-166.5)$

$= - 1902574$ KJ/hr $= - 1902574 / 3600 = - 528.5$ KW (KJ/sec)

Flowrate of shell side fluid = 7303 kg/hr

$Q=UA\Delta T_m$

$\Delta T_m = F_t * LMTD$

$$LMTD = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln\frac{(T_1 - t_2)}{(T_2 - t_1)}}$$

Putting values of inlet and outlet temperatures of shell and tube side fluids,

LMTD = 49.3 ^{o}C

$$R = \frac{(T_1 - T_2)}{(t_2 - t_1)} = 1.21$$

$$S = \frac{(t_2 - t_1)}{(T_1 - t_1)} = 0.48$$

From Figure 6.1 for 1 shell pass and 2 tube passes using values of S and R, we get

F_t = 0.713

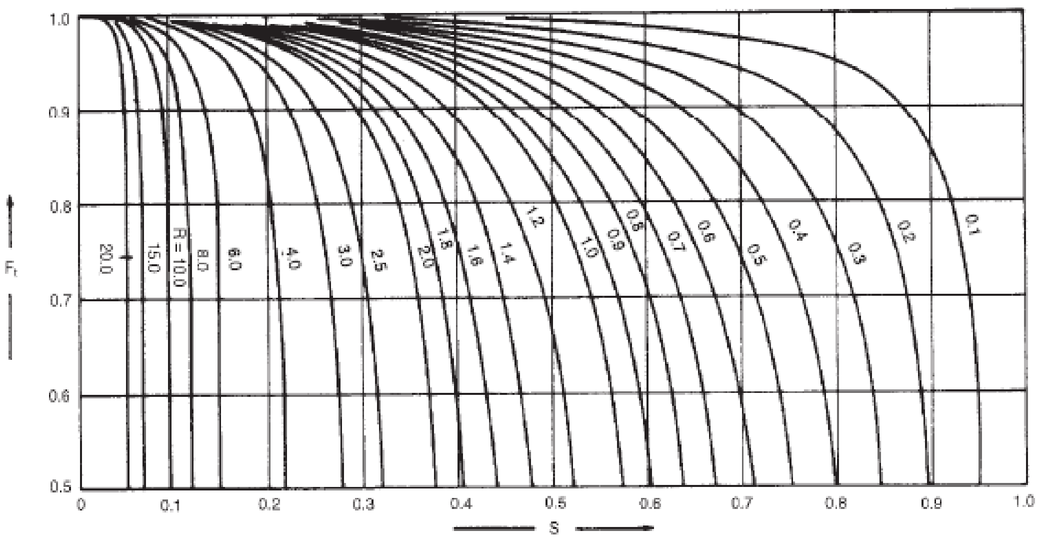


Figure 6.1: Correction factor (From Coulson and Richardson Volume 6)

$\Delta T_m = F_t * LMTD$

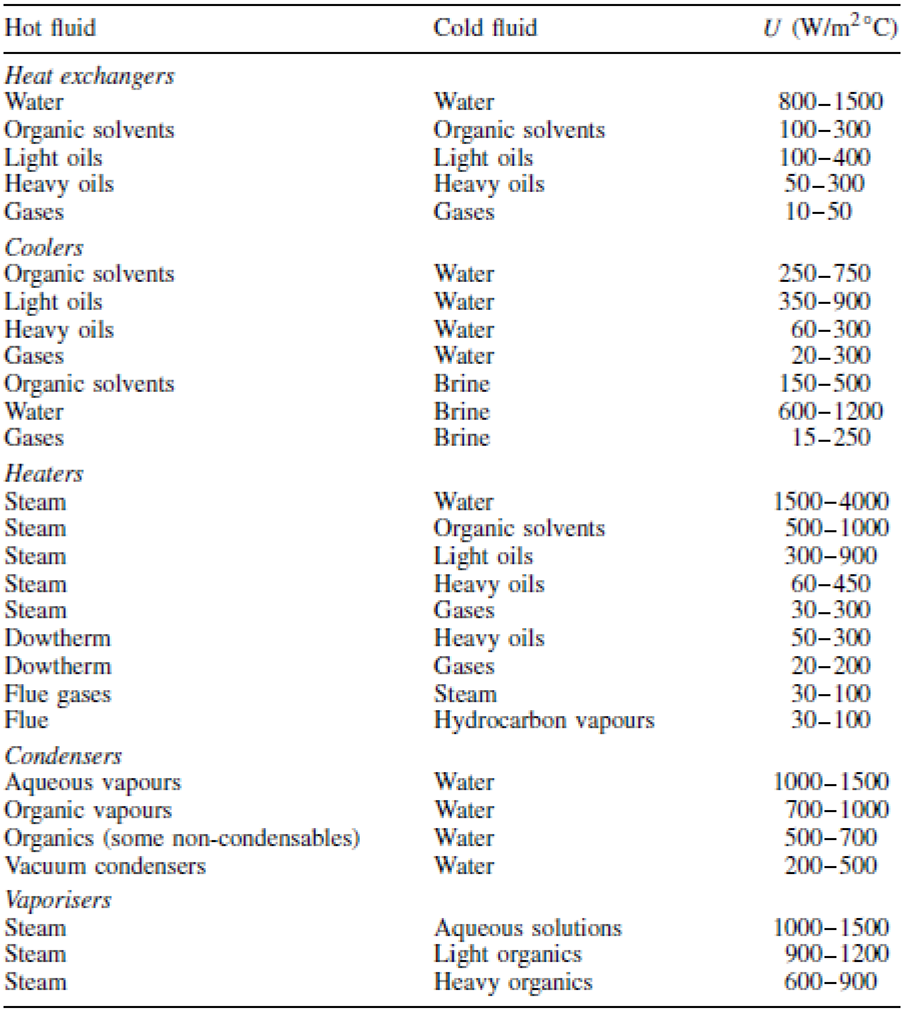
$\Delta T_m = 0.713 * 49.3 = 35.19$ ^{o}C

Overall heat transfer coefficient for this system is assumed.

U = 800 W/m^2. ^{o}C

Heat transfer Area = $A = \frac{Q}{U\Delta \mathrm{Tm}} = \frac{528.5}{800*35.19} = 18.77\ \mathrm{m}^2$

Table 6. 1: Heat Transfer Coefficient



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>Assumptions:</u>

Choose 20 mm outer diameter, 16 mm inner diameter and 3.66 m long tubes

Inner diameter of tube = d_i = 0.016 m

Outer Diameter of tube = d_o = 0.02 m

Length of tube = L = 3.66 m

Area of one tube = $\pi * L * d = \pi * 3.66 * 0.02 = 0.23\ \mathrm{m}^2$

Number of tubes = N_t = $\frac{Heat\ transfer\ area}{Area\ of\ one\ tube}$ = 18.77 / 0.23 = 82 tubes

Use triangular pitch,

Pitch = p_t = 1.25 * d_o = 1.25 * 0.02 = 0.025 m

Bundle diameter is calculated as follow:

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

K_1 and n_1 are constants. For 1 shell pass their values are,

K_1 = 0.319, n_1 = 2.142

By putting values in the above formula, bundle diameter is found.

$$D_b = 0.02 * \left(\frac{82}{0.319}\right)^{\frac{1}{2.142}} = 267.7 \text{ mm} = 0.2677 \text{ m}$$

Clearance for shell diameter = 0.088 m (from graph)

Shell Diameter = Bundle Diameter + Clearance

Shell Diameter = D_o = 0.2677 + 0.088 = **0.3557 m**

Baffle spacing = l_B = 0.5 * D_o = 0.5 * 0.3557 = 0.1778 m

Number of baffles = $\frac{length\ of\ tube}{Baffle\ Spacing}$ = 3.66 / 0.1778 = 20

Tube Side Co-efficient:

Mean temperature = $\frac{t_1+t_2}{2}$ = 85.75 ^oC

Density of solution = ρ = 888.2 kg/m^3

Viscosity of solution = μ = 0.0002723 kg/m.sec

Heat capacity of solution = C_p = 4468 J/kg ^oC

Thermal conductivity of solution = K = 0.6182 W/m. ^oC

Tube Cross-sectional area = a = $\frac{\pi\, d_i{}^2}{4} =$ (3.14 * 0.016 * 0.016) / 4

a = 201 mm^2 = 0.000201 m^2

Tubes per pass = total tubes / 2 = 82/2 = 41

Total flow area = Tube Cross-sectional area * Tubes per pass

Total flow area = 0.000201 * 41

Total flow area = 0.008 m^2

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{8355/3600}{0.008}$ = 279 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 279 / 888.2 = 0.315 m/sec

Reynolds Number in tubes = $Re_t = \frac{\rho\, d_i\, u_t}{\mu} = \frac{888.2 * 0.016 * 0.315}{0.0002723}$ = 16422.45

From graph between Re_t and J_H factor, it is found that

J_H = 0.0039

Prandtl number = $\frac{C_p \mu}{k}$ = (4468 * 0.0002723) / 0.6182 = 1.968

$$\frac{h_i}{d_i k} = J_H Re\, Pr^{0.33} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

By putting all values in above equation, we get

h_i = 3094.15 W/m^2 ^oC

Shell Side Co-efficient:

Mean temperature = $\frac{T_1+T_2}{2}$ = 135.25 ^oC

Density of solution = ρ = 912.5 kg/m^3

Viscosity of solution = μ = 0.000158kg/m.sec

Heat capacity of solution = C_p = 4511 J/kg °C

Thermal conductivity of solution = K = 0.6851W/m. °C

Cross flow area = A_s = $\frac{(p_t - d_o) D_s\, l_B}{p_t}$ = $\frac{(0.025-0.020)*0.3557*0.178}{0.025}$ = 0.01265 m^2

Equivalent diameter = d_e = $\frac{1.10\,({p_t}^2 - 0.917\,{d_o}^2)}{d_o}$ = 14.2 mm = 0.0142 m

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{7303}{0.01265}$ = 160.34 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 160.34 / 912.5 = 0.175 m/sec

Reynolds Number in tubes = Re_s = $\frac{\rho\, d_e\, u_t}{\mu} = \frac{912.5 * 0.0142 * 0.175}{0.000158}$ = 14411.63

From graph between Re_t and J_H factor, it is found that

J_H = 0.0038

Prandtl number = $\frac{C_p \mu}{k}$ = (4511 * 0.000158) / 0.6851 = 1.0403

$$\frac{h_i}{d_i k} = J_H Re\, Pr^{0.33} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

By putting all values in above equation, we get

h_s = 2676.701 W/m^2 °C

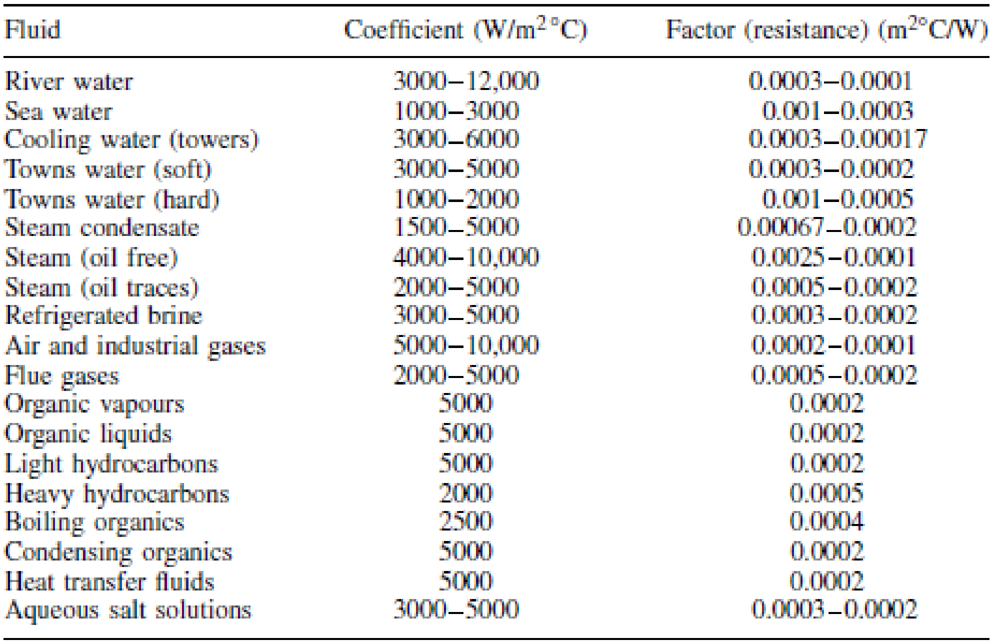
$$\frac{1}{U_o} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2\, k_w} + \frac{d_o}{d_i} * \frac{1}{h_{id}} + \frac{d_o}{d_i} * \frac{1}{h_i}$$

Fouling Factor for shell side = h_{od} = 5000 W/m^2 °C

Fouling Factor for tube side = h_{id} = 5000 W/m^2 °C

K_w = 50 W/m^2 °C

Table 6. 2: Fouling Factor



Fluid	Coefficient (W/m^2 °C)	Factor (resistance) (m^{2}°C/W)
River water	3000–12,000	0.0003–0.0001
Sea water	1000–3000	0.001–0.0003
Cooling water (towers)	3000–6000	0.0003–0.00017
Towns water (soft)	3000–5000	0.0003–0.0002
Towns water (hard)	1000–2000	0.001–0.0005
Steam condensate	1500–5000	0.00067–0.0002
Steam (oil free)	4000–10,000	0.0025–0.0001
Steam (oil traces)	2000–5000	0.0005–0.0002
Refrigerated brine	3000–5000	0.0003–0.0002
Air and industrial gases	5000–10,000	0.0002–0.0001
Flue gases	2000–5000	0.0005–0.0002
Organic vapours	5000	0.0002
Organic liquids	5000	0.0002
Light hydrocarbons	5000	0.0002
Heavy hydrocarbons	2000	0.0005
Boiling organics	2500	0.0004
Condensing organics	5000	0.0002
Heat transfer fluids	5000	0.0002
Aqueous salt solutions	3000–5000	0.0003–0.0002

Put all the values in above equation to get overall heat transfer coefficient.

U = 786 W/m^2 ^{o}C

The value of overall coefficient is almost equal to the assumed value so all assumptions regarding design are true.

Pressure drop:

Tube side pressure drop:

Reynolds number = 16422.5

From graph between Re and J_f,

Friction factor = J_f = 0.0039

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

Pressure drop is calculated by putting values of various factors in above equation:

$$\Delta P_t = 2\left[8 * 0.0039\left(\frac{3.66}{0.016}\right) + 2.5\right]\frac{888.2 * 0.315^2}{2}$$

ΔP_t = 847.54 N/m^2 = 0.847 KPa

Shell side pressure drop:

Reynolds number = 14411.6

From graph between Re and J_f,

Friction factor = J_f = 0.043

$$\Delta P_s = \left[8\, j_f\left(\frac{D_s}{d_e}\right)\left(\frac{L}{l_B}\right)\left(\frac{\mu}{\mu_w}\right)^{-0.14}\right]\frac{\rho {u_t}^2}{2}$$

$$\Delta P_s = \left[8 * 0.043\left(\frac{0.3557}{0.0142}\right)\left(\frac{3.66}{0.178}\right)\right]\frac{912.5 * 0.175^2}{2}$$

ΔP_s = 2498 N/m^2 = 2.498 KPa

Table 6. 3: Summary of Heat exchanger 1

Inner Diameter of tube	16 mm (0.16m)
Outer diameter of tube	20 mm (0.2 m)
Tube length	3.66 m
Number of tubes	82
Shell diameter	0.3557 m
Pitch	0.025 m
Number of baffles	20
Baffle spacing	0.178 m
Shell side coefficient	2676.7 W/m^2. ^{0}C
Tube side coefficient	3094.15 W/m^2. ^{0}C

Over all heat transfer coefficient	786 $W/m^2. {}^0C$
Shell side pressure drop	2.47 kPa
Tube side pressure drop	0.847 kPa
Effectiveness	0.6

6.2 Heat Exchanger 2

Exchanger type: Shell and tube Heat Exchanger

Shell Passes = 1

Tube Passes = N_p = 2

Shell Side Fluid: Lean Solution

Tube Side Fluid: Rich Solution

Lean solution inlet temperature = T_1 = 229 ^oC

Lean solution outlet temperature = T_2 = 166.5 ^oC

Rich solution inlet temperature = t_1 = 111.5 ^oC

Rich solution outlet temperature = t_2 = 163 ^oC

C_p of Lean Solution = 79 KJ/kgmol. ^oC

C_p of Rich Solution = 75 KJ/kgmol. ^oC

Molar flow rate of lean solution = 405.8825 kgmol/hr

Molar flow rate of rich solution = 467.4173 kgmol/hr

$Q_{shell} = m\ C_p\ \Delta T$

= 405.8825 * 79 * (104-166.5)

= - 1902574 KJ/hr = - 1902574 / 3600 = - 528.5 KW (KJ/sec)

Flowrate of shell side fluid = 7303 kg/hr

$Q = UA\Delta T_m$

$\Delta T_m = F_t * LMTD$

$$LMTD = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \frac{(T_1 - t_2)}{(T_2 - t_1)}}$$

Putting values of inlet and outlet temperatures of shell and tube side fluids,

LMTD = 60.33 ^oC

$$R = \frac{(T_1 - T_2)}{(t_2 - t_1)} = 1.21$$

$$S = \frac{(t_2 - t_1)}{(T_1 - t_1)} = 0.44$$

From Figure 6.1 for 1 shell pass and 2 tube passes using values of S and R, we get

$F_t = 0.829$

$\Delta T_m = F_t * LMTD$

$\Delta T_m = 0.829 * 60.33 = 50.02\ ^oC$

Overall heat transfer coefficient for this system is assumed.

U = 800 $W/m^2.\ ^oC$

Heat transfer Area = $A = \frac{Q}{U\Delta \text{Tm}} = \frac{528.5}{800*50.02} = 13.20\ m^2$

Assumptions:

Choose 20 mm outer diameter, 16 mm inner diameter and 3.66 m long tubes

Inner diameter of tube = d_i = 0.016 m

Outer Diameter of tube = d_o = 0.02 m

Length of tube = L = 3.66 m

Area of one tube = $\pi * L * d = \pi * 3.66 * 0.02 = 0.23\ m^2$

Number of tubes = $N_t = \frac{Heat\ transfer\ area}{Area\ of\ one\ tube}$ = 13.2 / 0.23 = 58 tubes

Use triangular pitch,

Pitch = $p_t = 1.25 * d_o = 1.25 * 0.02 = 0.025$ m

Bundle diameter is calculated as follow:

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

K_1 and n_1 are constants. For 1 shell pass their values are,

$K_1 = 0.319$, $n_1 = 2.142$

By putting values in the above formula, bundle diameter is found.

$$D_b = 0.02 * \left(\frac{58}{0.319}\right)^{\frac{1}{2.142}} = 228 \text{ mm} = 0.228 \text{ m}$$

Clearance for shell diameter = 0.087 m (from graph)

Shell Diameter = Bundle Diameter + Clearance

Shell Diameter = D_o = 0.228 + 0.087 = **0.315 m**

Baffle spacing = $l_B = 0.5 * D_o = 0.5 * 0.315 = 0.1575$ m

Number of baffles = $\frac{length\ of\ tube}{Baffle\ Spacing}$ = 3.66 / 0.1573 = 23

Tube Side Co-efficient:

Mean temperature = $\frac{t_1+t_2}{2}$ = 137.25 ^oC

Density of solution = $\rho = 834.2\ kg/m^3$

Viscosity of solution = μ = 0.000118 kg/m.sec

Heat capacity of solution = C_p = 4683 J/kg ^oC

Thermal conductivity of solution = K = 0.6062 W/m. ^oC

Tube Cross-sectional area = a = $\frac{\pi {d_i}^2}{4} =$ (3.14 * 0.016 * 0.016) / 4

a = 201 mm^2 = 0.000201 m^2

Tubes per pass = total tubes / 2 = 58/2 = 29

Total flow area = Tube Cross-sectional area * Tubes per pass

Total flow area = 0.000201 * 29

Total flow area = 0.006 m^2

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{8355/3600}{0.006}$

Mass velocity = 395 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 395 / 834.2 = 0.474 m/sec

Reynolds Number in tubes = Re_t = $\frac{\rho\, d_i\, u_t}{\mu} = \frac{834.2 * 0.016 * 0.474}{0.000118}$ = 53514

From graph between Re_t and J_H factor, it is found that

J_H = 0.003

Prandtl number = $\frac{C_p \mu}{k}$ = (4683 * 0.000118) / 0.6062 = 0.913

$$\frac{h_i}{d_i k} = J_H Re\, Pr^{0.33} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

By putting all values in above equation, we get

h_i = 5902 W/m^2 ^oC

Shell Side Co-efficient:

Mean temperature = $\frac{T_1+T_2}{2}$ = 197.75 ^oC

Density of solution = ρ = 850.5 kg/m^3

Viscosity of solution = µ = 0.00009042 kg/m.sec

Heat capacity of solution = C_p = 4881 J/kg ^oC

Thermal conductivity of solution = K = 0.6618 W/m. ^oC

Cross flow area = A_s = $\frac{(p_t - d_o)D_S\, l_B}{p_t}$ = $\frac{(0.025-0.020)*0.315*0.1573}{0.025}$ = 0.0099 m^2

Equivalent diameter = d_e = $\frac{1.10\,({p_t}^2-0.917\,{d_o}^2)}{d_o}$ = 14.2 mm = 0.0142 m

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}}$ = $\frac{7303}{0.0099}$ = 204.86 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 204.86 / 850.5 = 0.241 m/sec

Reynolds Number in tubes = Re_s = $\frac{\rho\, d_e\, u_t}{\mu}$ = $\frac{850.5 * 0.0142 * 0.241}{0.00009042}$ = 32174.48

From graph between Re_t and J_H factor, it is found that

J_H = 0.0032

Prandtl number = $\frac{C_p \mu}{k}$ = (4881 * 0.00009042) / 0.6618 = 0.6668

$$\frac{h_i}{d_i k} = J_H Re\, Pr^{0.33} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

By putting all values in above equation, we get

h_s = 4197.636 W/m^2 ^oC

$$\frac{1}{U_o} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2\,k_w} + \frac{d_o}{d_i} * \frac{1}{h_{id}} + \frac{d_o}{d_i} * \frac{1}{h_i}$$

Fouling Factor for shell side = h_{od} = 3000 W/m^2 ^oC

Fouling Factor for tube side = h_{id} = 3000 W/m^2 ^oC

K_w = 50 W/m^2 ^oC

Put all the values in above equation to get overall heat transfer coefficient.

U = 796.3 W/m^2 ^{o}C

The value of overall coefficient is almost equal to the assumed value so all assumptions regarding design are true.

Pressure drop:

Tube side pressure drop:

Reynolds number = 53514

From graph between Re and J_f,

Friction factor = J_f = 0.0031

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

Pressure drop is calculated by putting values of various factors in above equation:

$$\Delta P_t = 2 \left[8 * 0.0031 \left(\frac{3.66}{0.016}\right) + 2.5\right]\frac{834.2 * 0.474^2}{2}$$

ΔP_t = 1531 N/m^2 = 1.531 KPa

Shell side pressure drop:

Reynolds number = 32174.48

From graph between Re and J_f,

Friction factor = J_f = 0.041

$$\Delta P_s = \left[8\, j_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{l_B}\right)\left(\frac{\mu}{\mu_w}\right)^{-0.14}\right]\frac{\rho {u_t}^2}{2}$$

$$\Delta P_s = \left[8 * 0.041 \left(\frac{0.315}{0.0142}\right)\left(\frac{3.66}{0.1575}\right)\right]\frac{850.2 * 0.24^2}{2}$$

ΔP_s = 4171 N/m^2 = 4.171 KPa

Table 6. 4: Summary of Heat Exchanger 2

Inner Diameter of tube	16 mm (0.16m)
Outer diameter of tube	20 mm (0.2 m)
Tube length	3.66 m
Number of tubes	58
Shell diameter	0.315 m
Pitch	0.025 m
Number of baffles	23
Baffle spacing	0.157 m
Shell side coefficient	4197.6 W/m^2. ^{0}C
Tube side coefficient	5902.8 W/m^2. ^{0}C
Over all heat transfer coefficient	803 W/m^2. ^{0}C
Shell side pressure drop	4.17 kPa
Tube side pressure drop	1.53 kPa
Effectiveness	0.6

6.3 Cooler for Lean Solution

Exchanger type: Shell and tube Heat Exchanger

Shell Passes = 1

Tube Passes = N_p = 2

Shell Side Fluid: Lean Solution

Tube Side Fluid: Cooling Water

Lean solution inlet temperature = T_1 = 104 ^oC

Lean solution outlet temperature = T_2 = 38 ^oC

Cooling Water inlet temperature = t_1 = 30 ^oC

Cooling Water outlet temperature = t_2 = 60 ^oC

C_p of Lean Solution = 79 KJ/kgmol. ^oC

C_p of Cooling Water = 77.5 KJ/kgmol. ^oC

Molar flow rate of lean solution = 405.8825 kgmol/hr

Molar flow rate of Cooling Water = 910 kgmol/hr

$Q_{shell} = m\, C_p\, \Delta T$

= 405.8825 * 79 * (38-104)

= - 2116271 KJ/hr = - 2116271 / 3600 = - 587.85 KW (KJ/sec)

Flowrate of shell side fluid = 7303 kg/hr

$Q=UA\Delta T_m$

$\Delta T_m = F_t$ * LMTD

$$LMTD = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \frac{(T_1 - t_2)}{(T_2 - t_1)}}$$

Putting values of inlet and outlet temperatures of shell and tube side fluids,

LMTD = 21.12 ^oC

$$R = \frac{(T_1 - T_2)}{(t_2 - t_1)} = 2.2$$

$$S = \frac{(t_2 - t_1)}{(T_1 - t_1)} = 0.41$$

From graph for 1 shell pass and 2 tube passes using values of S and R, we get

$F_t = 0.82$

$\Delta T_m = F_t * LMTD = 0.82 * 21.12 = 17.32\ ^oC$

Overall heat transfer coefficient for this system is assumed.

$U = 1200\ W/m^2.\ ^oC$

Heat transfer Area = $A = \frac{Q}{U\Delta Tm} = \frac{587.85}{1200*17.32} = 28.29\ m^2$

<u>Assumptions:</u>

Choose 20 mm outer diameter, 16 mm inner diameter and 4.88 m long tubes

Inner diameter of tube = $d_i = 0.016$ m

Outer Diameter of tube = $d_o = 0.02$ m

Length of tube = L = 4.88 m

Area of one tube = π * L * d

= π * 4.88 * 0.02 = 0.31 m^2

Number of tubes = $N_t = \frac{Heat\ transfer\ area}{Area\ of\ one\ tube}$ = 28.29 / 0.31 = 92 tubes

Use triangular pitch,

Pitch = $p_t = 1.25 * d_o = 1.25 * 0.02 = 0.025$ m

Bundle diameter is calculated as follow:

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

K_1 and n_1 are constants. For 1 shell pass their values are,

$K_1 = 0.319$, $n_1 = 2.142$

By putting values in the above formula, bundle diameter is found.

$$D_b = 0.02 * \left(\frac{92}{0.319}\right)^{\frac{1}{2.142}} = 283 \text{ mm} = 0.283 \text{ m}$$

Clearance for shell diameter = 0.088 m (from graph)

Shell Diameter = Bundle Diameter + Clearance

Shell Diameter = D_o = 0.283 + 0.088 = **0.371 m**

Baffle spacing = $l_B = 0.5 * D_o = 0.5 * 0.371 = 0.186$ m

Number of baffles = $\frac{length\ of\ tube}{Baffle\ Spacing}$ = 4.88 / 0.186 = 26

Tube Side Co-efficient:

Mean temperature = $\frac{t_1+t_2}{2}$ = 45 ^{o}C

Density of Cooling Water = ρ = 992.1 kg/m^3

Viscosity of Cooling Water = μ = 0.0005939 kg/m.sec

Heat capacity of Cooling Water = C_p = 4318 J/kg ^{o}C

Thermal conductivity of Cooling Water = K = 0.6376 W/m. ^{o}C

Tube Cross-sectional area = a = $\frac{\pi\, d_i^{\,2}}{4} =$ (3.14 * 0.016 * 0.016) / 4

a = 201 mm^2 = 0.000201 m^2

Tubes per pass = total tubes / 2 = 92/2 = 46

Total flow area = Tube Cross-sectional area * Tubes per pass

Total flow area = 0.000201 * 46 = 0.009 m^2

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{16384/3600}{0.009}$ = 484 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 484 / 992.1 = 0.488 m/sec

Reynolds Number in tubes = $Re_t = \frac{\rho\, d_i\, u_t}{\mu} = \frac{992.1 * 0.016 * 0.488}{0.0005939}$ = 13043.43

From graph between Re_t and J_H factor, it is found that

$J_H = 0.004$

Prandtl number = $\frac{C_p \mu}{k}$ = (4318 * 0.0005939) / 0.6376 = 4.02

$$\frac{h_i}{d_i k} = J_H Re\, Pr^{0.33} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

By putting all values in above equation, we get

h_i = 3291 W/m^2 °C

Shell Side Co-efficient:

Mean temperature = $\frac{T_1+T_2}{2}$ = 71 °C

Density of solution = ρ = 896.7 kg/m^3

Viscosity of solution = μ = 0.00033 kg/m.sec

Heat capacity of solution = C_p = 4458 J/kg °C

Thermal conductivity of solution = K = 0.6161 W/m. °C

Cross flow area = $A_s = \frac{(p_t - d_o) D_s\, l_B}{p_t} = \frac{(0.025-0.020)*0.371*0.186}{0.025}$ = 0.0137 m^2

Equivalent diameter = $d_e = \frac{1.10\,({p_t}^2 - 0.917\,{d_o}^2)}{d_o}$ = 14.2 mm = 0.0142 m

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{7303}{0.0137}$ = 147.15 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 147.15 / 896.7 = 0.164 m/sec

Reynolds Number in tubes = $Re_s = \frac{\rho\, d_e\, u_t}{\mu} = \frac{896.7 * 0.0142 * 0.164}{0.00033}$ = 6326.73

From graph between Re_t and J_H factor, it is found that

$J_H = 0.0065$

Prandtl number = $\frac{C_p \mu}{k}$ = (4458 * 0.00033) / 0.6161 = 2.39

$$\frac{h_i}{d_i k} = J_H Re\, Pr^{0.33} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

By putting all values in above equation, we get

$h_s = 2378.4$ W/m^2 ^{o}C

$$\frac{1}{U_o} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2\, k_w} + \frac{d_o}{d_i} * \frac{1}{h_{id}} + \frac{d_o}{d_i} * \frac{1}{h_i}$$

Fouling Factor for shell side = h_{od} = 8000 W/m^2 ^{o}C

Fouling Factor for tube side = h_{id} = 15000 W/m^2 ^{o}C

K_w = 50 W/m^2 ^{o}C

Put all the values in above equation to get overall heat transfer coefficient.

U = 950 W/m^2 ^{o}C

The value of overall coefficient is almost equal to the assumed value so all assumptions regarding design are true.

Pressure drop:

Tube side pressure drop:

Reynolds number = 13043.43

From graph between Re and J_f,

Friction factor = J_f = 0.0045

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

Pressure drop is calculated by putting values of various factors in above equation:

$$\Delta P_t = 2 \left[8 * 0.0045 \left(\frac{4.88}{0.016}\right) + 2.5\right]\frac{992.1 * 0.488^2}{2}$$

ΔP_t = 3184 N/m^2 = 3.184 KPa

Shell side pressure drop:

Reynolds number = 6323.8

From graph between Re and J_f,

Friction factor = J_f = 0.052

$$\Delta P_s = \left[8\, j_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{l_B}\right)\left(\frac{\mu}{\mu_w}\right)^{-0.14}\right]\frac{\rho {u_t}^2}{2}$$

$$\Delta P_s = \left[8 * 0.052 \left(\frac{0.371}{0.0142}\right)\left(\frac{4.88}{0.186}\right)\right]\frac{896.7 * 0.164^2}{2}$$

ΔP_s = 3452 N/m^2 = 3.452 KPa

Table 6. 5: Summary of Lean Solution Cooler

Inner Diameter of tube	16 mm (0.16m)
Outer diameter of tube	20 mm (0.2 m)
Tube length	4.88 m
Number of tubes	92
Shell diameter	0.371 m
Pitch	0.025 m

Number of baffles	26
Baffle spacing	0.186 m
Shell side coefficient	2378.4 W/m^2. ^{0}C
Tube side coefficient	3291 W/m^2. ^{0}C
Over all heat transfer coefficient	950 W/m^2. ^{0}C
Shell side pressure drop	3.452 kPa
Tube side pressure drop	3.184 kPa
Effectiveness	0.8

6.4. Distillation Column

6.4.1 No of Plates

Number of equilibrium stages or no plates required to achieve desired separation are calculated as follows.

Firstly, Fenske's Equation is used to calculate minimum number of plates.

$$N_{min} = \frac{\ln[\, X_d(1-X_b)/\, X_b(1-X_d)\ ]}{\ln(\, \sigma_{ab})} \qquad \text{(Equation 6.1)}$$

Where

X_d (Concentration of more volatile component at top) = 0.995 = 99.5%

X_B (Concentration of more volatile component at bottom) = 0.01 = 1%

X_f (Concentration of more volatile component in feed) = 0.14032 = 14.032%

Y_f (Concentration of more volatile component in feed) = 0.85968 = 85.968

Relative Volatility = $\alpha_{AB} = \frac{(Y_{ae}/X_{ae})}{(Y_{be}/X_{be})}$

Where

Y_{ae} = Equilibrium Concentration of Component A in vapor phase = 0.999

Y_{be} = Equilibrium Concentration of Component B in vapor phase = 0.001

X_{ae} = Equilibrium Concentration of Component A in liquid phase = 0.995

X_{be} = Equilibrium Concentration of Component B in liquid phase = 0.005

Putting above values in equation 1 gives

N_{min} = Minimun No of Plates = 5.128746 $\simeq$ 6

For calculating actual/theoretical no of plates, Gilliand equation is used

$$\frac{N-N_{min}}{N+1} = 0.75\left[1-\left(\frac{R-R_{min}}{R+1}\right)^{0.566}\right] \qquad \text{(Equation 6.2)}$$

Where

N_{min} = Minimun No of Plates

N = Actual/Theoretical No of Plates

R_{min} = Minimum Reflux Ratio = $\frac{X_d - Y_f}{Y_f - X_f}$ = 1.8

R = Actual Reflux Ratio = $\frac{L}{D} = \frac{185.53}{61.84}$ = 3

Equation 6.2, on putting above values, gives

N/N_{th} = Actual No of Plates = 9.44 $\simeq$ 10

Effective number of plates or stages that are installed for operation are calculated as:

$$N_{eff} = N_{th} \cdot \left(\frac{S}{Eov}\right) \qquad \text{(Equation 6.3)}$$

Where

N_{eff} = Effective No of Stages/ No of stages to be actually installed

N_{th} = Theoretical No of Stages = 10

S = Safety Co-efficient = 1.4

E_{ov} = Overall Efficiency = 70% = 0.7

Substituting above values in equation 3 gives

Neff = No of stages/plates to be actually installed = 20

6.4.2. Height of Column

Using the effective no of stages (N_{eff}), Height of column can be calculated easily.

$H = N_{eff} . \Delta Z$ (Equation 6.4)

H = Height of column

N_{eff} = Effective No of Stages/ No of stages to be actually installed = 20

ΔZ = Spacing between the plates = 0.55m

Equation (4) on substituting the values, gives

H = Height of Column = 11 m

$H_{tot} = H + H_{min}$

H_{tot} = Actual Height of column N_{gmax}

H_{min} = Minimun Height (factor for calculating actual height) = 3. ΔZ = 1.65 m

So

H_{tot} = Actual Height of Column = 12.65 m

6.4.3. Diameter of Column

Internal Diameter of column is given by the following equation

$$D_i = 102.89 \sqrt{\frac{N_{Gmax} \, . \, T}{P \, . \, W_{Gzul}}}$$

N_{gmax} = Maximum vapor molar flow rate (kmol/s) = 247.38 kmol/hr (Equation 6.5)

$W_{g.zul}$ = Allowable vapor velocity referred to area = 0.909 m/s (Given by Equation 6.6)

T (K) = Column Operating Temperature = 493 K

P (Pa) = Column Operating Pressure = 3000000 Pa

For allowable vapor velocity, following equation is used

$$W_{G\ zul} = 0.7 * \frac{Aac}{AQ} * Kv * \sqrt{\frac{\rho L - \rho g}{\rho g}} \qquad \text{(Equation 6.6)}$$

Where

A_{ac} = active area of a tray [m²] = 4 m^2

A_Q = total column cross-sectional area [m²] = 5.51 m^2

K_v = capacity parameter of Souders/Brown [m/s] = 0.45 m/s

ρ_L = Liquid Density (kg/m^3) = 815 Kg/m^3

ρ_g = Vapor Density (kg/m^3) = 25.1 Kg/m^3

Equation (5) on substituting the values, gives

$W_{G\ zul}$ = 0.909 m/s

Substituting the above value and all other values in equation (6.5) gives,

D_i = Internal Diameter of Column = 1.19 m

6.5 Reboiler

Reboiler Type: kettle type

Exchanger Duty = 2265352 KJ/hr

Exchanger Duty = 629.264 kW

Maximum Flux is taken from the graph for aqueous solution = 22000 W/m^2

Heat Transfer area required = Exchanger duty/Maximum Flux = 629264/22000 = 28.6m^2

Assumptions:

Choose 25 mm outer diameter, 20 mm inner diameter and 3.66 m long tubes

Inner diameter of tube = d_i = 0.02 m

Outer Diameter of tube = d_o= 0.025 m

Length of tube = L = 3.66 m

Area of one tube = π * L * d = π * 3.66 * 0.025 = 0.287 m^2

Number of tubes = N_t = $\frac{Heat\ transfer\ area}{Area\ of\ one\ tube}$ = 28.6 / 0.287 = 100 tubes

Use triangular pitch,

Pitch = p_t = 1.25 * d_o = 1.25 * 0.025 = 0.03125 m

Bundle diameter is calculated as follow:

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

K_1 and n_1 are constants. For 1 shell pass their values are,

K_1 = 0.319, n_1 = 2.142

By putting values in the above formula, bundle diameter is found.

$D_b = 0.025 * \left(\frac{100}{0.319}\right)^{\frac{1}{2.142}}$ = 365 mm = 0.365 m

Clearance for shell diameter = 0.090 m (from graph)

Shell Diameter = Bundle Diameter + Clearance

Shell Diameter = D_o = 0.365 + 0.090 = **0.455 m**

Baffle spacing = l_B = 0.5 * D_o = 0.5 * 0.455 = 0.2275 m

Number of baffles = $\frac{length\ of\ tube}{Baffle\ Spacing}$ = 3.66 / 0.2275 = 16

Viscosity of liquid = μ_L = 0.00008459 Ns/m^2

Density of the vapor = ρ_v = 13.24 Kg/m^3

Conductivity of liquid = K_L = 0.5838 W/m.^{o}C

Density of the Liquid = ρ_L = 751.3 Kg/m^3

Cp of liquid = 5460 J/kg

Surface tension = σ = 0.028 N/m

Heat of vaporization = λ = 1775000 J/kg

Wall Surface temperature = T_w = 248 ^{o}C

Saturation temperature of boiling liquid = T_s = 229 ^{o}C

Saturation pressure corresponding to the wall temperature = Pw = 3844900 Pa

Saturation pressure corresponding to Ts = Ps = 2700000 Pa

$$h_{nb} = 0.00122 \left[\frac{k_L^{0.79} C_{pL}^{0.45} \rho_L^{0.49}}{\sigma^{0.5} \mu_L^{0.29} \lambda^{0.24} \rho_v^{0.24}} \right] (T_w - T_s)^{0.24} (p_w - p_s)^{0.75}$$

By putting values of all variables in above equation, we get

h_{nb} = 107648.5 W/m^2. ^{o}C

Taking the water coefficient as 10000 W/m^2. ^{o}C,

Fouling Factor for shell side = h_{od} = 3500 W/m^2 ^{o}C

Fouling Factor for tube side = h_{id} = 6000 W/m^2 ^{o}C

$$\frac{1}{U_o} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2\,k_w} + \frac{d_o}{d_i} * \frac{1}{h_{id}} + \frac{d_o}{d_i} * \frac{1}{h_i}$$

K_w = 50 W/m^2 ^{o}C

Put all the values in above equation to get overall heat transfer coefficient.

U = 1461.7 W/m^2 ^{o}C

Density of Steam = ρ = 18.34 kg/m^3

Viscosity of Steam = μ = 0.000001739 kg/m.sec

Tube Cross-sectional area = a = $\frac{\pi \, d_i^{\,2}}{4}$ = (3.14 * 0.020 * 0.020) / 4 = 0.000314 m^2

Tubes per pass = total tubes / 2 = 100/2 = 50

Total flow area = Tube Cross-sectional area * Tubes per pass

Total flow area = 0.000314 * 50 = 0.0312 m^2

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{0.3576}{0.0312}$ = 11.44 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 11.44 / 18.34 = 0.624 m/sec

Reynolds Number in tubes = Re_t = $\frac{\rho \, d_i \, u_t}{\mu} = \frac{11.44 * 0.020 * 0.624}{0.000001739}$ = 131564.4

For Shell,

Cross flow area = A_s = $\frac{(p_t - d_o) D_s \, l_B}{p_t} = \frac{(0.03125-0.025)*0.455*0.2275}{0.03125}$ = 0.0207 m^2

Equivalent diameter = d_e = $\frac{1.10 \, (p_t^{\,2} - 0.917 \, d_o^{\,2})}{d_o}$ = 24.6 mm = 0.0246 m

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{2.98}{0.0207}$ = 143.88 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 143.88 / 751.3 = 0.1915 m/sec

Reynolds Number in tubes = Re_s = $\frac{\rho \, d_e \, u_t}{\mu} = \frac{751.3 * 0.0246 * 0.1915}{0.0000846}$ = 41990.38

<u>Pressure drop:</u>

Tube side pressure drop:

Reynolds number = 131564.4

From graph between Re and J_f,

Friction factor = J_f = 0.0027

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

Pressure drop is calculated by putting values of various factors in above equation:

$$\Delta P_t = 2 \left[8 * 0.0027 \left(\frac{3.66}{0.020}\right) + 2.5\right]\frac{11.44 * 0.624^2}{2}$$

ΔP_t = 46.04 N/m^2 = 0.046 KPa

Shell side pressure drop:

Reynolds number = 41990.38

From graph between Re and J_f,

Friction factor = J_f = 0.042

$$\Delta P_s = \left[8\, j_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{l_B}\right)\left(\frac{\mu}{\mu_w}\right)^{-0.14}\right]\frac{\rho {u_t}^2}{2}$$

$$\Delta P_s = \left[8 * 0.042 \left(\frac{0.455}{0.0246}\right)\left(\frac{3.66}{0.2275}\right)\right]\frac{751.3 * 0.1915^2}{2}$$

ΔP_s = 1372.8 N/m^2 = 1.3728 KPa

6.6 Condenser

Condenser Type: Horizontal Exchanger, condensation in the shell

The vapors enter the condenser at 61 ^oC and condensate Leaves at 40 ^oC

The Cooling Water available at 25 ^oC and the temperature rise is limited to 30 ^oC.

Shell Side Fluid: Ammonia

Tube Side Fluid: Cooling Water

Ammonia inlet temperature = T_1 = 61 ^oC

Ammonia outlet temperature = T_2 = 40 ^oC

Cooling Water inlet temperature = t_1 = 25 ^oC

Cooling Water outlet temperature = t_2 = 55 ^oC

C_p of Cooling Water = 77.5 KJ/kgmol. ^oC

Flow rate of lean solution = 4203 kg/hr

Flow rate of Cooling Water = 32959.7 kg/hr

Q_c = 1186 kW

$Q=UA\Delta T_m$

$\Delta T_m = F_t * LMTD$

$$LMTD = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln\frac{(T_1 - t_2)}{(T_2 - t_1)}}$$

Putting values of inlet and outlet temperatures of shell and tube side fluids,

LMTD = 9.82 ^oC

$$R = \frac{(T_1 - T_2)}{(t_2 - t_1)} = 0.7$$

$$S = \frac{(t_2 - t_1)}{(T_1 - t_1)} = 0.83$$

From graph for 1 shell pass and 2 tube passes using values of S and R, we get

F_t = 0.72

$\Delta T_m = F_t * LMTD$

$\Delta T_m = 0.72 * 9.82$

$\Delta T_m = 7.07\ ^oC$

Overall heat transfer coefficient for this system is assumed.

$U = 1200\ W/m^2.\ ^oC$

Heat transfer Area = $A = \frac{Q}{U\Delta Tm} = \frac{1186}{1200*7.07} = 139.7\ m^2$

Assumptions:

Choose 25 mm outer diameter, 20 mm inner diameter and 6.1 m long tubes

Inner diameter of tube = d_i = 0.02 m

Outer Diameter of tube = d_o = 0.025 m

Length of tube = L = 6.1 m

Area of one tube = π * L * d

= π * 6.1 * 0.025 = 0.478 m^2

Number of tubes = N_t = $\frac{Heat\ transfer\ area}{Area\ of\ one\ tube}$ = 139.7 / 0.478 = 292 tubes

Use triangular pitch,

Pitch = $p_t = 1.25 * d_o = 1.25 * 0.025 = 0.03125$ m

Bundle diameter is calculated as follow:

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

K_1 and n_1 are constants. For 1 shell pass their values are,

$K_1 = 0.319$, $n_1 = 2.142$

By putting values in the above formula, bundle diameter is found.

$$D_b = 0.025 * \left(\frac{292}{0.319}\right)^{\frac{1}{2.142}} = 603\ mm = 0.603\ m$$

Clearance for shell diameter = 0.092 m (from graph)

Shell Diameter = Bundle Diameter + Clearance

Shell Diameter = D_s = 0.603 + 0.092 = **0.695 m**

Baffle spacing = l_B = 0.5 * D_o = 0.5 * 0.695 = 0.347 m

Number of baffles = $\frac{length\ of\ tube}{Baffle\ Spacing}$ = 6.1 / 0.347 = 18

Shell Side Coefficient:

Assuming the cofficient of condensation = 1300 W/m^2 ^{o}C

Mean Temperature:

Shell Side = (61+40)/2 = 50.5 ^{o}C

Tube Side = (25+55)/2 = 40 ^{o}C

Estimation of tube wall Temperature Tw;

(50.5 – T_w)*1300 = (50.5 - 40) *1200

T_w = 40.8 ^{o}C

Mean Temperature of condensate = (50.5 + 40.8) / 2 = 45.6 ^{o}C

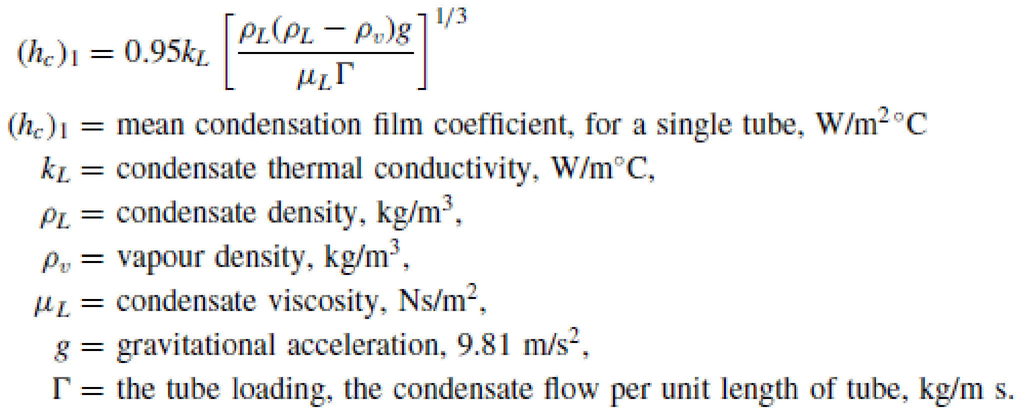
Physical Properties at 45.6 ^{o}C

μ_L = 0.001006 P

ρ_L = 570.3 Kg/m^3

K_L = 0.4322 W/m K

Vapour Density at the mean vapour Temperature = 4.471 kg/m^3



$$(h_c)_1 = 0.95 k_L \left[\frac{\rho_L(\rho_L - \rho_v) g}{\mu_L \Gamma}\right]^{1/3}$$

$(h_c)_1$ = mean condensation film coefficient, for a single tube, W/m^2 °C

k_L = condensate thermal conductivity, W/m°C,

ρ_L = condensate density, kg/m^3,

ρ_v = vapour density, kg/m^3,

μ_L = condensate viscosity, Ns/m^2,

g = gravitational acceleration, 9.81 m/s^2,

Γ = the tube loading, the condensate flow per unit length of tube, kg/m s.

By putting values of all variables, value of mean condensation film coefficient is found.

$(h_c)_1$ = 6923 W/m^2 °C

$(h_b)_1 = (h_c)_1 * N_r^{-1/6}$

Where N_r = (2/3) * number of tubes in center row = (2/3)*19 = 12.6

$(h_b)_1 = 6923 * (12.6)^{-1/6}$ = 4522.4 W/m^2 °C

Tube Side Coefficient:

Density of water of 40 °C = 992.1 kg/m^3

Viscosity of water = 0.0006 Pa.s

Tube Cross-sectional area = a = $\frac{\pi\, {d_i}^2}{4}$ = (3.14 * 0.02 * 0.02) / 4 = 0.000314 m^2

Tubes per pass = total tubes / 2 = 292/2 = 146

Total flow area = Tube Cross-sectional area * Tubes per pass

Total flow area = 0.000314 * 146

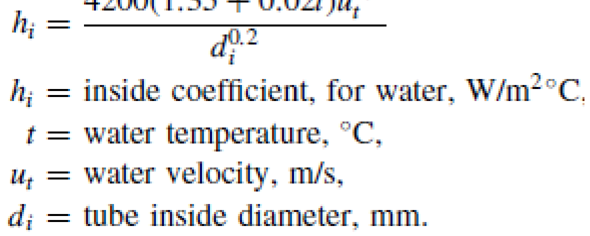
Total flow area = 0.0458 m^2

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{9.155}{0.0458}$ = 200 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 200 / 992.1 = 0.201 m/sec

Reynolds Number in tubes = $Re_t = \frac{\rho\, d_i\, u_t}{\mu} = \frac{992.1 * 0.020 * 0.201}{0.0006}$ = 6660.35

Density of water of 40 °C = 992.1 kg/m^3



$$h_i = \frac{4200(1.35 + 0.02t)u_t^{0.8}}{d_i^{0.2}}$$

h_i = inside coefficient, for water, W/m^2°C,

t = water temperature, °C,

u_t = water velocity, m/s,

d_i = tube inside diameter, mm.

Tube Velocity = $\frac{\text{Flow rate of water}}{\text{Tube cross sectional area} * \text{Density}}$

Tube Velocity = $\frac{9.155}{0.0458*992.1}$ = 0.201 m/s

Put values of t, tube diameter and water velocity to get inside coefficient.

h_i = 5479.4 W/m^2 ^{o}C

Fouling factors: As neither fluid is heavily fouling, use 6000 W/m2C for each side.

K_w = 50 W/m^2 ^{o}C

$$\frac{1}{U_o} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2\,k_w} + \frac{d_o}{d_i} * \frac{1}{h_{id}} + \frac{d_o}{d_i} * \frac{1}{h_i}$$

Overall coefficient = U = 1136.3 W/m^2 ^{o}C

Pressure drop:

Tube side pressure drop:

Reynolds number = 6660.35

From graph between Re and J_f,

Friction factor = J_f = 0.0056

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

Pressure drop is calculated by putting values of various factors in above equation:

$$\Delta P_t = 2 \left[8 * 0.0056 \left(\frac{6.1}{0.020}\right) + 2.5\right]\frac{992.1 * 0.201^2}{2}$$

ΔP_t = 540 N/m^2 = 0.540 KPa

Shell side pressure drop:

Cross flow area = $A_s = \frac{(p_t - d_o) D_s\, l_B}{p_t} = \frac{(0.03125-0.025)*0.695*0.347}{0.03125}$ = 0.048 m^2

Equivalent diameter = $d_e = \frac{1.10\,({p_t}^2 - 0.917\,{d_o}^2)}{d_o}$ = 19.7 mm = 0.0197 m

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{1.1675}{0.048}$ = 24.15 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 24.15 / 4.471 = 5.40 m/sec

Reynolds Number in tubes = $Re_s = \frac{\rho\, d_e\, u_t}{\mu} = \frac{4.471 * 0.0197 * 5.4}{0.00000909}$ =52480

From graph between Re and J_f,

Friction factor = J_f = 0.04

$$\Delta P_s = \left[8\, j_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{l_B}\right)\left(\frac{\mu}{\mu_w}\right)^{-0.14}\right]\frac{\rho {u_t}^2}{2}$$

$$\Delta P_s = \left[8 * 0.04 \left(\frac{0.695}{0.0197}\right)\left(\frac{6.1}{0.347}\right)\right]\frac{4.471 * 5.4^2}{2}$$

ΔP_s = 10320 N/m^2 = 10.320 KPa

6.7 Cooler for Hydrogen Stream

Exchanger type: Shell and tube Heat Exchanger

Shell Passes = 1

Tube Passes = N_p = 2

Shell Side Fluid: Hydrogen Stream

Tube Side Fluid: Cooling Water

Lean solution inlet temperature = T_1 = 170.3 ^oC

Lean solution outlet temperature = T_2 = 100 ^oC

Cooling Water inlet temperature = t_1 = 25 ^oC

Cooling Water outlet temperature = t_2 = 55 ^oC

C_p of Hydrogen Stream = 28.94 KJ/kgmol. ^oC

C_p of Cooling Water = 77.7 KJ/kgmol. ^oC

Molar flow rate of Hydrogen Stream = 512 kgmol/hr

Molar flow rate of Cooling Water = 448 kgmol/hr

$Q_{shell} = m\ C_p\ \Delta T$

$= 512 * 28.95 * (100-170.3)$

$= - 1041655$ KJ/hr $= - 1041655 / 3600 = - 289.3$ KW (KJ/sec)

Flowrate of shell side fluid = 1262 kg/hr

$Q=UA\Delta T_m$

$\Delta T_m = F_t * LMTD$

$$LMTD = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln\frac{(T_1 - t_2)}{(T_2 - t_1)}}$$

Putting values of inlet and outlet temperatures of shell and tube side fluids,

LMTD = 93.71 ^oC

$$R = \frac{(T_1 - T_2)}{(t_2 - t_1)} = 2.34$$

$$S = \frac{(t_2 - t_1)}{(T_1 - t_1)} = 0.21$$

From graph for 1 shell pass and 2 tube passes using values of S and R, we get

$F_t = 0.97$

$\Delta T_m = F_t * LMTD$

$\Delta T_m = 0.97 * 93.71$

$\Delta T_m = 90.9\ ^oC$

Overall heat transfer coefficient for this system is assumed.

$U = 60\ W/m^2.\ ^oC$

Heat transfer Area = $A = \frac{Q}{U\Delta \text{Tm}} = \frac{289.3}{60*90.9} =$ 53.05 m^2

Assumptions:

Choose 20 mm outer diameter, 16 mm inner diameter and 3.66 m long tubes

Inner diameter of tube = d_i = 0.016 m

Outer Diameter of tube = d_o = 0.020 m

Length of tube = L = 3.66 m

Area of one tube = π * L * d

= π * 3.66 * 0.02 = 0.23 m^2

Number of tubes = $N_t = \frac{Heat\ transfer\ area}{Area\ of\ one\ tube}$ = 53.05 / 0.23 = 230 tubes

Use triangular pitch,

Pitch = $p_t = 1.25 * d_o$ = 1.25 * 0.02 = 0.025 m

Bundle diameter is calculated as follow:

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

K_1 and n_1 are constants. For 1 shell pass their values are,

$K_1 = 0.319$, $n_1 = 2.142$

By putting values in the above formula, bundle diameter is found.

$D_b = 0.02 * \left(\frac{230}{0.319}\right)^{\frac{1}{2.142}}$ = 433 mm = 0.433 m

Clearance for shell diameter = 0.090 m (from graph)

Shell Diameter = Bundle Diameter + Clearance

Shell Diameter = D_o = 0.433 + 0.090 = **0.523 m**

Baffle spacing = l_B = 0.5 * D_o = 0.5 * 0.523 = 0.261 m

Number of baffles = $\frac{length\ of\ tube}{Baffle\ Spacing}$ = 3.66 / 0.261 = 14

Tube Side Co-efficient:

Mean temperature = $\frac{t_1+t_2}{2}$ = 40 °C

Density of Cooling Water = ρ = 1000 kg/m^3

Viscosity of Cooling Water = μ = 0.000696 kg/m.sec

Heat capacity of Cooling Water = C_p = 4318 J/kg °C

Thermal conductivity of Cooling Water = K = 0.6376 W/m. °C

Tube Cross-sectional area = a = $\frac{\pi\, {d_i}^2}{4} =$ (3.14 * 0.016 * 0.016) / 4

a = 201 mm^2 = 0.000201 m^2

Tubes per pass = total tubes / 2 = 230/2 = 115

Total flow area = Tube Cross-sectional area * Tubes per pass

Total flow area = 0.000201 * 115

Total flow area = 0.0231 m^2

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{8064/3600}{0.0231}$

Mass velocity = 96 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 96 / 1000 = 0.096 m/sec

Reynolds Number in tubes = $Re_t = \frac{\rho\, d_i\, u_t}{\mu}$

$$Re_t = \frac{1000 * 0.016 * 0.096}{0.000696} = 2212.088$$

From graph between Re_t and J_H factor, it is found that

$$J_H = 0.0029$$

Prandtl number = $\frac{C_p \mu}{k}$ = (4318 * 0.000696) / 0.6376

= 4.7

$$\frac{h_i}{d_i k} = J_H Re\, Pr^{0.33} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

By putting all values in above equation, we get

h_i = 426.3 W/m^2 ^{o}C

Shell Side Co-efficient:

Mean temperature = $\frac{T_1+T_2}{2}$ = 135.15 ^{o}C

Density of Hydrogen Stream = ρ = 2.123 kg/m^3

Viscosity of Hydrogen Stream = μ = 0.0000112 kg/m.sec

Heat capacity of Hydrogen Stream = C_p = 11.65 J/kg ^{o}C

Thermal conductivity of Hydrogen Stream = K = 0.2104 W/m. ^{o}C

Cross flow area = $A_s = \frac{(p_t - d_o) D_s\, l_B}{p_t}$

$$A_s = \frac{(0.025-0.020)*0.523*0.261}{0.025} = 0.0273 \text{ m}^2$$

Equivalent diameter = $d_e = \frac{1.10\,(p_t{}^2 - 0.917\, d_o{}^2)}{d_o}$

d_e = 14.2 mm = 0.0142 m

Mass velocity = $\frac{\text{Mass flowrate}}{\text{Flow Area}} = \frac{1262}{0.0273}$

Mass velocity = 12.8 kg/sec. m^2

Linear velocity = u_t = (mass velocity / density) = 12.8 / 2.123 = 6 m/sec

Reynolds Number in tubes = $Re_s = \frac{\rho\, d_e\, u_t}{\mu}$

$Re_s = \frac{2.123 * 0.0142 * 6}{0.0000112}$ = 16233.053

From graph between Re_t and J_H factor, it is found that

J_H = 0.0028

Prandtl number = $\frac{C_p \mu}{k}$ = (11.65 * 0.0000112) / 0.2104

= 0.0006202

$$\frac{h_i}{d_i k} = J_H Re\, Pr^{0.33} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

By putting all values in above equation, we get

h_s = 58.85 W/m^2 °C

$$\frac{1}{U_o} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2\,k_w} + \frac{d_o}{d_i} * \frac{1}{h_{id}} + \frac{d_o}{d_i} * \frac{1}{h_i}$$

Fouling Factor for shell side = h_{od} = 10,000 W/m^2 °C

Fouling Factor for tube side = h_{id} = 6000 W/m^2 °C

K_w = 50 W/m^2 °C

Put all the values in above equation to get overall heat transfer coefficient.

U = 49.3 W/m² °C

The value of overall coefficient is almost equal to the assumed value so all assumptions regarding design are true.

Pressure drop:

Tube side pressure drop:

Reynolds number = 2212

From graph between Re and J_f,

Friction factor = J_f = 0.007

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

Pressure drop is calculated by putting values of various factors in above equation:

$$\Delta P_t = 2 \left[8 * 0.007 \left(\frac{3.66}{0.016}\right) + 2.5\right]\frac{1000 * 0.096^2}{2}$$

ΔP_t = 141.6 N/m² = 0.141 KPa

Shell side pressure drop:

Reynolds number = 16233

From graph between Re and J_f,

Friction factor = J_f = 0.048

$$\Delta P_s = \left[8\, j_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{l_B}\right)\left(\frac{\mu}{\mu_w}\right)^{-0.14}\right]\frac{\rho {u_t}^2}{2}$$

$$\Delta P_s = \left[8 * 0.048 \left(\frac{0.523}{0.0142}\right)\left(\frac{3.66}{0.261}\right)\right]\frac{2.123 * 6^2}{2}$$

ΔP_s = 7640 N/m² = 7.640 KPa

Table 6. 6: Summary of HRU Cooler

Inner Diameter of tube	16 mm (0.16m)
Outer diameter of tube	20 mm (0.2 m)
Tube length	3.66 m
Number of tubes	230
Shell diameter	0.523 m
Pitch	0.025 m
Number of baffles	14
Baffle spacing	0.261 m
Shell side coefficient	58.85 W/m^2. ^{0}C
Tube side coefficient	426.33 W/m^2. ^{0}C
Over all heat transfer coefficient	49.3 W/m^2. ^{0}C
Shell side pressure drop	7.64 kPa
Tube side pressure drop	0.141 kPa

6.8 Absorber 1

Liquid flowrate = m_L = 323.8 kgmol/hr

Gas flowrate = m_v = 1004 kgmol/hr

Ammonia in gas feed = y_2 = 5.5 %

Ammonia in liquid feed = x_1 = 1 %

Ammonia in exit gas = y_1 = 0.2 %

Ammonia in liquid exit = x_2 = 15 %

M_L = 18 kg/kgmol

M_v = 10.68 kg/kgmol

ρ_v = 10.6 kg/m^3

ρ_L = 55 kg/m^3

m_L = 1.5 m_{Lmin}

Minimum liquid flowrate = 215.9 kgmol/hr

$$\frac{m_{L\,m}}{m_V{}'} = \frac{y_{2k}{}' - y_{1k}{}'}{\frac{y_{2k}{}'}{K_k} - X_{1k}{}'}$$

By putting values in above equation, we get

Equilibrium constant = K_k = 0.214

$$A_A = \frac{m_L}{K_{K*}\,m_v} = 1.504$$

$$(A_A)^{Ne} = \frac{y_{2k}{}' - K_k\,X_{1k}{}'}{y_{1k} - K_k\,X_{1k}{}'} * \left(1 - \frac{1}{A_A}\right) + \frac{1}{A_A}$$

Number of equilibrium stages = N_e = 5.6

Efficiency = 30%

Actual Number of stages = N_e / efficiency = 19

Volumetric flowrate of gas = 139.4 m^3/hr

$$A = \frac{V_v{}'}{V_v}$$

$$A = \frac{\pi\,D^2}{4}$$

$$V_V = K_V\left[\frac{(\rho_L{}' - \rho_V{}')}{\rho_V{}'}\right]^{1/2}$$

$$K_V = 0.75\,k\,(\frac{\sigma}{20})^{0.2}$$

$$K = f\ [\ \frac{m_L{}'\ M_L{}'}{m_V{}'\ M_V{}'}\left(\frac{\rho_V{}'}{\rho_L{}'}\right)^{0.5},\ Z_T\ ]$$

Value of k is taken from Figure 6.1 with the help of tray spacing for high pressure and the flow parameter upon which k depends.

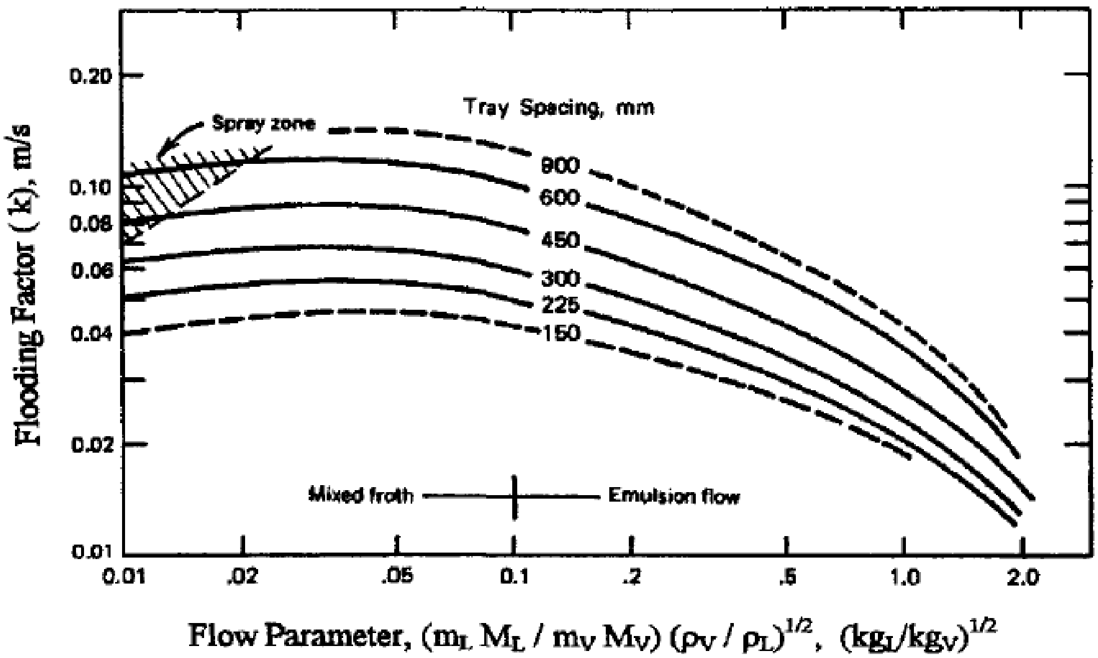


Figure 6.2: Flooding Factor for Absorber 1 (From Coulson and Richardson Volume 6)

Surface tension = σ = 69 dyne/cm

Value of K_v by using surface tension and k = 0.03, is found to be,

K_v = 0.03 m/s

v_v = 0.06 m/s

Now, $A = V_v / v_v = 0.66\ m^2$

From area, **Diameter = D = 0.91 m = 3 ft**

Height of column:

$$Z = N_A \; Z_T + 3\,ft + 0.25\,D + L_S$$

$$L_S = 0.06\,N_A + 2.0$$

L_s = 3.12

Height = Z = 9.96 m = 10 m

6.9 Absorber 2

Liquid flowrate = m_L = 82.162 kgmol/hr

Gas flowrate = m_v = 48 kgmol/hr

Ammonia in gas feed = y_2 = 17.2 %

Ammonia in liquid feed = x_1 = 1 %

Ammonia in exit gas = y_1 = 0.1 %

Ammonia in liquid exit = x_2 = 14 %

M_L = 18 kg/kgmol

M_v = 12.77 kg/kgmol

ρ_v = 12.38 kg/m^3

ρ_L = 55 kg/m^3

m_L = 1.5 m_{Lmin}

Minimum liquid flowrate = 54.77 kgmol/hr

$$\frac{m_{L\,m}}{m_V{}'} = \frac{y_{2k}{}' - y_{1k}{}'}{\frac{y_{2k}{}'}{K_k} - X_{1k}{}'}$$

By putting values in above equation, we get

Equilibrium constant = K_k = 1.076

$$A_A = \frac{m_L}{K_{K*}\,m_v}$$

A_A = 1.59

$$(A_A)^{Ne} = \frac{y_{2k}{}' - K_k\,X_{1k}{}'}{y_{1k} - K_k\,X_{1k}{}'} * \left(1 - \frac{1}{A_A}\right) + \frac{1}{A_A}$$

Number of equilibrium stages = N_e = 9

Efficiency = 80%

Actual Number of stages = N_e / efficiency = 11

Volumetric flowrate of gas = 50 m^3/hr

$$A = \frac{V_v'}{V_v}$$

$$A = \frac{\pi D^2}{4}$$

$$V_V = K_V \left[\frac{(\rho_L' - \rho_V')}{\rho_V'}\right]^{1/2}$$

$$K_V = 0.75\, k\, \left(\frac{\sigma}{20}\right)^{0.2}$$

$$K = f\ \left[\frac{m_L'\ M_L'}{m_V'\ M_V'} \left(\frac{\rho_V'}{\rho_L'}\right)^{0.5},\ Z_T\right]$$

Value of k is taken from Figure 6.2 with the help of tray spacing for high pressure and the flow parameter upon which k depends.

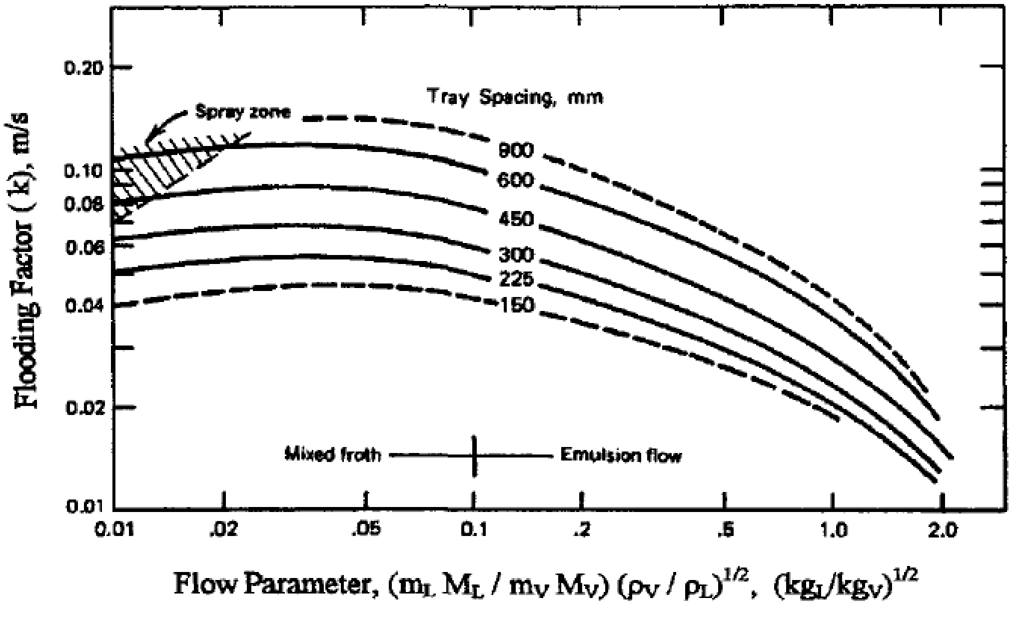


Figure 6.3: Flooding Factor for Absorber 2 (From Coulson and Richardson Volume 6)

Surface tension = σ = 69 dyne/cm

Value of K_v by using surface tension and k = 0.05, is found to be,

K_v = 0.05 m/s, v_v = 0.08 m/s

Now, $A = V_v / v_v = 0.17\ m^2$

From area, **Diameter = D = 0.47 m = 1.53 ft**

Height of column:

$$Z = N_A\ Z_T + 3\,ft + 0.25\,D + L_S$$

$$L_S = 0.06\,N_A + 2.0 = 2.67$$

Height = Z = 5.63 m

6.10 Compressor for HRU

Flowrate of gas = Q = 512 kgmol/hr

R = 8.314 J/mol.K

Inlet temperature = T = 4 ^oC = 277 K

$\eta = 0.85$

$\gamma = 1.4$

$\psi = P/P_1$

Inlet pressure = P_1 = 6 kg/cm^2_g

Outlet pressure = P_2 = 28 kg/cm^2_g

For multistage compressors, the inter stage pressure is,

$P = \sqrt{P_1 P_2}$

P = 12.9 kg/cm^2_g

Power = $E = \frac{1}{\eta}\ Q\ \frac{\gamma RT}{\gamma - 1}\ [\ (\psi)^{\frac{\gamma}{\gamma-1}} - 1\ ]$

Put all variables to get power of compressor,

E = 663 kW

6.11 Expander for HRU

Flowrate of gas = Q = 950 kgmol/hr

R = 8.314 J/mol.K

Inlet temperature = T = 48.8 ^{o}C = 321.8 K

η = 0.8, γ = 1.4

$\psi = P/P_1$

Inlet pressure = P_1 = 120 kg/cm^2_g

Outlet pressure = P_2 = 60 kg/cm^2_g

For multistage compressors, the inter stage pressure is,

$P = \sqrt{P_1 P_2}$ = 84.85 kg/cm^2_g

$$E = \frac{1}{\eta} Q \frac{\gamma RT}{\gamma - 1} [1 - (\psi)^{\frac{\gamma-1}{\gamma}}]$$

Put all variables to get power of compressor,

E = 554.9 kW

6.12 Membrane

Feed:

Table 6. 7: Inlet Composition (HRU)

Components	**Inlet Compositions**
H_2	63 %
N_2	21.5 %
Ar	2 %
CH_4	13.3 %
NH_3	0.2 %

Feed flowrate = 950 kgmol/hr

Feed pressure = 60 kg/cm^2_g

Permeate pressure = 6 kg/cm^2_g

Stage cut = 0.54

Membrane = Polysulfone Membrane Dense membrane, Glassy type

Arrangement = Spiral Wound

Table 6. 8: Permeabilities

Component	**Permeability (kgmol/hr.1000kPa.m^2)**
H_2	0.0337
N_2	0.0006024
Ar	0.0006024
CH_4	0.00003
NH_3	0.00001

Permeate flowrate = stage cut * feed flowrate

= 0.54 * 950

= 511 kgmol/hr

Retentate flowrate = Feed flowrate – Permeate flowrate

= 950 – 511 = 439 kgmol/hr

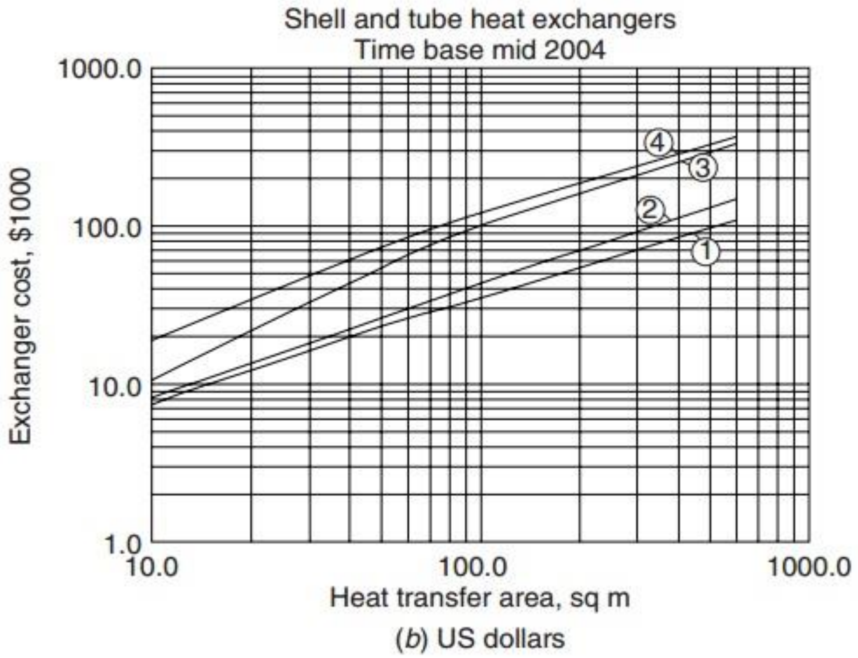
Chapter 7: Costing

Cost Estimates for the Proposed Design

Costing was done with reference from Coulson Richardson's Chemical Engineering Design Volume 6^{th}. Following are the graphs and figures used in cost estimation.

Graph 1

This graph is used for Heat exchangers cost estimation.



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

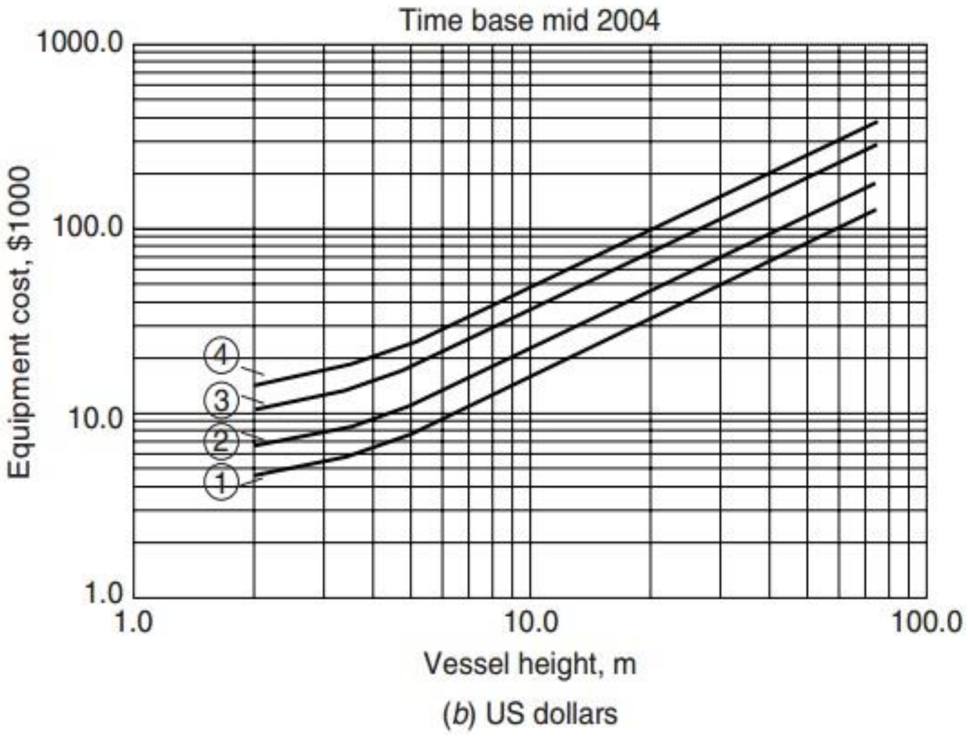
Pressure factors	
1–10 bar	× 1.0
10–20	× 1.1
20–30	× 1.25
30–50	× 1.3
50–70	× 1.5

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

Figure 7.1: Costing of Heat Exchangers (From Coulson and Richardson Volume 6)

Graph 2

Cost of columns is estimated using this graph.



Diameter, m	
①—0.5	③—2.0
②—1.0	④—3.0

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

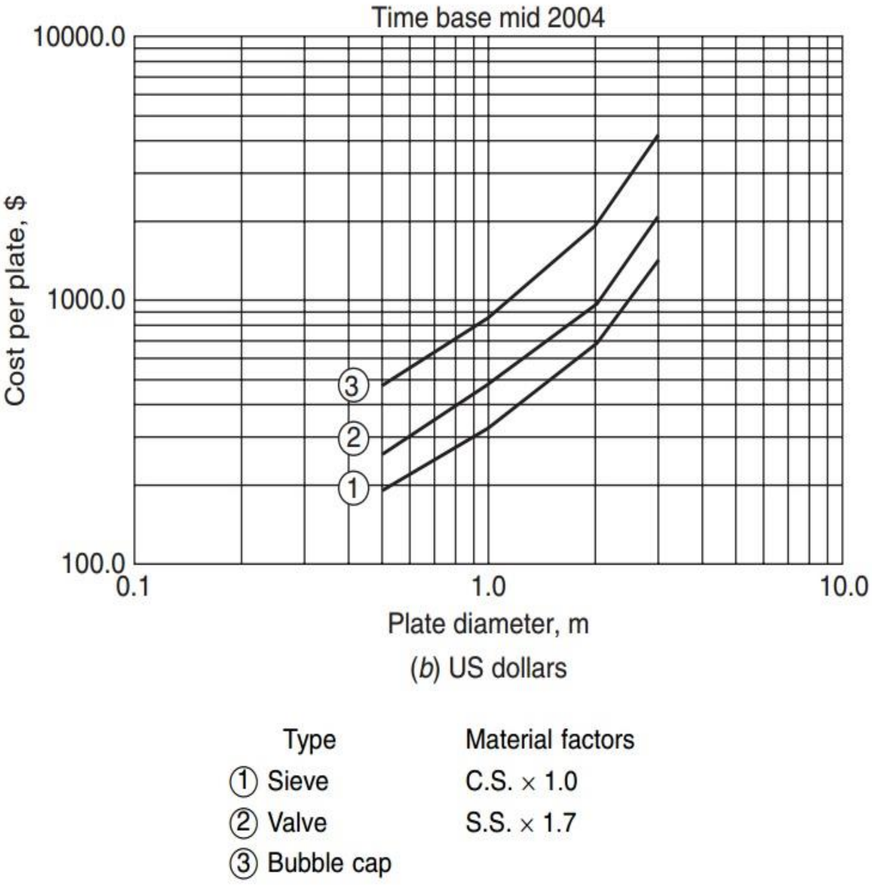
Pressure factors	
1–5 bar	× 1.0
5–10	× 1.1
10–20	× 1.2
20–30	× 1.4
30–40	× 1.6
40–50	× 1.8
50–60	× 2.2

Temperature up to 300°C

Figure 7.2: Cost of Vessels (From Coulson and Richardson Volume 6)

Graph 3

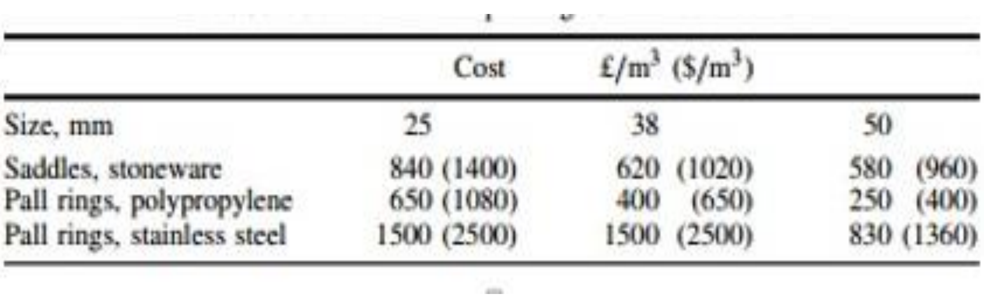
Cost of plates is estimated using this graph.



(*b*) US dollars

Figure 7.3: Cost of Plates (From Coulson and Richardson Volume 6)

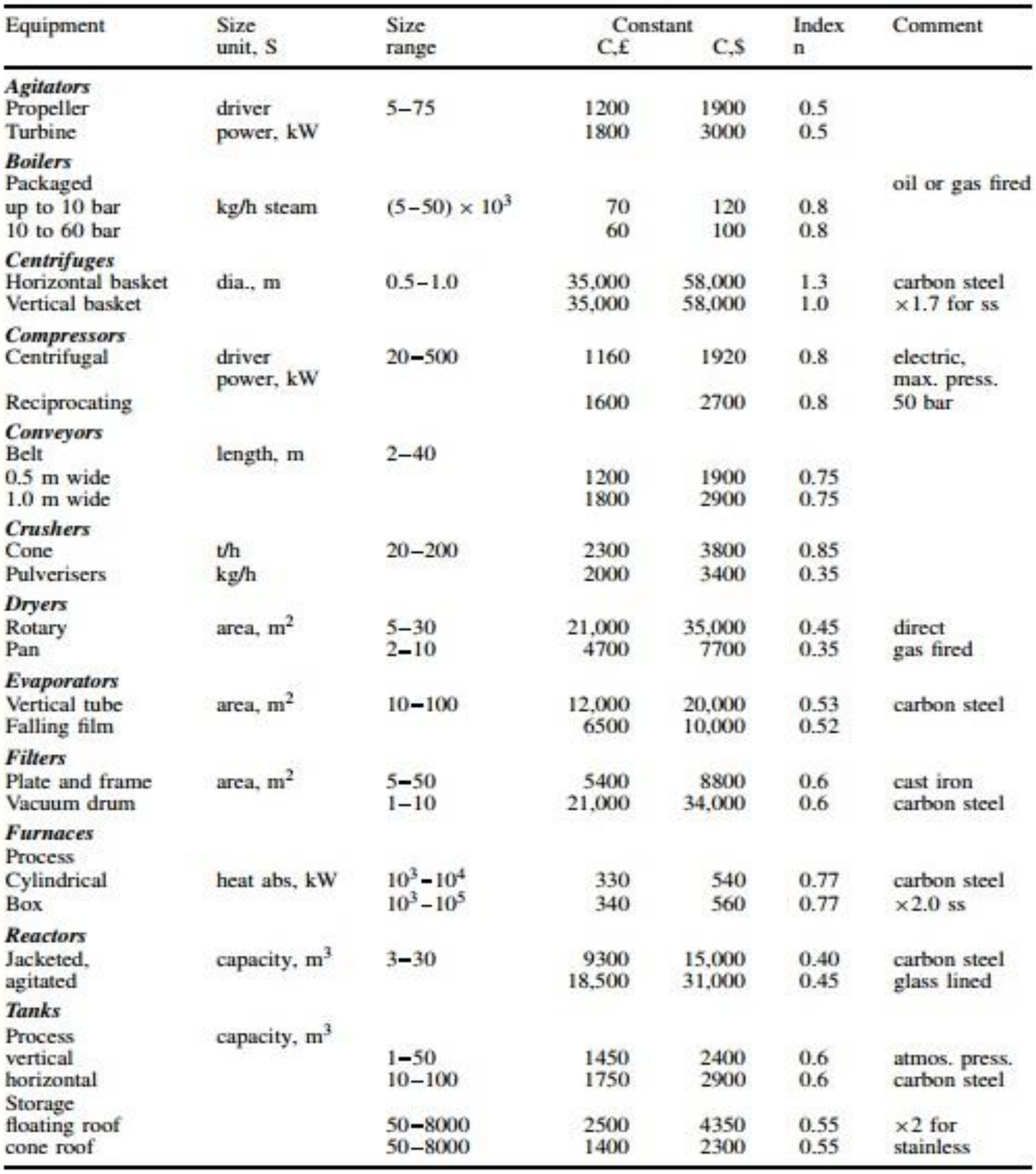
Table 7. 1: Cost of Packing



	Cost	£/m³ ($/m³)	
Size, mm	25	38	50
Saddles, stoneware	840 (1400)	620 (1020)	580 (960)
Pall rings, polypropylene	650 (1080)	400 (650)	250 (400)
Pall rings, stainless steel	1500 (2500)	1500 (2500)	830 (1360)

This table helps in cost estimation of other equipment.

Table 7. 2: Cost of Equipment



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. 50 bar
Reciprocating			1600	2700	0.8	
Conveyors						
Belt	length, m	2–40				
0.5 m wide			1200	1900	0.75	
1.0 m wide			1800	2900	0.75	
Crushers						
Cone	t/h	20–200	2300	3800	0.85	
Pulverisers	kg/h		2000	3400	0.35	
Dryers						
Rotary	area, m^2	5–30	21,000	35,000	0.45	direct
Pan		2–10	4700	7700	0.35	gas fired
Evaporators						
Vertical tube	area, m^2	10–100	12,000	20,000	0.53	carbon steel
Falling film			6500	10,000	0.52	
Filters						
Plate and frame	area, m^2	5–50	5400	8800	0.6	cast iron
Vacuum drum		1–10	21,000	34,000	0.6	carbon steel
Furnaces						
Process						
Cylindrical	heat abs, kW	10^3-10^4	330	540	0.77	carbon steel
Box		10^3-10^5	340	560	0.77	×2.0 ss
Reactors						
Jacketed,	capacity, m^3	3–30	9300	15,000	0.40	carbon steel
agitated			18,500	31,000	0.45	glass lined
Tanks						
Process	capacity, m^3					
vertical		1–50	1450	2400	0.6	atmos. press.
horizontal		10–100	1750	2900	0.6	carbon steel
Storage						
floating roof		50–8000	2500	4350	0.55	×2 for
cone roof		50–8000	1400	2300	0.55	stainless

Ammonia Recovery Unit

Rich Solution Exchanger 1 (ARU)

Shell and tube heat exchanger of floating head type.

Heat Transfer Area = 18.77 m^2

Bare Equipment cost = $ 12500 (Using Graph 1)

Pressure factor = 1.3

Type factor = 1.0

Cost = Bare equipment cost * Pressure factor * Type factor

Cost = $ 16250

Rich Solution Exchanger 2 (ARU)

Shell and tube heat exchanger of floating head type.

Heat Transfer Area = 13.20 m^2

Bare Equipment cost = $ 9000 (Using Graph 1)

Pressure factor = 1.3

Type factor = 1.0

Cost = Bare equipment cost * Pressure factor * Type factor

Cost = $ 11700

Lean Solution Cooler (ARU)

Shell and tube heat exchanger of floating head type.

Heat Transfer Area = 28.29 m^2

Bare Equipment cost = $ 16500 (Using Graph 1)

Pressure factor = 1.25

Type factor = 1.0

Cost = Bare equipment cost * Pressure factor * Type factor

Cost = $ 20625

Condenser (ARU)

Shell and tube heat exchanger of floating head type.

Heat Transfer Area = 20.2 m^2

Bare Equipment cost = $ 13700 (Using Graph 1)

Pressure factor = 1.25

Type factor = 1.0

Cost = Bare equipment cost * Pressure factor * Type factor

Cost = $ 17125

Reboiler (ARU)

Shell and tube heat exchanger of kettle type.

Heat Transfer Area = 28.6 m^2

Bare Equipment cost = $ 16750 (Using Graph 1)

Pressure factor = 1.25

Type factor = 1.3

Cost = Bare equipment cost * Pressure factor * Type factor

Cost = $ 27220

Purge Gas Absorber (ARU)

Height of Column = 9.96 m

Column’s Diameter = 0.915 m

Bare Vessel Cost = $ 14000 (Using Graph 2)

Pressure Factor = 2.2

Material Factor = 1.0

Cost = Bare vessel cost * Pressure factor * Material factor

Cost of column = $ 30800

No of plates in column = 19

Cost per plate = $ 180

Cost of 19 plates = $ 3420

Total cost of absorption column = $ 34220

Off Gas Absorber (ARU)

Height of Column = 5.63 m

Column's Diameter = 0.47 m

Bare Vessel Cost = $ 4800 (Using Graph 2)

Pressure Factor = 1.4

Material Factor = 1.0

Cost = Bare vessel cost * Pressure factor * Material factor

Cost of column = $ 6720

No of plates in column = 12

Cost per plate = $ 130

Cost of 19 plates = $ 1560

Total cost of absorption column = $ 8280

Distillation Column (ARU)

Height of Column = 12.65 m

Column's Diameter = 1.19 m

Bare Vessel Cost = $ 22175 (Using Graph 2)

Pressure Factor = 1.4

Material Factor = 1.0

Cost = Bare vessel cost * Pressure factor * Material factor

Cost of column = $ 31045

No of plates in column = 20

Cost per plate = $ 250

Cost of 19 plates = $ 5000

Total cost of distillation column = $ 36045

Mixer (ARU)

Volume = 9.16 m^3 (Height = 2.219 m, Diameter = 0.739 m)

From Table 4,

Size unit = 9.16 m^3

Cost Constant = C = $ 2900

Index = n = 0.6

Cost of equipment = $C * S^n$

Cost = $ 10955

Pump 1 (ARU)

Power = 0.84 KW

Centrifugal Type Pump

Capacity = 1.76 m^3/hr

Cost = $ 100 (Price estimate from Alibaba.com)

Pump 2 (ARU)

Power = 17.51 KW

Centrifugal Type Pump

Capacity = 5.86 m^3/hr

Cost = $ 750 (Price estimate from Alibaba.com)

Total purchase cost of major equipment items (PCE) = $ 183271

Estimation of fixed capital cost for fluids processing plant

f1	Equipment erection	0.40
f2	Piping	0.70
f3	Instrumentation	0.20
f4	Electrical	0.10
f5	Buildings	none required
f6	Utilities	not applicable
f7	Storages	provided in PCE
f8	Site development	not applicable
f9	Ancillary buildings	none required

Total physical plant cost (PPC) = 183271 * (1 + 0.4 + 0.7 + 0.2 + 0.1) =~ $ 439850

f10	Design and Engineering	0.30
f11	Contractor’s Fee	none (unlikely to be used for a small, plant project)
f12	Contingencies	0.10

Fixed capital = 439850 (1 + 0.3 + 0.1) = $ 615790

Investment Needed for Plant (rounded) = $ 620000

Working capital (5% of fixed capital) = $ 30780

Operating time allowing for attainment = 365*0.95 = 347 days

Variable Costs

1. Cooling water required = 3190 kgmol /hr = 57420 kg/hr = 57.42 t/hr

Cooling water cost = 57.42t * 0.1 $/t = $ 5.742

Cooling water cost/year = 347 * 24 * 5.742 = $ 47819

2. Steam required = 71.5 kgmol/hr = 1287 kg/hr =1.287 t/hr

Steam cost = 1.287t * 12$/t = $ 15.444/hr

Steam cost/year = 347 * 24 * 15.444 = $ 128617

3. Process Water required = 408 kgmol/hr = 7344 kg/hr = 7.344 t/hr

Process water cost = 7.3444t * 0.5$/t = $ 3.672

Process water cost/year = 347 * 24 * 3.672 = $ 30580

4. Electricity required for ARU = 4085 KWh

Electricity cost = 4085 KWh * 0.0015 $/KWh = $ 6.1275

Electricity cost/year = 347 * 24 * 6.1275 = $ 51030

Total Variable cost = TVC = $ 258046

Fixed Costs

5. Maintenance (5% of fixed capital) = C5= $ 30789.5

6. Operating labor say = C6 = $ 150000 / year

7. Supervision, no additional supervision would be needed

8. Plant overheads (50% of operating labor) = C8 = $ 75000

9. Laboratory (take as 30% of operating labor) = C9 = $ 45000

10. Capital charges (6% of fixed capital) = C10 = $ 36947

11. Insurance (1% of fixed capital) = C11 = $ 6158

12. Local taxes are neglected

13. Royalties not applicable

Total fixed cost **= C5 + C6 + C8 + C9 + C10 + C11 = TFC = $ 343895**

14. Sales expense not applicable

15. General overheads not applicable

16. Research and development not applicable

Annual operating cost, rounded = TVC + TFC = $ 601940

Hydrogen Recovery Unit (HRU):

Total Plant Investment (TPI)

Total membrane Module Cost (MC) = \$5/ft^2 = \$ 306615

Installed Compressor Cost (CC) = \$ 347550 (Using Table 4)

HRU Cooler Cost (HCC) = \$ 31850

Fixed Cost (FC) = MC + CC + HCC = \$ 686015

Base Plant Cost (BPC) = 1.12 × FC = \$ 768337

Project Contingency (PC) = 0.2 × BPC = \$ 153667

Total Facilities Investment (TFI) = BPC + PC = 922004

Start-up Cost (SC) = 0.10 × VOM = \$ 49792

Total Plant Investment (TPI) = TFI + SC = \$ 971796

Annual Capital related Cost (CRC) = 0.2 × TPI = \$ 194360

Annual Variable Operating & Maintenance Cost (VOM)

Contract & Material Maintenance Cost (CMC) = 0.05 × TFI = \$ 46100

Local Taxes & Insurances (LTI) = 0.015 × TFI = \$ 13830

Direct labor Cost (DLC) \$15/hr = \$ 124920

Labor Overhead Cost (LOC) = 1.15 × DL = \$ 143658

Membrane Replacement Cost (MRC) \$ 2.5/ ft2 of membrane = \$ 153307

Utility Cost (UC) \$ 0.07 /kWh = \$ 16104

Annual variable Operating & Maintenance Cost (VOM) = CMC + LTI + DLC + LOC + MRC + UC = \$ 497919

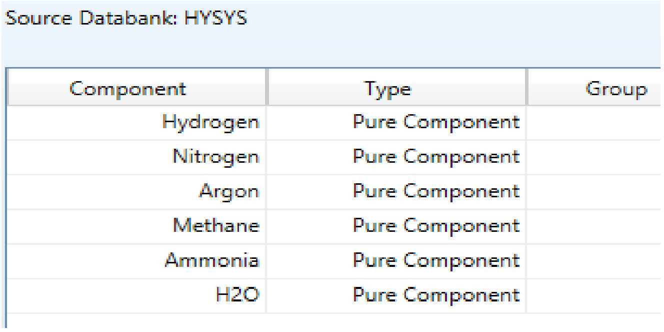
Total cost of HRU (Installation + Operating) = TPI + CRC + VOM = \$ 1664075

Chapter 8: Simulation

Aspen Hysys is used for simulation of the project. Simulation was done for the verification of results of ARU (Ammonia Recovery Unit) and HRU (Hydrogen Recovery Unit).

Membranes material balance is done on the Aspen Hysys. The system consists of following components:

- Hydrogen
- Nitrogen
- Ammonia
- Methane
- Argon
- Water



Source Databank: HYSYS

Component	Type	Group
Hydrogen	Pure Component	
Nitrogen	Pure Component	
Argon	Pure Component	
Methane	Pure Component	
Ammonia	Pure Component	
H2O	Pure Component	

Figure 5.1: Component list

8.1 Property Package

Property package is chosen according to the components and the operating conditions. The property package we used is ***Peng Robinson.***

Type of Component System:

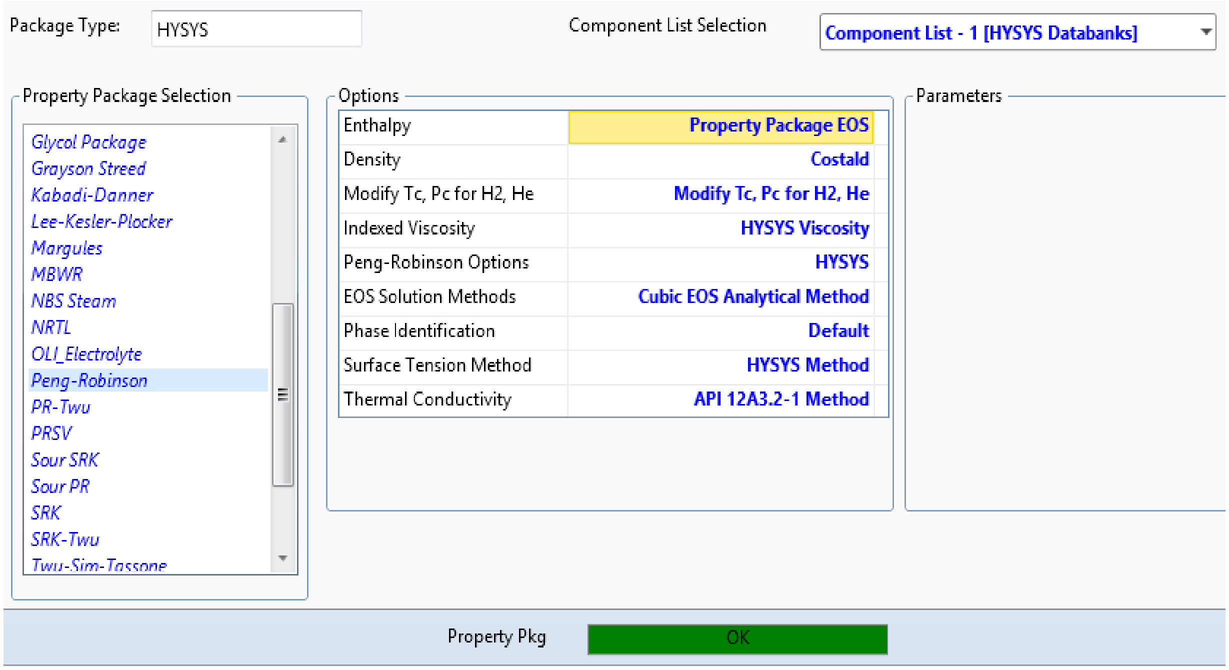
Chemical

Pressure:

High Pressure (>10 bar)

Property Packages Available for this system:

BWRS, GCEOS, Glycol Package, Kabadi-Danner, Lee-Kesler-Plocker, MBWR, **Peng-Robinson**, PR-Twu, PRSV, Sour SRK, Sour PR, SRK, SRK-Twu, or Zudkevitch-Joffee



Package Type: HYSYS

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

Property Package Selection

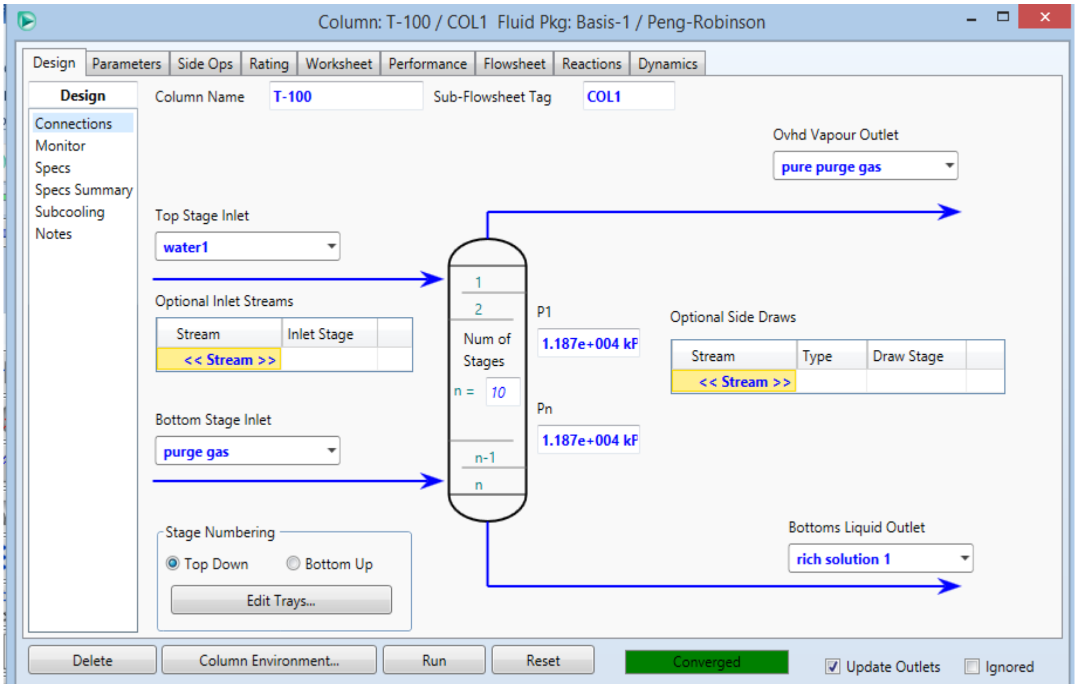
- Glycol Package
- Grayson Streed
- Kabadi-Danner
- Lee-Kesler-Plocker
- Margules
- MBWR
- NBS Steam
- NRTL
- OLI_Electrolyte
- Peng-Robinson
- PR-Twu
- PRSV
- Sour SRK
- Sour PR
- SRK
- SRK-Twu
- Twu-Sim-Tassone

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

Figure 8.2: Property Package

8.2 Absorber 1



Column: T-100 / COL1 Fluid Pkg: Basis-1 / Peng-Robinson

Column Name: T-100

Sub-Flowsheet Tag: COL1

Ovhd Vapour Outlet: pure purge gas

Top Stage Inlet: water1

Bottom Stage Inlet: purge gas

Bottoms Liquid Outlet: rich solution 1

P1: 1.187e+004 kF

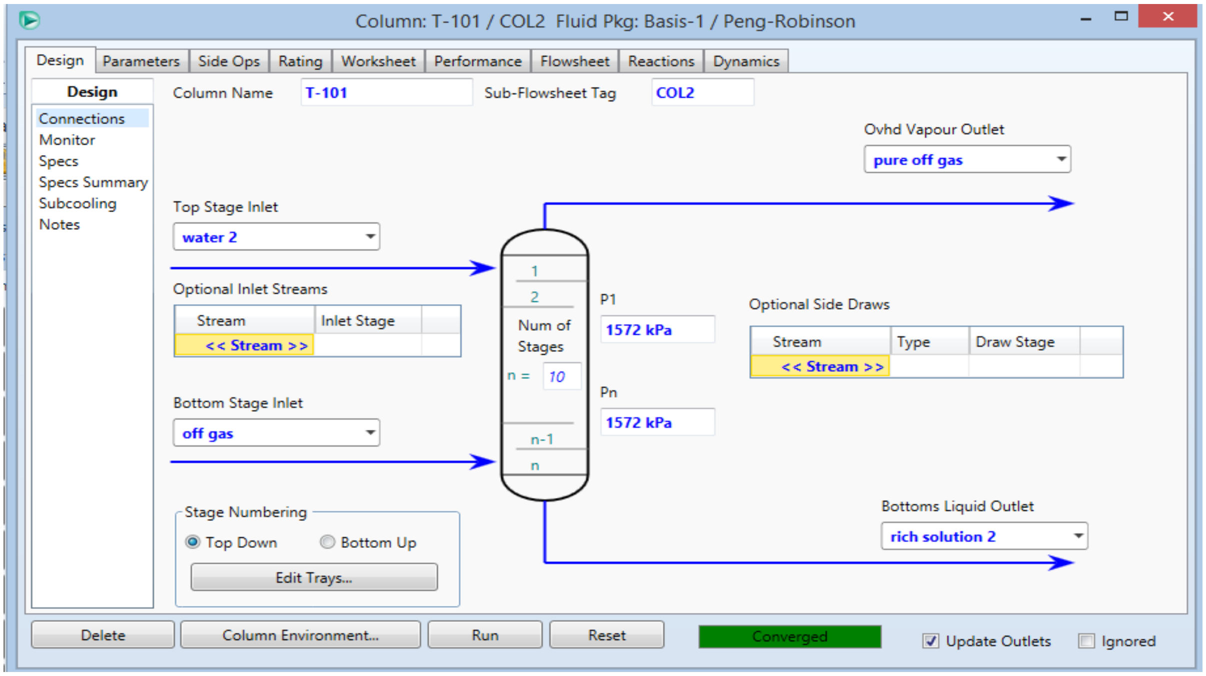
Pn: 1.187e+004 kF

Num of Stages: n = 10

Stage Numbering: ● Top Down ○ Bottom Up

Figure 8.3: Simulation of Absorber 1

8.3 Absorber 2



Column: T-101 / COL2 Fluid Pkg: Basis-1 / Peng-Robinson

Column Name: T-101

Sub-Flowsheet Tag: COL2

Ovhd Vapour Outlet: pure off gas

Top Stage Inlet: water 2

Bottom Stage Inlet: off gas

P1: 1572 kPa

n = 10

Pn: 1572 kPa

Bottoms Liquid Outlet: rich solution 2

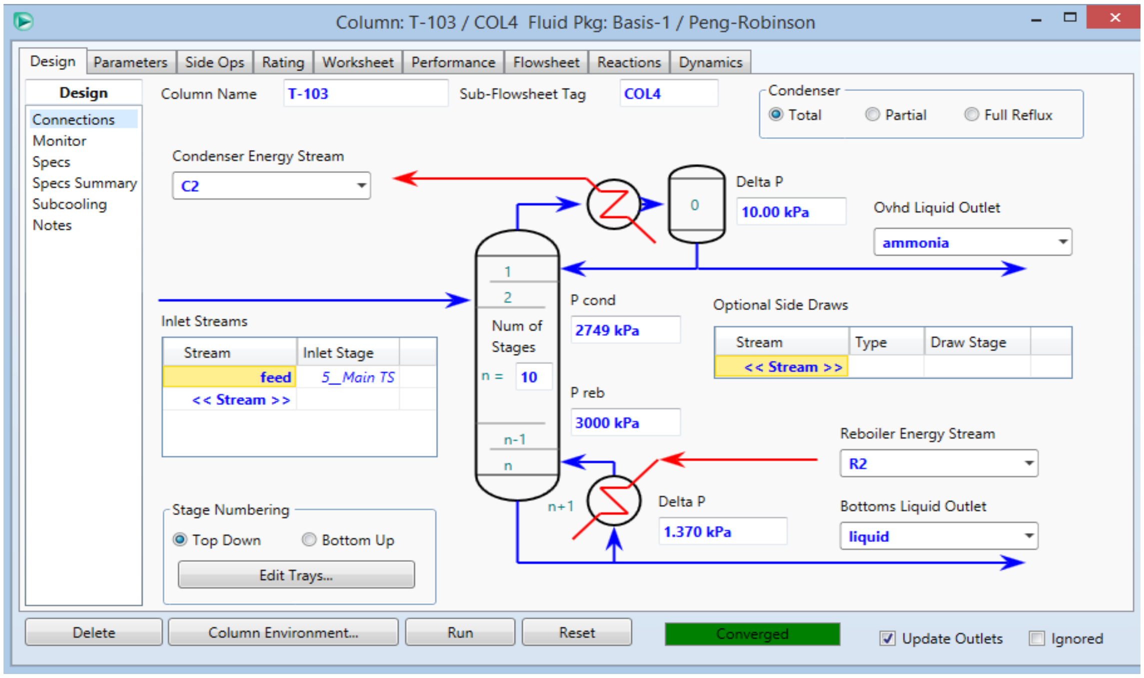
Stage Numbering: ● Top Down ○ Bottom Up

Converged

☑ Update Outlets ☐ Ignored

Figure 8.4: Simulation of Absorber 2

8.4 Distillation



Column: T-103 / COL4 Fluid Pkg: Basis-1 / Peng-Robinson

Column Name: T-103

Sub-Flowsheet Tag: COL4

Condenser: ● Total ○ Partial ○ Full Reflux

Condenser Energy Stream: C2

Delta P: 10.00 kPa

Ovhd Liquid Outlet: ammonia

Inlet Streams

Stream	Inlet Stage
feed	5__Main TS
<< Stream >>	

P cond: 2749 kPa

n = 10

P reb: 3000 kPa

Reboiler Energy Stream: R2

Delta P: 1.370 kPa

Bottoms Liquid Outlet: liquid

Stage Numbering: ● Top Down ○ Bottom Up

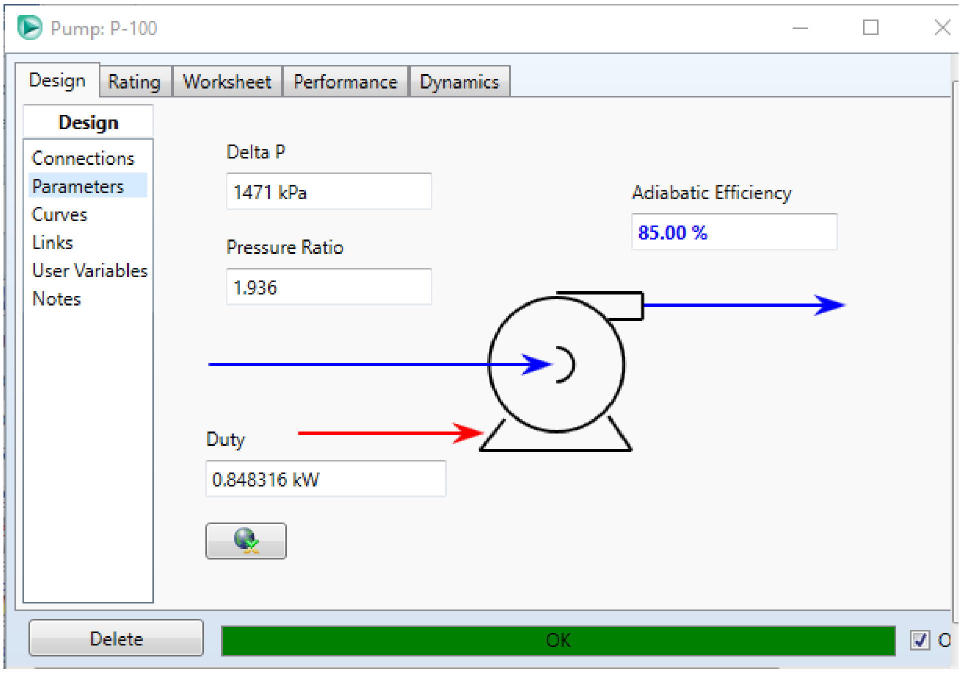
Converged

☑ Update Outlets ☐ Ignored

Figure 8.5: Simulation of Distillation Column

8.5 Pumps

Power of pumps used in ARU are taken from the Hysys. All operating conditions (temperature and pressure), flow rates of streams and change in pressure required are entered into hysys and it calculated the power required for the pumps as shown below:



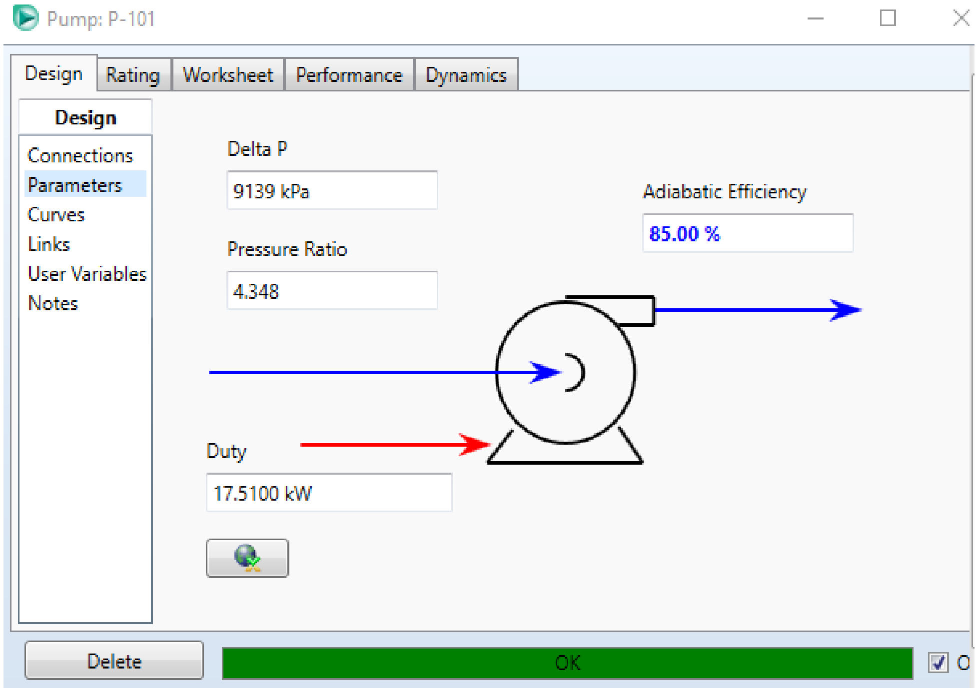
Delta P: 1471 kPa

Pressure Ratio: 1.936

Adiabatic Efficiency: 85.00 %

Duty: 0.848316 kW

Figure 8.6: Pump 1



Delta P: 9139 kPa

Pressure Ratio: 4.348

Adiabatic Efficiency: 85.00 %

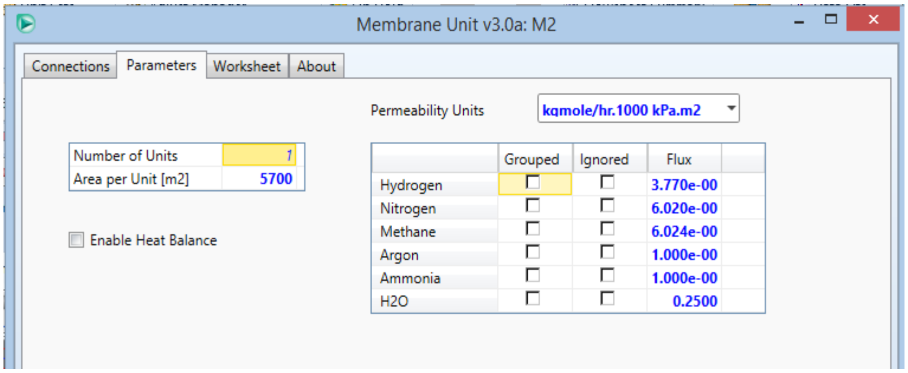
Duty: 17.5100 kW

Figure 8.7: Pump 2

Similarly all the operating conditions are used for shell and tube heat exchangers and our results are verified.

8.6 Hydrogen Recovery Unit (HRU)

Membranes are used for hydrogen recovery from the purge gas. We knew the permeability values of components of purge gas for the membrane selected for separation. These values are converted into permeance using thickness of membrane and the values of permeance are put into HYSYS membrane model. The operating conditions and flow rate are also entered. From simulation, we got permeate and retentate flow rates and their respective compositions. It also gave the total area of membrane needed for the required recovery.



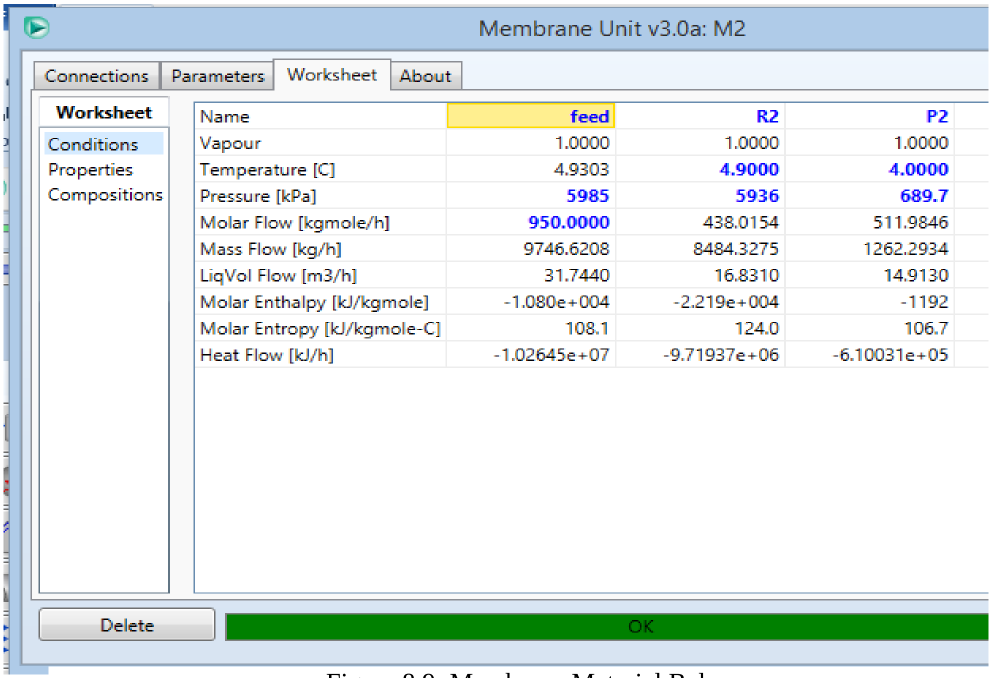
Permeability Units: kgmole/hr.1000 kPa.m2

Number of Units	1
Area per Unit [m2]	5700

☐ Enable Heat Balance

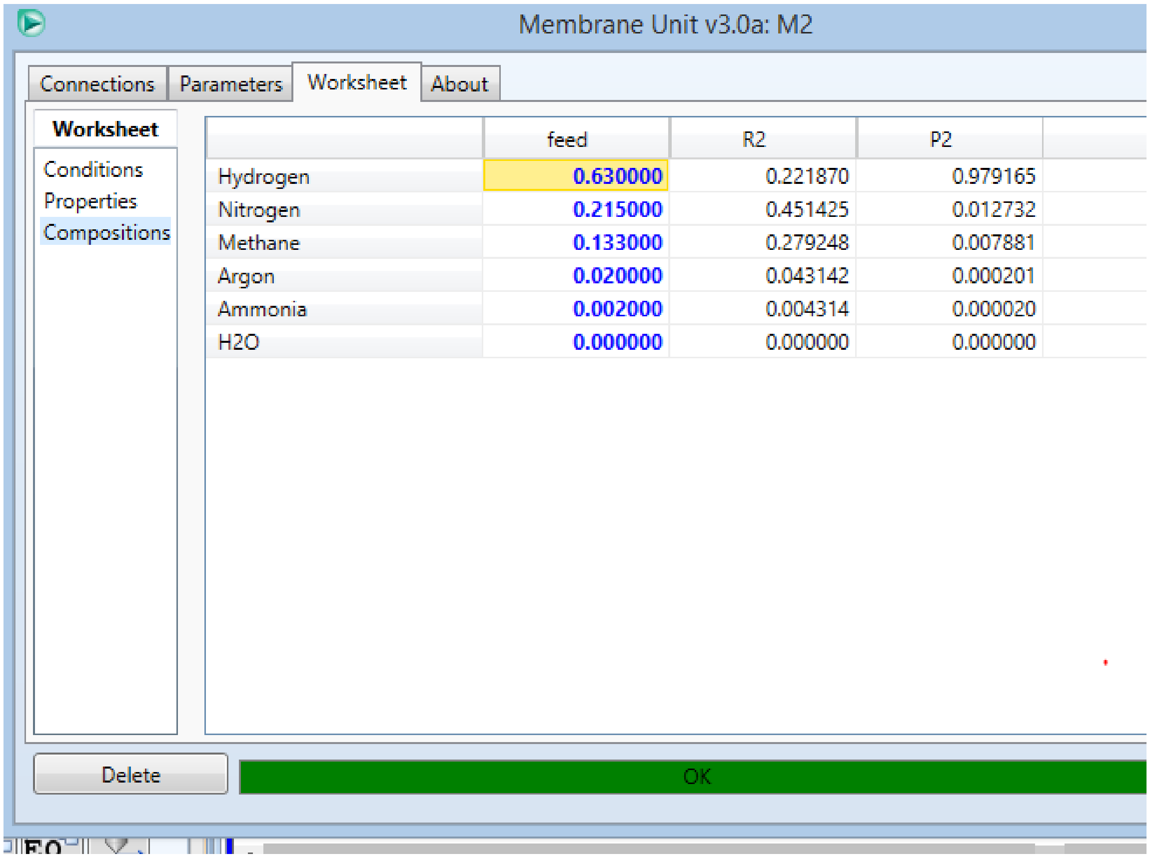
	Grouped	Ignored	Flux
Hydrogen	☐	☐	3.770e-00
Nitrogen	☐	☐	6.020e-00
Methane	☐	☐	6.024e-00
Argon	☐	☐	1.000e-00
Ammonia	☐	☐	1.000e-00
H2O	☐	☐	0.2500

Figure 8.8: Membrane Permeability



Name	feed	R2	P2
Vapour	1.0000	1.0000	1.0000
Temperature [C]	4.9303	4.9000	4.0000
Pressure [kPa]	5985	5936	689.7
Molar Flow [kgmole/h]	950.0000	438.0154	511.9846
Mass Flow [kg/h]	9746.6208	8484.3275	1262.2934
LiqVol Flow [m3/h]	31.7440	16.8310	14.9130
Molar Enthalpy [kJ/kgmole]	-1.080e+004	-2.219e+004	-1192
Molar Entropy [kJ/kgmole-C]	108.1	124.0	106.7
Heat Flow [kJ/h]	-1.02645e+07	-9.71937e+06	-6.10031e+05

Figure 8.9: Membrane Material Balance



	feed	R2	P2
Hydrogen	0.630000	0.221870	0.979165
Nitrogen	0.215000	0.451425	0.012732
Methane	0.133000	0.279248	0.007881
Argon	0.020000	0.043142	0.000201
Ammonia	0.002000	0.004314	0.000020
H2O	0.000000	0.000000	0.000000

Figure 8.10: Composition of HRU Stream

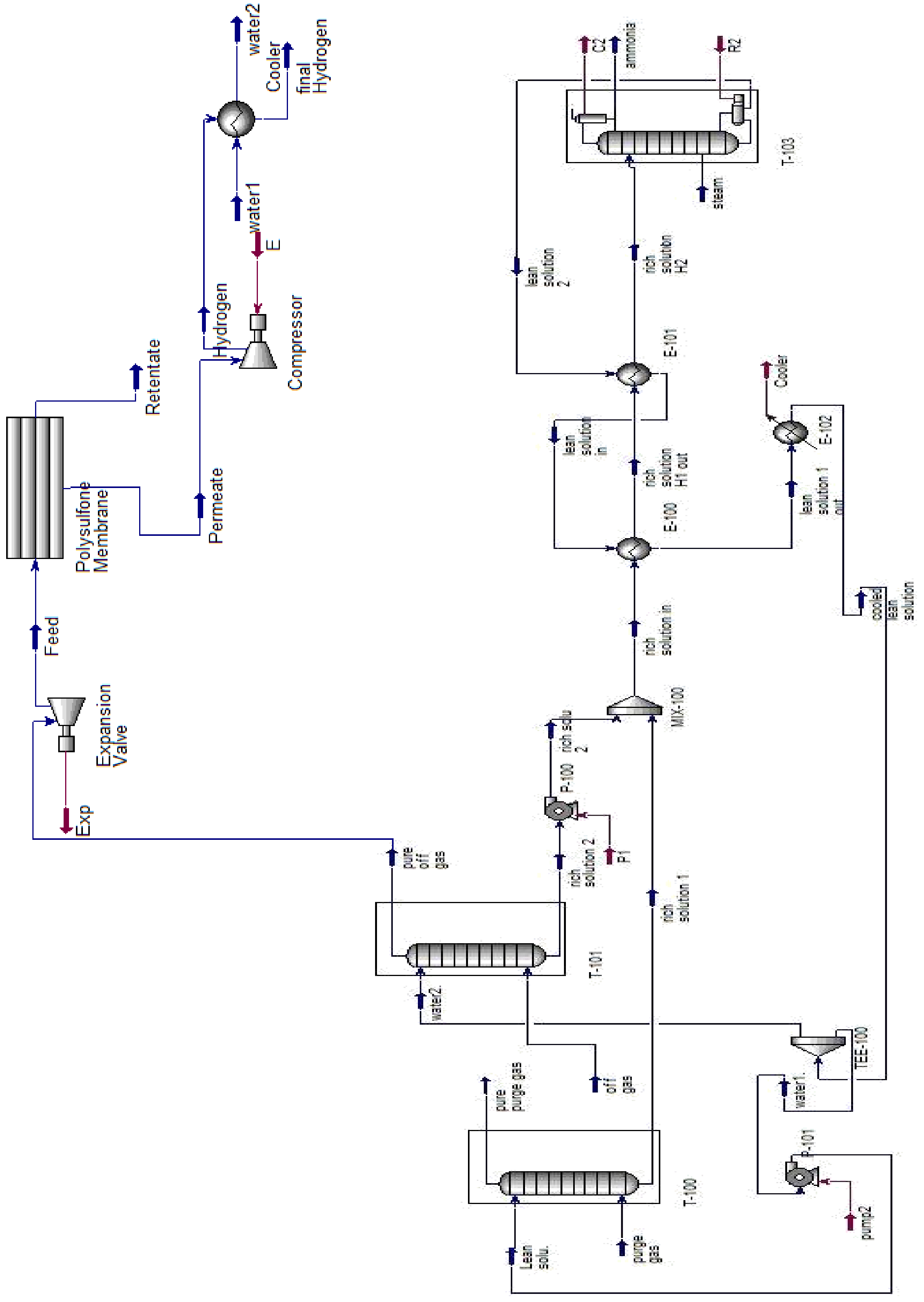
Recovery:

$$\text{Hydrogen Recovery} = \frac{Hydrogen\ flowrate\ in\ permeate}{Hydrogen\ flowrate\ in\ feed}$$

$$= \frac{0.979*512}{0.63*950} * 100 = 83.73\ \%$$

$$\text{Stage cut} = \frac{Permeate\ flow\ rate}{Feed\ flowrate}$$

$$= 512/950 = 0.54$$


Expansion Valve
Exp
Feed
Polysulfone Membrane
Retentate
Permeate
Hydrogen
Compressor
E
water1
Cooler
water2
final Hydrogen
pure off gas
T-101
water2.
off gas
pure purge gas
Lean solu.
purge gas
T-100
rich solution 1
rich solution 2
P1
P-100
rich solu 2
MIX-100
rich solution in
E-100
rich solution H1 out
lean solution in
E-101
rich solution H2
lean solution 2
T-103
steam
C2
ammonia
R2
Cooler
E-102
lean solution 1 out
cooled lean solution
TEE-100
water1.
P-101
pump2

Chapter 9: Hazard and Operability (HAZOP) Study

Equipment: **Absorber**

Intention: To remove ammonia from the purge gas by absorption in water.

Guide Word	Deviation	Cause	Consequence & Action
Line 1 – Intention: To provide water for ammonia absorption from purge gas			
No	Flow	Pump Failure or Power loss	No ammonia absorption will take place. Liquid Level sensor is needed.
More	Flow	Failure of flow control valve	Weeping occurs in absorption column. Level sensor is needed. Liquid flow should be decreased.
Less	Flow	Failure of flow control valve	Flooding occurs in absorption column. Level sensor is needed. Liquid flow should be increased.

Guide Word	Deviation	Cause	Consequence & Action
Line 2 – Intention: Getting ammonia free purge gas from absorber.			
No	Concentration	Water pump failure or purge gas control valve failure	No ammonia absorption will take place. Ammonia comes out along with gas from top. Level sensor and alarm for gas flow failure is needed.

Less	Concentration	Failure of liquid flow control valve or purge gas control valve	Little amount of ammonia is absorbed. Level sensor and alarm for gas flow failure is needed.

Guide Word	**Deviation**	**Cause**	**Consequence & Action**
Line 3 – Intention: To keep the pressure high (optimum) in the absorption column.			
No	Pressure	High pressure drop, high liquid loading	Inefficient absorption of ammonia and weeping occurs. Pressure controller is required at the upstream of absorption column. Gas flow controller is also needed for the exit gas stream. Good packing material should be used.
More	Pressure	Pressure drop is low	Efficient absorption takes place. Pressure controller is required at the upstream of absorption column. Gas flow controller is also needed for the exit gas stream.

Equipment: **Distillation Column**

Intention: To recover Ammonia with maximum purity

Guide Word	**Deviation**	**Cause**	**Consequence & Action**
Line 4– Intention: To provide feed of rich ammonia solution to the distillation column.			

No	Flow	Pump failure or Power Loss	No separation takes place. Level sensor is required for distillation column.
More	Flow	Failure of Flow control Valve	Temperature falls and inefficient separation of ammonia. Level sensor is required for distillation column.
Less	Flow	Failure of Flow control Valve	Temperature rises inside column, inefficient use of large size column for separation. Level sensor is required for distillation column.

Guide Word	**Deviation**	**Cause**	**Consequence & Action**
Line 5- Intention: To send the reflux back to the Column			
No	Flow	Reflux Control Valve Failure	Ammonia product is not purified and contains water as well. Composition analyser must be used for product.
Less	Flow	Reflux Control Valve Failure	Required product purification is not achieved. Composition analyser must be used for product.
More	Flow	Reflux Control Valve Failure	Distillate obtained will be less. Purification is more than

			required. Composition analyzer must be used for product.

Guide Word	Deviation	Cause	Consequence & Action
Line 6– Intention: To provide the steam for heating to the reboiler			
More	Flow	Steam Flow Control Valve Failure	Temperature of reboiler will increase and may become uncontrolled. More heat would be wasted. Temperature sensor for bottom stream should be used.
Less	Flow	Steam Flow Control Valve Failure	Temperature will decrease and the heat required for re boiling would not be sufficient enough which will cause inefficient separation in the column.

Guide Word	Deviation	Cause	Consequence & Action
Line 7 – Intention: To move the NH3 produced towards the Hydrogen Recovery Unit			
More	Flow	The downstream pumping increases	The level in the reflux drum falls, and more feed to the column and hence higher flowrates would be required.

Less	Flow	The Flow control valve or the pump downstream fails	The level in the reflux drum increases to dangerous levels and pressure may build in it which may cause serious damage to the equipment.

Chapter 10: Instrumentation and Process Control

10.1 Absorbers

The liquid and vapor flow rates are kept optimum to get maximum absorption of ammonia from purge gas. Absorption occurs at high pressure and low temperature. Pressure sensor is placed at the gas exit of absorber. If the pressure is too low, the inlet gas flow rate and exit gas flowrates are controlled by using flow control valve. A specific amount of liquid is also required for absorption and to avoid the problems of weeping and flooding, a level sensor is used for measuring liquid level inside column. This liquid level is maintained by controlling the water flow using flow controller valve. A composition analyzer is also placed at the top of absorber for finding exit gas compositions and checking the ammonia recovery from the purge gas. If enough ammonia is not removed, then gas velocity and pressure conditions are changed by using flow control valves.

10.2 Cooler for lean solution

The lean solution from the bottom of distillation column after passing through series of heat exchangers, is cooled in a water cooler. Cooling water removes its energy and decreases its temperature to that required for absorption. A temperature sensor is used for measuring the outlet temperature of the lean solution. A flow control valve for cooling water is used to achieve the required temperature of lean solution.

If the outlet temperature of stream is less than desired temperature, signal is sent to the cooling water flow control valve and flowrate of cooling water is decreased. If the outlet temperature of stream is more than desired temperature, signal is sent to the cooling water flow control valve and flowrate of cooling water is increased.

10.3 Distillation Column

The purpose of distillation is to separate ammonia from the water as ammonia is the desired product and water obtained could be recycled for absorption. Flow control valve is used for feed. Level sensor detects the liquid level in the column and helps the feed control valve to operate accordingly. A reflux control valve is used after condenser for controlling reflux flow and distillate flow. Composition analyzer is used which check the purity of ammonia in the product stream. If it is less than the desired amount, reflux is increased by using reflux valve while in other case, reflux is decreased by using reflux flow control valve.

Make up steam is added in the distillation column as some water is lost along with distillate (ammonia product). Flow control valve for this feed stream is used. The composition analyzer gives the amount of water lost and this signal is transmitted to the steam controller and its amount is controlled accordingly.

10.4 Condenser

In the condenser, top product of distillation column is condensed and cooled up to desired temperature. This temperature is achieved by cooling water. A temperature sensor is used for measuring the outlet temperature of the product. A flow control valve for cooling water is used to achieve the required product temperature.

If the outlet temperature of stream is less than desired temperature, signal is sent to the cooling water flow control valve and flowrate of cooling water is decreased. If the outlet temperature of stream is more than desired temperature, signal is sent to the cooling water flow control valve and flowrate of cooling water is increased.

10.5 Reboiler

Bottom of distillation column enters reboiler. Some amount of liquid is collected as product while remaining vapors are sent back to distillation column. The reboiler ratio is controlled by control valve. It depends upon the purity of bottom product. Flow of vapors recycled and bottom product taken are changed by the flow controller depending upon the requirement.

This bottom liquid is heated by a steam. The temperature of bottom product is maintained by the steam flow. A temperature sensor is used for measuring the outlet temperature of the product. A flow control valve for steam is used to achieve the required product temperature.

If the outlet temperature of stream is less than desired temperature, signal is sent to the cooling water flow control valve and flowrate of steam is increased. If the outlet temperature of stream is more than desired temperature, signal is sent to the steam flow control valve and flowrate of steam is decreased.

10.6 Cooler after Hydrogen Recovery

Hydrogen gas obtained after hydrogen recovery unit is compressed to achieve the required pressure for recycling. During this process, its temperature rises and it is necessary to cool it down so that it can be recycled. This cooling is done by cooling water.

A temperature sensor is used for measuring the outlet temperature of the hydrogen stream. A flow control valve for cooling water is used to achieve the required temperature of hydrogen stream.

If the outlet temperature of stream is less than desired temperature, signal is sent to the cooling water flow control valve and flowrate of cooling water is decreased. If the outlet temperature of stream is more than desired temperature, signal is sent to the cooling water flow control valve and flowrate of cooling water is increased.

Conclusion

The proposed design for the ARU is environmentally friendly as it keeps the ammonia from the purge and off gases from entering the environment, hence decreasing air pollution. As in FFC, purge gas is sent to boilers so ARU prevents NOx formation in boilers. This plant can be established with an annual cost of $601940, while the recovered ammonia can be used for various purposes like fertilizers, explosives & detergents etc. The design for HRU can be established and operated with an annual cost of $1664075 which is comparatively less than the cryogenic unit. Moreover the recovered Hydrogen can be used as a raw material for Ammonia production and various other applications. As Hydrogen will be recycled to the Ammonia synthesis loop, it will reduce the fresh Hydrogen feed requirement hence making the process economically efficient.

References

[1] **Crisalle, Oscar D. 2013.** *Ammonia Gas Absorption.* 2013.

[2] David DortMundt, Mark Schott. *Development of Gas Separation Membranes for Highly Sour Service.* USA : Illinois 60017.

[3] Fernando, Sushil Adhikari and Sandun. *Hydrogen Membrane Separation Techniques.* s.l. : Department of Agricultural and Biological Engineering, Mississippi State UniVersity, Mississippi State,.

[4] Freeman, Benny D. 1998. *Basis of Permeability/Selectivity Tradeoff Relations in Polymeric Gas.* North Carolina : s.n., 1998.

[5] Holmes, Michael. 2010. *Hydrogen Separation Membranes.* 2010.

[6] Hu, Professor Xijun. *Advanced Physico-Chemical Treatment Processes.*

[7] Kolmetz, Karl. 2013. *REBOILER ENGINEERING DESIGN* . #03-12 Block Aronia,Jalan Sri Perkasa 2 Taman Tampoi Utama 81200 Johor Bahru : s.n., 2013.

[8] L. M. Robesont, W. F. Burgoyne, M. Langsam, A. C. Savoca and C. F. Tien. *High performance polymers for membrane.* USA : Air Products and Chemicals, Inc., Allentown, PA 18195,.

[9] *Membrane Technology for Gas Separation.* **Paola Bernardo, Gabriele Clarizia. 2013.** Italy : s.n., 2013, Vol. 32. 978-88-95608-23-5.

[10] Michael Pilutti, PE, Julia E. Nemeth, PE,. *TECHNICAL AND COST REVIEW OF COMMERCIALLY AVAILABLE.*

[11] Mukherjee, Rajiv. 1998. *CHEMICAL ENGINEERING PROGRESS.* s.l. : Engineers India Ltd, 1998.

[12] Nenoff, Nathan W. Ockwig and Tina M. 2006. *Membranes for Hydrogen Separation.* New Mexico : s.n., 2006.

[13] *Process Design Analyses of CO2 Capture from Natural Gas by Polymer Membrane.* **Arshad Hussain, Habib Nasir and Muhammad Ahsan.** s.l. : J.Chem.Soc.Pak., 2014, Vol. 36.

[14] R.Hughes. 2000. *Hydrogen Recovery using Inorganic Membranes.* Salford, UK : Academics Press, 2000.

[15] REYES, GERALD ZUNIGA. *LAYER OF PROTECTION ANALYSIS APPLIED TO AMMONIA REFRIGERATION SYSTEMS.* 2008.

[16] Richardson's, and Coulson. *CHEMICAL ENGINEERING DESIGN, Volume 6, Edition 4.* 2005. ISBN 0 7506 6538 6.

[17] S. Sridhar, Sharifah Bee and Suresh K. Bhargava. *Membrane-based Gas Separation: Principle, Applications and Future Potential.* s.l. : Hyderabad, India – 500007.

[18] Sander, Rolf. *Compilation of Henry's Law Constants for Inorganic and Organic Species of Potential Importance in Environmental Chemistry.* 1999. .

[19] W.Baker, Richard. *Membrane Technology and Applications.* England : McGraw-Hill, 2000.

[20] Warren L.McCabe, Julian C.Smith, Peter Harriott. *Unit Operations of Chemical Engineering.* 1993. 0-07-112738-0.

[21] A. Mersmann, M. Kind, J. Stichlmair: *Thermal Separation Technology. Berlin, Heidelberg: Springer 2011; D.W. Green (ed.): Perry's Chemical Engineers'* Handbook, New York: McGraw-Hill 2008

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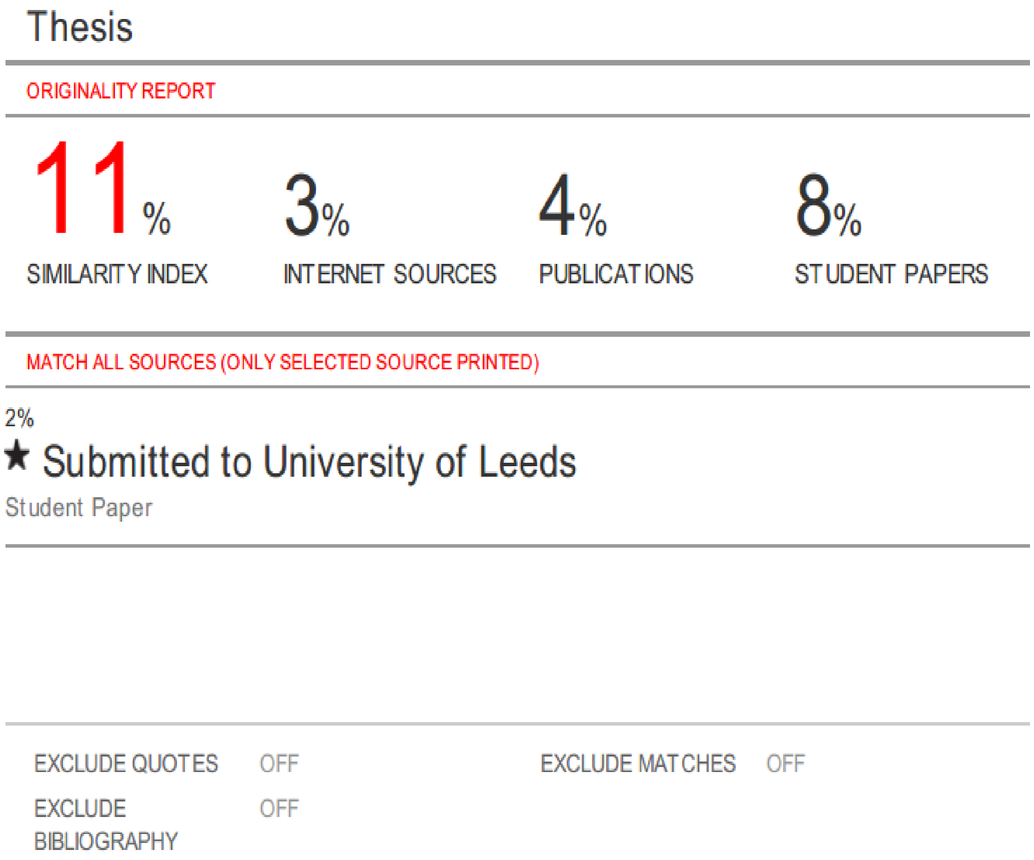

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