

Optimization of Ammonia Recovery Unit at FFC Plant-1



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Optimization of Ammonia Recovery

Unit at FFC Plant-1



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Certificate

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Dedication

To our parents, to our teachers, to our friends, and to SCME♥

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Abstract

In the past 30 years, the chemical industry has been the subject of notable changes due to higher demands, and increasing production and operational costs. Now, there is great competition in product pricing and quality in local and global markets. As a result, it has become imperative for chemical industries to focus on optimization of industrial units and processes. Chemical processes have to be made as efficient as physically possible and within the reasonable constraints. Chemical engineers are determined to find better-operating strategies and processes controls. The current situation of Purge Gas Recovery Section at FFC Plant-1 offers an opportunity for energy optimization of the Ammonia Recovery Unit. The main idea behind the study is to reduce energy consumption for the plant and make a positive contribution towards its profitability.

Therefore, this study focused on the statistical determination of optimum operating parameters for a purge gas Ammonia separating unit at FFC-1. In order to reduce the overall energy requirement of the unit and reduce production cost of NH_3 required for Urea; ultimately increasing profitability for FFC.

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Nomenclature

| | |
|-----------------|---------------------------------|
| 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 |
| d _o | Outer diameter of tubes |
| d _i | Internal diameter of tubes |
| G | Mass Velocity |
| α | Selectivity |
| μ | Viscosity |
| λ | 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 |
| $^{\circ}C$ | Centigrade |
| $^{\circ}K$ | Kelvin |
| P | Pressure |
| ρ | Density |
| KJ/kg | Kilo Joule per kilogram |
| U | Overall Heat Transfer Coefficient |
| Pt | Tube Pitch |
| Ds | Shell Diameter |
| Ib | Baffle Spacing |
| Gs | Mean Velocity |
| Ft | Friction Factor |
| W | Watt |
| H2 | Hydrogen |
| N2 | Nitrogen |
| Re | Reynolds Number |
| ΔP | Pressure Difference |
| PCE | Purchase Cost of Equipment |
| hi | Internal Pipe Heat Transfer Coefficient |
| O.D | Outer diameter |
| As | Cross-Sectional Area |
| Pr | Prandtl Number |

| | |
|--------|--|
| Nu | Nusselt Number |
| de | Equivalent diameter |
| Np | Number of Tube Passes |
| ut | Tube Side Velocity |
| jh | Heat Transfer Factor |
| s | Second |
| do | Tube Outside Diameter |
| Dv | Diffusivity of vapor |
| Z | Height of Column |
| ARU | Ammonia Recovery Unit |
| MTPD | Metric Tons Per Day |
| NMC | Normal Cubic Meter per hour |
| MMSCFD | Metric Million Standard Cubic Feet per day |
| FFC | Fauji Fertilizer Company Ltd. |

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.[4]

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. [2, 18] Purge gas contains Ammonia, Hydrogen and other trace gases. Ammonia can be recovered from purge gas to increase total Ammonia gain. In this way, the process can be made more efficient.

Ammonia Section of Plant-I, FFC Goth Machhi employs Ammonia recovery Unit (ARU) for treating purge gas. The Ammonia Recovery Unit (ARU) is installed to recover the ammonia present in purge and off gas from Ammonia Synthesis Loop. [20]

The recovery of Ammonia from Purge gases is essential to:

1. Keep the overall NO_x emission of the plant within safe environmental regulations ranges (400mg/l) under Punjab Environmental Quality Standards [1]
2. It also results in small increase in production of ammonia (~14.8 MTPD)

1.2 Problem Statement

The ARU at Plant-1 is designed to handle 22,499 Nm³/h considering the future installation of an economically feasible Hydrogen Recovery Unit (HRU). But currently, due to the limitations of the gas burner nozzles, this number of Purges cannot be sent to the boilers. So, the ARU is being operated at a lower purge gas flow rate. Due to such conditions, it is required by FFC to optimize the unit for current purge gas flow rates. Therefore, the steam consumption for the ARU must be lowered keeping

in view of ammonia recovery and environmental constraints.

1.3 Purpose of the Study

The purpose of this project is to address the energy issues because of limited natural gas reserves and to favor ammonia production gain. The minimization of energy required to operate the plant along with better operating strategies was also the focus of study.

The purpose mentioned above was met by a simulations-based study involving chemical process design and optimization. The changes to the current design were not proposed and recalculations for the current case process design were done. After comparison with the installed unit it was suggested to perform a statistical study of operating parameters of the ARU to minimize the energy requirement.

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 Information about Product:

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 $-33.3\text{ }^\circ\text{C}$ and has a freezing point of $-77.7\text{ }^\circ\text{C}$. Its density is about one half the density of air. [2]

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. Alkalis 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.
- Dilute solutions of Ammonia are extensively used as detergents and household cleaners.

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: (McCabe 1993) [5]

- 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 product are formed. When absorption of acid gases in basic solvent occurs, mostly chemical absorption occurs. Chemical absorption is also called reactive absorption.

As suggested by Coulson 1977 for absorption, high pressure and low temperature conditions are required. [6] 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 phenomenon 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, 3]

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 [6] 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 top. High temperature and low pressure are 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. (Richardson-1999) [7]

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 exists. 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. [6]

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 Equipment

2.2.1 Heat Exchanger

It is a device which is used to transfer heat between one or more than one fluid, which are either in direct contact or separated by a wall so that the liquids do not mix with each other. (Kern, 2002) Heat Exchangers are widely used in various fields like natural gas processing, petroleum industry etc.

There are different types of Heat Exchangers, but we will be discussing the shell and tube heat 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 phase 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.

2.2.2 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. [11]

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. [7] (Richardson, 1999)

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 is 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.

2.2.3 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 sub cool the product, so firstly vapors are converted into liquid state and then extra heat is removed before collection of final products. [7]

2.2.4 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 operates 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.

2.3 Optimization of Processes

We need to understand the concept and need for Optimization before we can proceed with it. It is the employment of certain tools or methods to arrive at the most efficient and cost-effective solution to a process problem or a design. A process can constitute of some experimental data or even represented by some mathematical relations. We chose a single performance variable such as maximum profitability or minimum resource requirement. The objective of the optimization is to find the numerical values of certain factors in the process that yield the best value of the performance variable. There's often times a compromise between capital costs and operations cost in process optimization. The process model, its factors and the performance variable constitute what we call an "optimization problem".[4]

Following are the 6 steps that are often followed to analyze, evaluate and solve an optimization problem as referred by Himmelblau (1998):

1. Define the process variables and specific characteristics; by making a list of all the variables involved we have analyzed the problem.
2. Determine clearly what needs to be optimized by specifying a function in terms of the variables enlisted in the first part. This process is called the model formulation.
3. Using mathematical expressions, structure a reasonable model for the design that relates our input output variables for the process. Use the basic principles of engineering (mass and energy balances, thermodynamics and transport). Perform degree of freedom analysis to check feasibility of problem solution.
4. If the size of the problem is too large or beyond scope:
 - a. Segmentalize it down to manageable parts
 - b. Re-define the objective function in a simpler way
5. Apply suitable techniques to the mathematical model of the problem.
6. Perform analysis and examine the results. Often, we have to check the sensitivity of the result to the changes in our assumptions and the coefficients present in the problem statement.

Using this 6-step approach, we can move towards better designs, operating procedures and control systems. We can add value to any product, operation or industry using optimization techniques. [4]

2.3.1 Determination of Optimum Operating Parameters:

Once an equipment or a unit is designed, we cannot make physical changes to it. For a column, its number of trays, diameter, tray spacing and height is fixed, and very few degrees of freedom can be manipulated to reduce the operating cost.[4, 6, 12] In these cases we have to determine the optimum operating parameters such as the temperature, pressures, concentrations and levels, which can be easily implemented at a unit to reduce its operating cost; while resulting in greater economic profitability.

There are number of techniques available for the solution of optimization problems, ranging from programming, analytical methods, computational techniques and iterative approximations. But the choice of methods totally depends on the type of problem at hand. We will be discussing these techniques in detail before choosing one for our objective. [4]

2.3.2 Linear Programming

When the problem presents us linear relationships between different variables, we employ the methods of linear programming to that problem. It is easy and simple to comprehend and apply. An objective function and constraints are determined. Each constraint has a linear relation that can be displayed analytically using straight lines and a feasible region is determined. Various algorithms exist to solve the linear programming equations. Once we find out the critical points of the feasible region, we can find out the response of the performance parameter at these critical points. In this way we can determine the maximum or minimum response of the performance parameter at the respective conditions of the variables. Linear Programming is extremely useful in managing resources and inventories which have direct relations with each other and can be evaluated with ease.

2.3.3 Advanced Calculus

Solving optimization problems using calculus is also very renowned technique. The objective function needs to be specified on differential equations in terms of two or more variables. The differential equation is then optimized using critical point

techniques. The derivative of the equations can tell us more about the trends of the response over a range of the values of the variables. This technique is often employed and suitable for the case of design and dimensions of the equipment for a process where primary equations are variables are present and the performance parameters directly relate to economics of the project.

2.3.4 Approximations

Suitable in the case where both the objective function and variables cannot be described in linear or differential terms easily.[4] Approximations consist of numerical methods to evaluate a problem that cannot be solved by conventional techniques of differential or integrals. We can also employ this technique when we only have the experimental data and need to find out trends and objective functions out of it. This type of analysis, lacking on fixed accurate values, yet is very powerful and can be used on almost all types of problems. [17]

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, the clean gases are sent to the boilers. The whole process is described in detail below:

3.1 Process Flow Diagram:

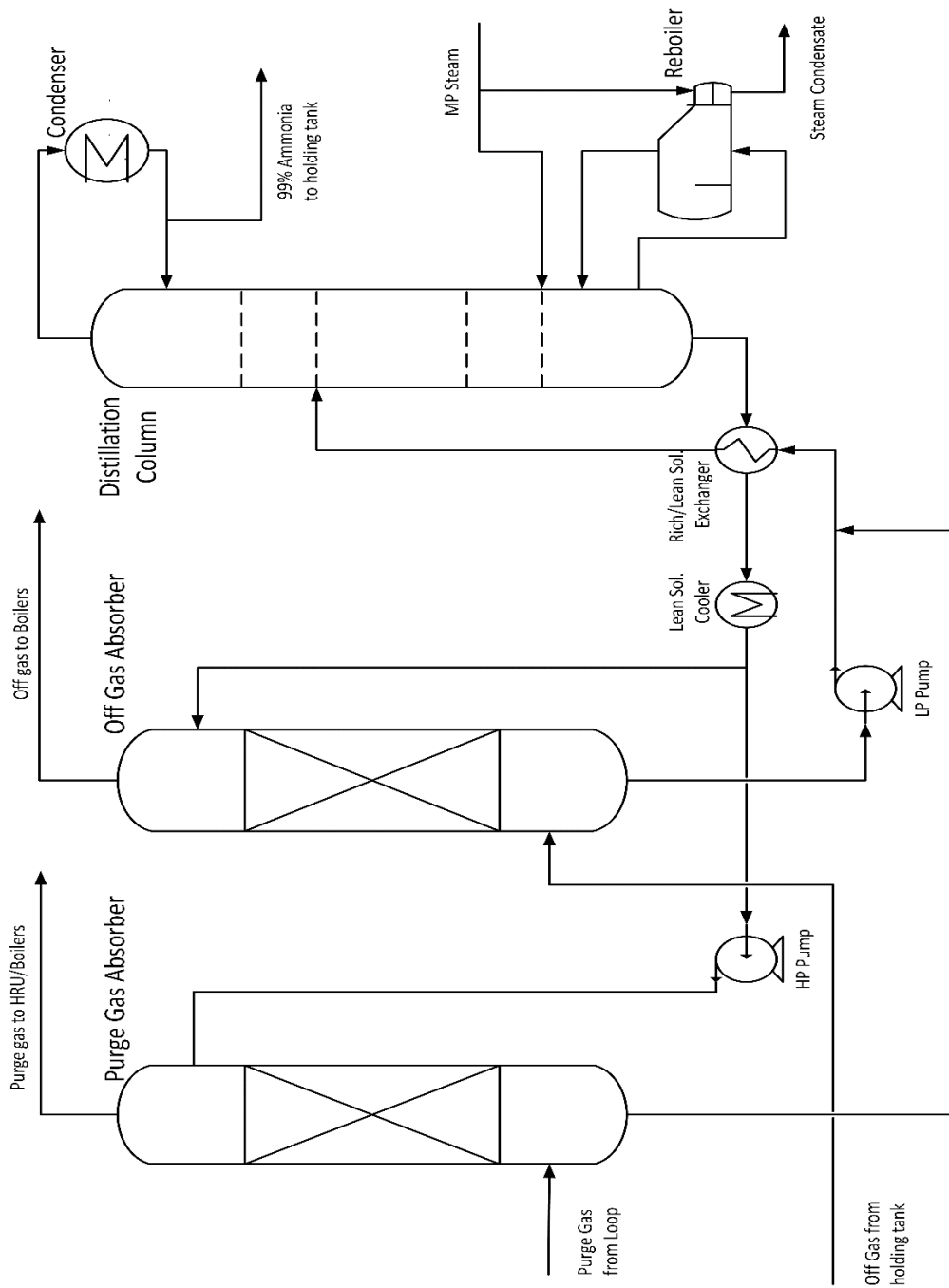


Figure 1 Process Flow Diagram (PFD)

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 °C and 231 bar. Flow of purge gas is at maximum 11,000 Nm³/h. 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 and it leaves from the bottom as rich solution. Rest of the gas mixture i.e. purified purge gas leaves from the top of the column boilers as fuel gas. In the future, the clean purge gases will be sent to the HRU for hydrogen recovery.

3.2.2 Off-Gas Absorber

Off gas from the letdown vessel (vessel in main process) enters the second absorber at 40 °C and 24.1 bar. Flow of off gas is 1066 Nm³/h. 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 °C. 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. About 600kg/h of 99% pure Ammonia is received at the distillate stream. Sent to the Ammonia Let down vessel and to storage.

3.2.4 Rich Solution Exchanger and LP Circulation Pump

For distillation of Rich solution, it must enter the column at high temperature (163 °C) 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 380C at the exit of this heat exchanger. The recovered water is then used in both absorbers as a solvent.

Chapter 4: Material Balance

4.1 Purge Gas Absorber

Purge Gas = P_1

Lean Solution (Water) = W_1

Purified Purge Gas = G_1

Rich Solution (Water + Ammonia) = R_1

Purge Gas (P_1):

Molar Flow (P_1) = $11,000\text{Nm}^3/\text{hr} = 490.76\text{ kgmol/hr}$

Inlet Temperature = $12\text{ }^\circ\text{C}$

Inlet Pressure = $231\text{ kg/cm}^2_g = 22.75\text{ MPa}$

Table 1 Inlet Gas Composition (Absorber 1)

| Components | Compositions | Flowrates (kgmol/hr) |
|-----------------------|--------------|----------------------|
| H ₂ | 59.5% | 292.00 |
| N ₂ | 20.3% | 99.62 |
| Ar | 2.0% | 10.31 |
| CH ₄ | 12.6% | 61.84 |
| NH₃ | 5.6% | 26.99 |

Lean Solution (W₁):

Molar Flow of lean solution = W₁ (kgmol/hr)

Inlet Temperature = 40 °C

Inlet Pressure = 120 kg/cm²_g = 11.87 MPa

Composition of lean solution is assumed as shown below:

Table 2 Inlet Water Composition (Absorber 1)

| Components | Compositions |
|------------------|--------------|
| H ₂ O | 99% |
| NH ₃ | 1% |

H₂, N₂, CH₄ and Ar act as inert and they don't dissolve in water. These components come in same amount in exit gas.

It is assumed that the exit gas contains 0.2% ammonia. Apply component balance on these inert

H₂: H₂ in P₁ = H₂ in G₁

$$0.595 * (490.76 \text{ kgmol/hr}) = \text{H}_2 \text{ in G}_1 = 292.00 \text{ kgmol/hr}$$

N₂: N₂ in P₁ = N₂ in G₁

$$0.203 * (490.76 \text{ kgmol/hr}) = \text{N}_2 \text{ in G}_1 = 99.62 \text{ kgmol/hr}$$

Ar: Ar in P₁ = Ar in G₁

$$0.021 * (490.76 \text{ kgmol/hr}) = \text{Ar in G}_1 = 10.31 \text{ kgmol/hr}$$

CH₄: CH₄ in P₁ = CH₄ in G₁

$$0.126 * (490.76 \text{ kgmol/hr}) = \text{CH}_4 \text{ in G}_1 = 61.84 \text{ kgmol/hr}$$

Overall Balance on G₁: G₁ = N₂ + H₂ + Ar + CH₄ + NH₃

$$G_1 = 292 + 99.62 + 10.31 + 61.84 + 0.002G_1$$

$$G_1 = 463.77 + 0.002G_1$$

$$0.998G_1 = 463.77$$

$$G_1 = 464.70 \text{ kgmol/ hr}$$

Rich Solution (R₁):

We are assuming rich solution compositions to be

Table 3 Outlet Solution Composition (Absorber 1)

| Components | Compositions |
|------------------|--------------|
| H ₂ O | 85% |
| NH ₃ | 15% |

Apply ammonia balance,

NH₃: Ammonia in P₁ + Ammonia in W₁ = Ammonia in G₁ + Ammonia in R₁

$$(0.055 * 490.76) + (0.01 * W_1) = (0.002 * G_1) + (0.15 * R_1)$$

$$26.99 + 0.01 W_1 = 0.929 + 0.15 R_1$$

$$26.06 = 0.15 R_1 - 0.01 W_1$$

Apply water balance,

H₂O: Water in Lean Solution = Water in Rich Solution

$$0.99 W_1 = 0.85 R_1$$

$$W_1 = 0.859 R_1$$

Solving the two balances

$$26.06 = 0.15 R_1 - 0.01 (0.859 R_1)$$

$$26.06 = 0.15 R_1 - 0.00859 R_1$$

$$26.06 = 0.14141 R_1$$

$$R_1 = 187.76 \text{ kgmol/ hr}$$

$$W_1 = 0.859 * 187.76$$

$$W_1 = 161.29 \text{ kgmol/ hr}$$

Also, by applying overall molar flow balance on the absorber,

$$P_1 + W_1 = R_1 + G_1$$

$$W_1 = 187.76 + 464.29 - 490.76$$

$$W_1 = 161.29 \text{ kgmol/ hr}$$

Table 4 Flowrates (Absorber 1)

| Streams | Flowrates (kgmole/hr) |
|---|------------------------------|
| Purge Gas (P₁) | 490.76 |
| Purified Purge Gas (G₁) | 464.29 |
| Lean Solution (W₁) | 161.29 |
| Rich Solution (R₁) | 187.76 |

Table 5 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 Off Gas Absorber

Off Gas = P_2

Lean Solution (Water) = W_2

Purified Off Gas = G_2

Rich Solution (Water + Ammonia) = R_2

Off Gas (P_1):

Molar Flow (P_2) = $1066 \text{ Nm}^3/\text{hr} = 48 \text{ kgmol/hr}$

Inlet Temperature = $40 \text{ }^\circ\text{C}$

Inlet Pressure = $24.1 \text{ kg/cm}^2_{\text{g}} = 2.47 \text{ MPa}$

Table 6 Inlet Gas Composition (Absorber 2)

| Components | Compositions | Flowrates (kgmole/hr) |
|-----------------------|--------------|--------------------------|
| H ₂ | 43 % | 20.64 |
| N ₂ | 17.6% | 8.45 |
| Ar | 2.0% | 0.96 |
| CH ₄ | 20.2% | 9.70 |
| NH₃ | 17.2% | 8.26 |

Lean Solution (W₂):

Molar Flow of lean solution = W₂ (kgmol/hr)

Inlet Temperature = 38 °C

Inlet Pressure = 20 kg/cm²_g = 1.96 MPa

Composition of lean solution is assumed as shown below:

Table 7 Inlet Water Composition (Absorber 2)

| Components | Compositions |
|------------------|--------------|
| H ₂ O | 99% |
| NH ₃ | 1% |

H₂, N₂, CH₄ and Ar act as inert and they don't dissolve in water. These components come in same amount in exit gas.

It is assumed that the exit gas contains 0.1% ammonia. Apply component balance on these inert

H₂: H₂ in P₂ = H₂ in G₂

0.43 * (48 kgmol/hr) = H₂ in G₂ = 20.64 kgmol/ hr

N₂: N₂ in P₂ = N₂ in G₂

0.176 * (48 kgmol/hr) = N₂ in G₂ = 8.45 kgmol/ hr

Ar: Ar in P₂ = Ar in G₂

0.02 * (48 kgmol/hr) = Ar in G₂ = 0.96 kgmol/ hr

CH₄: CH₄ in P₂ = CH₄ in G₂

0.202 * (48 kgmol/hr) = CH₄ in G₂ = 9.70 kgmol/ hr

Overall Balance on G₂: $G_2 = N_2 + H_2 + Ar + CH_4 + NH_3$

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

$$0.999G_2 = 39.744$$

$$G_2 = \mathbf{39.78 \text{ kgmol/ hr}}$$

Rich Solution (R₂):

We are assuming rich solution compositions to be

Table 8 Outlet Solution Composition (Absorber 2)

| Components | Compositions |
|------------------|--------------|
| H ₂ O | 90% |
| NH ₃ | 10% |

Apply ammonia balance,

NH₃: Ammonia in P₂ + Ammonia in W₂ = Ammonia in G₂ + Ammonia in R₂

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

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

$$8.216216 = 0.1 R_2 - 0.01 W_2$$

Apply water balance,

H₂O: Water in Lean Solution = Water in Rich Solution

$$0.99 W_2 = 0.9 R_2$$

$$1.1 W_2 = R_2$$

Solving the two balances

$$8.216216 = 0.1 (1.1 W_2) - 0.01 W_2$$

$$8.216216 = 0.11 W_2 - 0.01 W_2$$

$$W_2 = \mathbf{82.16 \text{ kgmol/ hr}}$$

$$R_2 = 1.1 W_2$$

$$R_2 = \mathbf{90.38 \text{ kgmol/ hr}}$$

Also, by applying overall molar flow balance on the absorber,

$$P_2 + W_2 = R_2 + G_2$$

$$R_2 = 82.16 + 48 - 39.78$$

$$W_2 = \mathbf{90.38 \text{ kgmol/ hr}}$$

Table 9 Flowrates (Absorber 2)

| Streams | Flowrates (kgmol/hr) |
|---|-----------------------------|
| Off Gas (P₂) | 48.00 |
| Purified Off Gas (G₂) | 39.78 |
| Lean Solution (W₂) | 82.16 |
| Rich Solution (R₂) | 90.38 |

Table 10 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

$$R_1 + R_2 = R_3$$

$$187.76 + 90.38216 = R_3$$

$$R_3 = \mathbf{278.4 \text{ kgmol/hr}}$$

$$\text{H}_2\text{O in } R_3 = x$$

Apply water balance,

$$0.85 * R_1 + 0.9 * R_2 = x * R_3$$

$$0.85 * 187.76 + 0.9 * 90.38 = x * 278.4$$

$$\mathbf{H_2O:} \quad x = 0.866$$

$$\mathbf{NH_3:} \quad 1 - x = 1 - 0.866 = 0.134$$

4.4 Distillation Column

Feed of distillation column = F

Top product = D

Bottom Product = B

Vapors coming from top of column = V

Reflux amount = R

Make-up steam = M

Now, Feed of Distillation:

$$F = R_3 = \mathbf{278.4 \text{ kgmol/hr}}$$

$$\text{H}_2\text{O} = 86.6\%, \text{NH}_3 = 13.4\%$$

Assumption:

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

and 1% ammonia.

Apply Ammonia balance,

NH₃: Ammonia in feed = Ammonia in top product + Ammonia in bottom product

$$0.134 * 278.4 = 0.995 * D + 0.01 * B$$

$$37.3 = 0.995D + 0.01B$$

Apply Water balance,

H₂O: Water in feed + Make up Steam = Water in top product + Water in bottom product

$$0.866 * 278.4 + M = 0.005 * D + 0.99 * B$$

As we know,

$$B = W_1 + W_2$$

$$B = 161.29 + 82.16 = \mathbf{243.45 \text{ kgmol/ hr}}$$

From ammonia balance

$$37.3 = 0.995D + (0.01 * 243.45)$$

$$D = \mathbf{35.05 \text{ kgmol/ hr}}$$

By applying overall balance around distillation column,

Feed + Make up Steam = Top Product + Bottom Product

$$\mathbf{Overall:} \quad F + M = D + B$$

$$M = 35.05 + 243.45 - 278.4$$

$$M = \mathbf{0.1 \text{ kgmol/ hr}}$$

Reflux ratio: (2.5)

$R = \text{reflux amount} = (2.5) * D = (2.5) * 35.05 = 87.63 \text{ kgmol/hr}$

Vapors coming out from distillation column = Reflux + Top Product = $D + R = 35.05 + 87.63$

$V = 122.68 \text{ kgmol/hr}$

Table 11 Flowrates (Distillation Column)

| Streams | Flowrates (kgmole/hr) |
|---------------------|------------------------------|
| Feed (F) | 278.4 |
| Top Product (D) | 35.05 |
| Bottom Product (B) | 243.45 |
| Reflux Amount (R) | 87.63 |
| Vapors from top (V) | 122.68 |
| Make up Steam (M) | 0.1 |

Table 12 Compositions (Distillation)

| Components | Feed | Top Product | Bottom Product |
|-------------------|-------------|--------------------|-----------------------|
| NH3 | 0.134 | 0.995 | 0.99 |
| H2O | 0.866 | 0.005 | 0.10 |

Chapter 5: Energy Balance

5.1 Purge Gas Absorber

The outlet streams are at higher temperatures than inlet; this is due to the exothermic nature of absorption of ammonia gas. Applying an energy balance over the system. The heat of solution can be considered to heat the system.

$$\Delta U + \Delta K.E. + \Delta P.E. + P\Delta V = Q + W$$

$$\Delta H = Q$$

$$Q_{abs} = \Delta H_{gas} + \Delta H_l$$

$$\Delta H = m C_p \Delta T$$

For unabsorbed gas, only temperature is increasing, and the average C_p of the gases can be found out by individual C_p of components and compositions.

$$M_{gas} = 464.29 \text{ Kgmol/hr}$$

$$C_p = 31.44 \text{ KJ/kgmol. } ^\circ\text{C}$$

$$\text{Inlet temperature of gas} = 12 \text{ } ^\circ\text{C}$$

$$\text{Outlet temperature of gas} = 43.3 \text{ } ^\circ\text{C}$$

$$Q_g = m C_p \Delta T$$

$$= 464.29 \times 31.44 \times (43.3 - 12)$$

$$= 4.569 \times 10^5 \text{ KJ/hr}$$

For Lean Solution, which is mostly water, we get the sensible change.

$$M_{Lean} = 161.29 \text{ Kgmol/hr}$$

$$C_p = 77.48 \text{ KJ/kgmol. } ^\circ\text{C}$$

$$\text{Inlet temperature of lean solution} = 38.85 \text{ } ^\circ\text{C}$$

$$\text{Outlet temperature of lean solution} = 49 \text{ } ^\circ\text{C}$$

$$\begin{aligned}
 Q_L &= m C_p \Delta T \\
 &= 161.29 \times 77.48 \times (49 - 40) \\
 &= 1.125 \times 10^5 \text{ KJ/hr}
 \end{aligned}$$

$$\begin{aligned}
 Q_{\text{abs}} &= Q_g + Q_L \\
 Q_{\text{abs}} &= 4.569 \times 10^5 + 1.125 \times 10^5 \text{ KJ/hr} \\
 &= 5.694 \times 10^5 \text{ KJ/hr}
 \end{aligned}$$

NH₃ absorbed= 26.06 kgmol/hr

Heat of vaporization of NH₃= 23,300 KJ/kgmol

$$Q_{\text{NH}_3} = 26.06 * 23300 = 607,198 \text{ KJ/hr}$$

5.2 Off Gas Absorber

$$Q_{\text{abs}} = Q_g + Q_L$$

$$Q = m C_p \Delta T$$

For gas,

$$M_{\text{gas}} = 39.78 \text{ kgmol/hr}$$

$$C_p = 31 \text{ kJ/kgmol. } ^\circ\text{C}$$

Inlet temperature of gas = 40 °C

Outlet temperature of gas = 42°C

$$\begin{aligned}
 Q_{\text{gas}} &= m C_p \Delta T \\
 &= 39.78 \times 31 \times (42 - 40) \\
 &= 2466.36 \text{ kJ/hr}
 \end{aligned}$$

For Lean Solution

$$M_L = 82.16 \text{ kgmol/hr}$$

$$C_p = 77.86 \text{ kJ/kgmol } ^\circ\text{C}$$

Inlet temperature of lean solution = $38 \text{ } ^\circ\text{C}$

Outlet temperature of lean solution = $73.7 \text{ } ^\circ\text{C}$

$$\begin{aligned} Q_L &= m C_p \Delta T \\ &= 82.16 \times 77.86 \times (73.7 - 38) \\ &= 2.283 \times 10^5 \text{ KJ/hr} \end{aligned}$$

$$Q_{\text{abs}} = Q_g + Q_L$$

$$\begin{aligned} Q_{\text{abs}} &= 2.283 \times 10^5 + 2466.36 \\ &= 2.308 \times 10^5 \text{ kJ/hr} \end{aligned}$$

NH_3 absorbed = 9.04 kgmol/hr

Heat of ammonia solution = 23350 kJ/hr

$$Q_{\text{NH}_3} = 211,084 \text{ KJ/hr}$$

5.3 Lean Solution Heat Exchanger

Table 13 Heat Exchanger Summary (E-523)

| | Shell Side | Tube Side |
|--------------------|-------------------------|-------------------------|
| Fluid | Lean Solution | Rich Solution |
| Flow Rate | 243.45 kgmol/hr | 278.4 kgmol/hr |
| C _P | 75 kJ/kgmol °C | 79 kJ/kgmol °C |
| Inlet Temperature | T ₁ = 229 °C | t ₁ = 60 °C |
| Outlet Temperature | T ₂ = | t ₂ = 163 °C |

$$\begin{aligned}Q_{\text{tube}} &= m C_p \Delta T \\&= 278.4 \times 79 \times (163 - 60) \\&= 2.27 \times 10^6 \text{ KJ/hr}\end{aligned}$$

Applying energy balance with no work or external heat transfer,

Energy given by shell side = Energy gained by tube side

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

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

$$-2.27 \times 10^6 = 243.45 \times 75 \times (T_2 - 229)$$

$$T_2 = 104.93 \text{ °C}$$

5.4 Lean Solution Cooler

Table 14 Heat Exchanger Summary (E-524)

| | Shell Side | Tube Side |
|--------------------|-------------------|------------------|
| Fluid | Lean Solution | Cooling Water |
| Flow Rate | 405.8825 kgmol/hr | m_{cw} |
| C_p | 75 kJ/kgmol °C | 77.5 kJ/kgmol °C |
| Inlet Temperature | $T_1 = 104.93$ °C | $t_1 = 30$ °C |
| Outlet Temperature | $T_2 = 38$ °C | $t_2 = 60$ °C |

Consider we need to find out the flow rate of utility (cooling water) required.

$$\begin{aligned}
 Q_{\text{shell}} &= m C_p \Delta T \\
 &= 243.45 \times 75 \times (38-104.93) \\
 &= - 1.222 \times 10^5 \text{ kJ/hr}
 \end{aligned}$$

Applying energy balance,

Energy given by shell side = Energy gained by tube side

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

$$Q_{\text{tube}} = m_{cw} C_p \Delta T$$

$$1.222 \times 10^6 = m_{cw} \times 77.5 \times (60-30)$$

$$m_{cw} = 525.62 \text{ kgmol/hr}$$

5.5 Distillation Column

Condenser Duty

Apply energy balance around condenser

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

$$Q_C + H_D + H_R = H_V$$

Table 15 Condenser Energy Balance

| | Distillate | Reflux | Vapors |
|-------------------|----------------|----------------|-----------------|
| Specific Enthalpy | 371.9 kJ/kg | 371.9 kJ/kg | 1595.78 kJ/kg |
| Temperature | 40 °C | 40 °C | 82 °C |
| Flow rate | 35.05 kgmol/hr | 87.63 kgmol/hr | 122.68 kgmol/hr |

$$\text{Energy of the distillate} = H_D = m * \hat{h} = (35.05 \times 17.03) \text{ kg} \times (371.9 \text{ kJ/kg})$$

$$= 2.216 \times 10^5 \text{ kJ}$$

$$\text{Energy of the reflux} = H_R = m * \hat{h} = (87.63 * 17.03) \text{ kg} * (371.9 \text{ kJ/kg})$$

$$= 5.55 \times 10^5 \text{ KJ/hr}$$

$$\text{Energy of vapors} = H_V = m * \hat{h} = (122.68 \times 17.03) \text{ kg} \times (1595.78 \text{ kJ/kg})$$

$$= 3.334 \times 10^6 \text{ kJ/hr}$$

Condenser Duty:

$$Q_C = H_V - H_D - H_L$$

$$Q_c = (3.33 - 0.221 - 0.555) \times 10^6$$

$$Q_c = 2.557 \times 10^6 \text{ kJ/hr}$$

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

$$C_p \text{ of water} = 77.72 \text{ KJ/kgmol. } ^\circ\text{C}$$

Temperature rise for cooling water is limited to 30 °C

$$Q = m C_p \Delta T$$

$$Q = 2.557 \times 10^6 \text{ kJ/hr}$$

$$m_{cw} = \frac{(2.56 \times 10^6)}{77.72 * 30} = 1096.83 \text{ kgmol/hr}$$

Reboiler Duty

Reboiler energy is calculated by applying overall energy balance on the column

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

$$Q_B + H_B = Q_C + H_D + H_B$$

Table 16 Reboiler Energy Balance

| | Distillate | Bottoms | Feed |
|-------------------|----------------|-----------------|----------------|
| Specific Enthalpy | 371.9 kJ/kg | 985 kJ/kg | kJ/kg |
| Temperature | 40 °C | 229 °C | 163 °C |
| Flow rate | 35.05 kgmol/hr | 243.45 kgmol/hr | 278.4 kgmol/hr |

$$\begin{aligned}\text{Energy of the distillate} = H_D &= m \times \hat{h} = (35.05 \times 17.03) \text{ kg} \times (371.9 \text{ kJ/kg}) \\ &= 2.216 \times 10^5 \text{ kJ}\end{aligned}$$

$$\begin{aligned}\text{Energy of the bottoms} = H_B &= m \times \hat{h} = (243.45 \times 18.00) \text{ kg} \times (985 \text{ kJ/kg}) \\ &= 4.316 \times 10^6 \text{ KJ/hr}\end{aligned}$$

$$\begin{aligned}\text{Energy of the feed} = H_F &= m \times \hat{h} = (278.4 \times 17.88) \text{ kg} \times (645.15 \text{ kJ/kg}) \\ &= 3.211 \times 10^6 \text{ KJ/hr}\end{aligned}$$

$$\begin{aligned}Q_B &= Q_c + H_D + H_B - H_F \\ &= (2.56 + 0.22 + 4.32 - 3.21) \times 10^6 \\ &= 3.89 \times 10^6 \text{ kJ/hr}\end{aligned}$$

Amount of steam required for reboiler is calculated as below:

Steam is available at 248 °C,

Latent heat of steam = 31670 kJ/kgmol

$$Q_B = 3,890,000 \text{ kJ/hr}$$

$$\text{Steam required} = 3.89 \times 10^6 / 31670$$

$$\text{Steam required} = 122.83 \text{ kgmol/hr}$$

Chapter 6: Equipment Design

6.1 Purge Gas Absorber C-523:

Liquid flowrate = $m_L = 161.29$ kgmol/h

Gas flowrate = $m_v = 490.76$ kgmol/h

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 = $x_2 = 15$ %

Molar mass of Lean solution: $M_L = 18$ kg/kgmol

Avg. Molar Mass of gas: $M_v = 10.68$ kg/kgmol

Density of gas: $\rho_v = 10.6$ kg/m³

Density of Lean solution: $\rho_L = 55$ kg/m³

$m_L = 1.5 m_{Lm}$

Minimum liquid flowrate: $m_{Lm} = 107.52$ kgmol/h

$$\frac{m_L m}{m_v'} = \frac{y_{2k}' - y_{1k}'}{\frac{y_{2k}'}{K_k} - X_{1k}'}$$

By putting values in above equation, we get

Equilibrium constant = $K_k = 0.214$

Absorption coefficient can be calculated through:

$$A_A = \frac{m_L}{K_k \times m_v} = 1.53$$

And the Number of Equilibrium stages through:

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

Number of equilibrium stages = $N_e = 5.6$

Volumetric flowrate of gas = 11,000 N m³/h = 3.05 N m³/s = 0.0045 m³/s (ideal gas relation)

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

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

$$V_v = K_v \left[\frac{(\rho_L' - \rho_v')}{\rho_v'} \right]^{1/2}$$

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

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

Consider a stage spacing of 0.5m

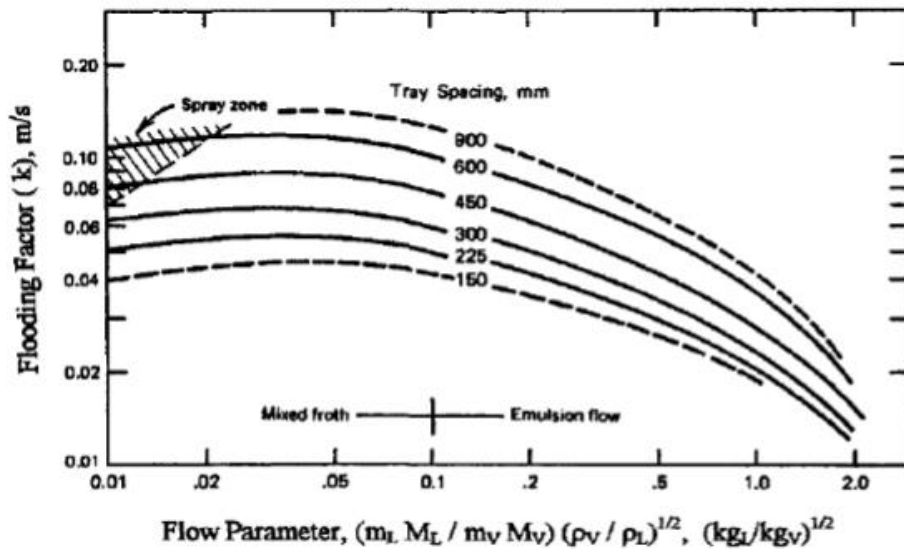


Figure 2 Absorber Flooding Factor Plot (Richardson Vol 6, 1999)

Surface tension = $\sigma = 0.069$ N/m, Flow parameter = 0.243

Value of K_v by using surface tension and $k=0.055$ m/s is :

$$K_v = 0.013 \text{ m/s}$$

$$\text{Gas velocity} = v_v = 0.027 \text{ m/s}$$

$$\text{Now, } A = V_v / v_v = 0.166 \text{ m}^2$$

From area, Diameter = $D = 0.46 \text{ m} = 1.5 \text{ ft}$

Height of column:

$$Z = HTU \times NTU = HTU \times \int_{x_2}^{x_1} \frac{dx}{x_e - x}$$

HTU= Height equivalent of one transfer unit = 0.5 (from table given in Richardson-Coulson)

NTU= Number of transfer units

x_e = Equilibrium NH_3 concentration in liquid (from VLE plot)



Figure 3 $\text{NH}_3\text{-H}_2\text{O}$ VLE plot (@120bar) from HYSYS (AspenTech Eq Unit)

The equilibrium concentration for $x=0.15$ comes out to be 0.05

$$NTU = \int_{0.15}^{0.002} \frac{dx}{0.05 - x}$$

$$NTU = 18.56$$

$$\text{Efficiency of Column} = \frac{Ne}{NTU} = \frac{5.6}{18.56} = 30.2\%$$

Height of the column = $18.56 \times 0.5 = 9.28\text{m}$

Packing Type:

It is a packed column with a packing of SS-316 MellaPak Plus (a type of structured packing). The packing gives extremely controlled pressure drop against a high surface contact area for mass transfer.

The packing unit is 1inch, therefore a suggested HTU of 0.5m

6.2 Off Gas Absorber C-522**Specification:**

Liquid flowrate = $m_L = 82.162 \text{ kgmol/hr}$

Gas flowrate = $m_v = 48 \text{ kgmol/hr}$

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 = $x_2 = 14 \%$

$M_L = 18 \text{ kg/kgmol}$

$M_v = 12.77 \text{ kg/kgmol}$

$\rho_v = 12.38 \text{ kg/m}^3$

$\rho_L = 55 \text{ kg/m}^3$

$m_L = 1.5 m_{L_{\min}}$

Packing Type:

It is a packed column with a packing of CS Pall Rings. The packing is ideal for absorption purposes because of versatility, wettability, and a high resistance to fouling. It can be adaptable to very high loads with low pressure drop and high mass transfer efficiency. Mechanical strength of Pall Rings is also very desirable in cases of high temperature applications.

The packing unit is 1inch, therefore a suggested HTU of 0.5m

Applying the same procedure for the Off-Gas Absorber:

| | |
|----------------------------------|---------------------|
| No. of Equilibrium Stages | 16.8 |
| Packing Material | Carbon Steel (CS) |
| Packing Type | Pall Rings |
| Cross-Sectional Area | 0.187m ² |
| Diameter | 0.48m (1.6ft) |
| HTU | 0.5 |
| NTU | 22.48 |
| Height | 11.24m |
| Efficiency | 74.7% |

6.3 Distillation Column C-521:

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

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

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%

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

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} = Minimum No of Plates ≈ 6

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

$$\frac{N - N_{min}}{N + 1} = 0.75 \left[1 - \left(\frac{R - R_{min}}{R + 1} \right)^{0.5666} \right]$$

Where

N_{min} = Minimum No of Plates

N = Actual/Theoretical No of Plates

R_{min} = Minimum Reflux Ratio = $(X_d - Y_f)/(Y_f - X_f) = 1.13$

R = Actual Reflux Ratio = L/D

$$= 87.63/35.05 = 2.5$$

Putting Values in Gilliland Equation:

N/N_{th} = Actual No of Plates ≈ 10

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

$$N_{eff} = N_{th} \cdot \left(\frac{S}{E_{ov}} \right)$$

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 above equation gives

$N_{\text{eff}} = \text{No of stages/plates to be actually installed} = 20$

Height of Column:

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

$$H = N_{\text{eff}} \cdot \Delta Z$$

H = Height of column

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

$\Delta Z = \text{Spacing between the plates} = 0.55\text{m}$

Substituting the values above give:

H = Height of Column = 10.3 m

Total height is given by:

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

$H_{\text{tot}} = \text{Actual Height of column } N_{\text{gmax}}$

$H_{\text{min}} = \text{Minimum Height (factor for calculating actual height)} = 3.$

$\Delta Z = 1.65 \text{ m}$

So

$H_{\text{tot}} = \text{Actual Height of Column} = 11.34\text{m}$

Diameter of Column

Internal Diameter of column is given by the following equation:

$$D_i = 102.89 \sqrt{\frac{N_{G\text{max}} \cdot T}{P \cdot W_{G\text{zul}}}}$$

$N_{\text{gmax}} = \text{Maximum vapor molar flow rate (kmol/s)} = 122.68 \text{ kmol/hr}$

$W_{\text{g.zul}} = \text{Allowable vapor velocity referred to area} = 0.909 \text{ m/s}$

T (K) = Column Operating Temperature = 493 K

P (Pa) = Column Operating Pressure = 3000000 Pa

For allowable vapor velocity, following equation is used:

$$W_{G\text{zul}} = 0.7 * \frac{A_{ac}}{A_Q} * K_v * \sqrt{\frac{\rho_L - \rho_g}{\rho_g}}$$

Where

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

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

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

ρ_L = Liquid Density (kg/m³) = 815 Kg/m³

ρ_g = Vapor Density (kg/m³) = 25.1 Kg/m³

Substituting the values in above equation, gives:

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

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

$$D_i = \text{Internal Diameter of Column} = 0.62\text{m} = 620\text{mm}$$

6.4 Heat Exchanger E-523 (Rich-Lean Exchanger)

Heat exchanger employed are Shell and Tube Exchanger

| | Shell Side | Tube Side |
|--------------------|---------------------------|------------------------|
| Fluid | Lean Solution | Rich Solution |
| Flow Rate | 243.45 kgmol/hr | 278.4 kgmol/hr |
| C_p | 75 kJ/kgmol °C | 79 kJ/kgmol °C |
| Inlet Temperature | $T_1 = 229 \text{ °C}$ | $t_1 = 60 \text{ °C}$ |
| Outlet Temperature | $T_2 = 104.93 \text{ °C}$ | $t_2 = 163 \text{ °C}$ |

$$\begin{aligned} Q_{\text{shell}} = m C_p \Delta T &= Q_{\text{tube}} = m C_p \Delta T \\ &= 278.4 \times 79 \times (163 - 60) \\ &= 2.27 \times 10^6 \text{ KJ/hr} \end{aligned}$$

$$Q = UA\Delta T_m$$

$$\Delta T_m = F_t * \text{LMTD}$$

$$LMTD = \frac{[(T_1 - t_2) - (T_2 - t_1)]}{\ln \left[\frac{T_1 - t_2}{T_2 - t_1} \right]}$$

Substituting value give result

$$LMTD = 54.79 \text{ } ^\circ\text{C}$$

$$R = (T_1 - T_2) / (t_2 - t_1) = 1.20$$

$$S = (t_2 - t_1) / (T_1 - t_1) = 0.61$$

From figure, using values of S and R, we get F_t

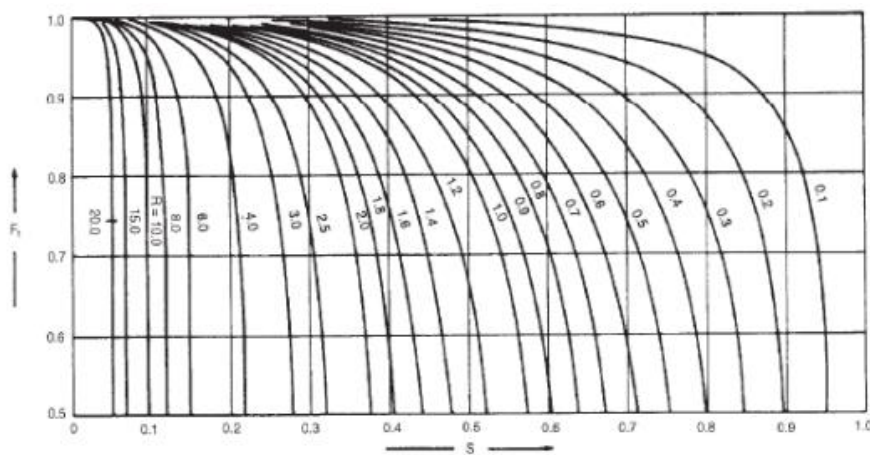


Figure 4 Correction Factor for S&T LMTD (Kern, 2002)

$$\Delta T_m = F_t * LMTD$$

$$\Delta T_m = 0.713 * 77.96 = 55.59 \text{ } ^\circ\text{C}$$

Assumptions:

Choose 20 mm outer diameter, 16 mm inner diameter and 5 m long tubes

Inner diameter of tube = $d_i = 0.016 \text{ m}$

Outer Diameter of tube = $d_o = 0.02 \text{ m}$

Length of tube = $L = 5 \text{ m}$

Area of one tube = $\pi * L * d = \pi * 5 * 0.02 = 0.314 \text{ m}^2$

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

Pitch = $P_t = 1.25 * d_o = 1.25 * 0.02 = 0.025 \text{ m}$

Number of Baffles = Length/Baffle spacing = 20

Tube Side Co-efficient:

$$\text{Mean temperature} = \frac{t_1 + t_2}{2} = 85.75 \text{ }^\circ\text{C}$$

$$\text{Density of solution} = \rho = 888.2 \text{ kg/m}^3$$

$$\text{Viscosity of solution} = \mu = 0.0002723 \text{ kg/m}\cdot\text{sec}$$

$$\text{Heat capacity of solution} = C_p = 4468 \text{ J/kg }^\circ\text{C}$$

$$\text{Thermal conductivity of solution} = K = 0.6182 \text{ W/m}\cdot^\circ\text{C}$$

$$\text{Tube Cross-sectional area} = a = \pi \frac{d_i^2}{4} = (3.14 * 0.016 * 0.016) / 4 = 201 \text{ mm}^2 = 0.000201 \text{ m}^2$$

$$\text{Tubes per pass} = \text{total tubes} / 2 = 82/2 = 41$$

$$\text{Total flow area} = \text{Tube Cross-sectional area} * \text{Tubes per pass}$$

$$\text{Total flow area} = 0.000201 * 41$$

$$\text{Total flow area} = 0.008 \text{ m}^2$$

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

$$\text{Linear velocity} = u_t = (\text{mass velocity} / \text{density}) = 279 / 888.2 = 0.315 \text{ m/sec}$$

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

From graph between Re and Jh factor,

$$\text{it is found that } J_H = 0.0039 \text{ and Prandtl number} = C_p \frac{\mu}{k} = (4468 * 0.0002723) / 0.6182 = 1.968$$

$$h_i = 3094.15 \text{ W/m}^2\cdot^\circ\text{C}$$

Shell Side Coefficient:

$$\text{Mean temperature} = \frac{T_1 + T_2}{2} = 135.25 \text{ }^\circ\text{C}$$

$$\text{Density of solution} = \rho = 912.5 \text{ kg/m}^3$$

$$\text{Viscosity of solution} = \mu = 0.000158 \text{ kg/m}\cdot\text{sec}$$

$$\text{Heat capacity of solution} = C_p = 4511 \text{ J/kg }^\circ\text{C}$$

$$\text{Thermal conductivity of solution} = K = 0.6851 \text{ W/m}\cdot^\circ\text{C}$$

$$\text{Cross flow area} = A_s = (P_t - d_o) \times D_s \times \frac{LB}{P_t} = (0.025 - 0.020) \times 0.3557 \times \frac{0.178}{0.025}$$

$$A_s = 0.01265 \text{ m}^2$$

$$\text{Equivalent diameter} = d_e = 1.10 \frac{P_t^2 - 0.917 d_o^2}{d_o} = 14.2 \text{ mm} = 0.0142 \text{ m}$$

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

$$\text{Linear velocity} = u = (\text{mass velocity} / \text{density}) = 160.34 / 912.5 = 0.175 \text{ m/sec}$$

$$\text{Reynolds Number in tubes} = Re_s = \rho \times d_e \times \frac{\mu_t}{\mu} = 912.5 * 0.0142 * 0.175 / 0.000158 = 14411.63$$

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

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

By putting all values in above equation,

we get $h_s = 2676.701 \text{ W/m}^2 \text{ }^\circ\text{C}$

$$\frac{1}{U_o} = \frac{1}{h_o} + \frac{1}{h_o d} + \left[d_o \frac{\ln d_o}{d_{io}} 2 k_w \right] + \frac{d_o}{d_i} * \frac{1}{h_i d} + \frac{d_o}{d_i} * \frac{1}{h}$$

Fouling Factor for shell side = $h_o d = 5000 \text{ W/m}^2 \text{ }^\circ\text{C}$

Fouling Factor for tube side = $h_i d = 5000 \text{ W/m}^2 \text{ }^\circ\text{C}$

$K_w = 50 \text{ W/m}^2 \text{ }^\circ\text{C}$

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

$U = 786 \text{ W/m}^2 \text{ }^\circ\text{C}$

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

Pressure Drop:

Tube side calculation:

Reynolds number = 16422.5

From graph between Re and J_f ,

Friction factor = $J_f = 0.0039$

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

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

$$\Delta P_t = 3847.54 \text{ N/m}^2 = 3.847 \text{ KPa} = 0.512 \text{ psi}$$

Shell side calculations:

Reynolds number = 14411.6

From graph between Re and J_f ,

Friction factor = $J_f = 0.043$

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

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

$$\Delta P_s = 980 \text{ N/m}^2 = 0.980 \text{ KPa} = 0.142 \text{ psi}$$

Summary of E-523:

| | |
|--|---------------------------|
| Inner Diameter of tube | 16mm |
| Outer diameter of tube | 20mm |
| Tube length | 5m |
| Number of tubes | 82 |
| Shell diameter | 0.3557m |
| Pitch | 0.025m |
| Number of baffles | 20 |
| Baffle spacing | 0.178m |
| Shell side coefficient | 2676 W/m ² .°C |
| Tube side coefficient | 3094 W/m ² .°C |
| Overall heat transfer coefficient | 786 W/m ² .°C |
| Effective Heat Transfer Area | 18.77 m ² |
| Tube side pressure drop | 0.512 psi |
| Shell side pressure drop | 0.142 psi |

6.5 Lean Solution Cooler E-524:

| | Shell Side | Tube Side |
|--------------------|-------------------|------------------|
| Fluid | Lean Solution | Cooling Water |
| Flow Rate | 405.8825 kgmol/hr | 525.62 kgmol/hr |
| C_p | 75 kJ/kgmol °C | 77.5 kJ/kgmol °C |
| Inlet Temperature | $T_1 = 104.93$ °C | $t_1 = 30$ °C |
| Outlet Temperature | $T_2 = 38$ °C | $t_2 = 60$ °C |

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

$$= 243.45 \times 75 \times (38-104.93)$$

$$= - 1.222 \times 10^5 \text{ kJ/hr}$$

Summary of E-524:

For the following specifications we perform the same procedure as for exchanger I and obtain the results as in the table below:

| | |
|---|-----------------------------|
| Inner Diameter of tube | 0.016 m |
| Outer diameter of tube | 0.02 m |
| Tube length | 5 m |
| Shell diameter | 0.315 m |
| Number of tubes | 58 |
| Pitch | 0.025 m |
| Number of baffles | 23 |
| Baffle spacing | 0.157 m |
| Shell side coefficient | 4197.6 W/m ² .°C |
| Tube side coefficient | 5902.8 W/m ² .°C |
| Over all heat transfer coefficient | 803 W/m ² . °C |
| Effective Heat Transfer Area | 28.29 m ² |
| Shell side pressure drop | 0.88 psi |
| Tube side pressure drop | 3.84 psi |

6.6 Reboiler E-521:

| | Shell Side | Tube Side |
|-----------------------|------------------|------------------|
| Fluid | Ammonia Water | Steam |
| Flow Rate | 300 kgmol/hr | 122.83 kgmol/hr |
| C_p | 84.43 KJ/kgmol K | 66.46 KJ/kgmol K |
| Operating Temperature | 229 °C | 248 °C |

Latent heat of steam = 31670 kJ/kgmol

$$Q_B = 3,890,000 \text{ kJ/hr}$$

$$\text{Steam required} = 3.89 \times 10^6 / 31670$$

$$\text{Steam required} = 122.83 \text{ kgmol/hr}$$

Using the manuals of Kern, 2002[7] and Kolmetz, 2013[11], we can apply a similar approach to a double tube pass kettle type reboiler and find out an effective area and number of tubes required.

Summary for E-521

| | |
|---|----------------------------|
| Inner Diameter of tube | 0.016 m |
| Outer diameter of tube | 0.02 m |
| Tube length | 3 m |
| Number of tubes | 97 |
| Pitch | 0.025 m |
| Over all heat transfer coefficient | 4201 W/m ² . °C |
| Effective Heat Transfer Area | 28.6 m ² |
| Shell side pressure drop | 4.7 psi |
| Tube side pressure drop | 0.45 psi |

6.7 Condenser E-522:

| | Shell Side | Tube Side |
|-----------------------|-----------------|------------------|
| Fluid | Ammonia | Cooling Water |
| Flow Rate | 122.68 kgmol/hr | 1096.83 kgmol/hr |
| C_p | 0.66 KJ/kgmol K | 77 KJ/kgmol K |
| Operating Temperature | 62-40 °C | 32-43 °C |

$$Q = m C_p \Delta T$$

$$Q = 2.557 \times 10^6 \text{ kJ/hr}$$

$$m_{cw} = \frac{(2.56 \times 10^6)}{77.72 * 30} = 1096.83 \text{ kgmol/hr}$$

Summary for E-521

| | |
|---|----------------------------|
| Inner Diameter of tube | 0.016 m |
| Outer diameter of tube | 0.02 m |
| Tube length | 5 m |
| Number of tubes | 195 |
| Pitch | 0.025 m |
| Over all heat transfer coefficient | 2514 W/m ² . °C |
| Effective Heat Transfer Area | 20.2 m ² |
| Shell side pressure drop | 2.2 psi |
| Tube side pressure drop | 3.6 psi |

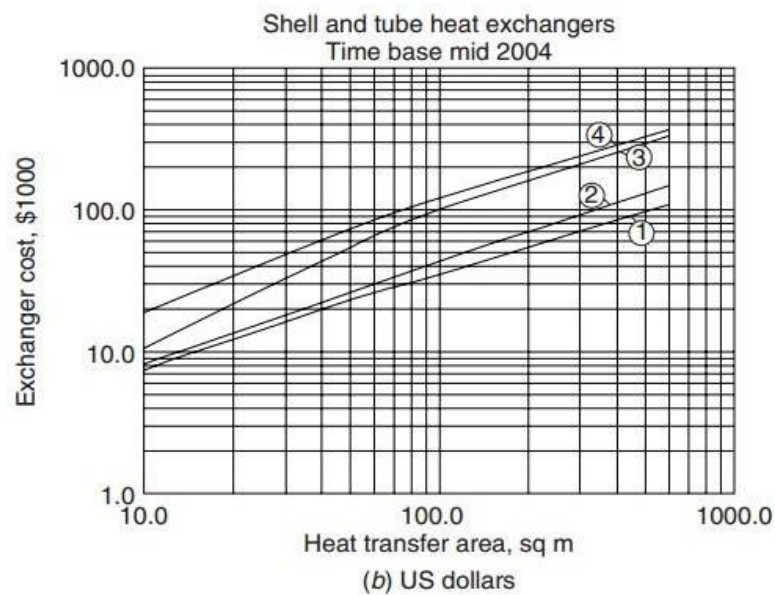
Chapter 7: Costing

Cost Estimates for the Design

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

Graph 1

This graph is used for Heat exchangers cost estimation.

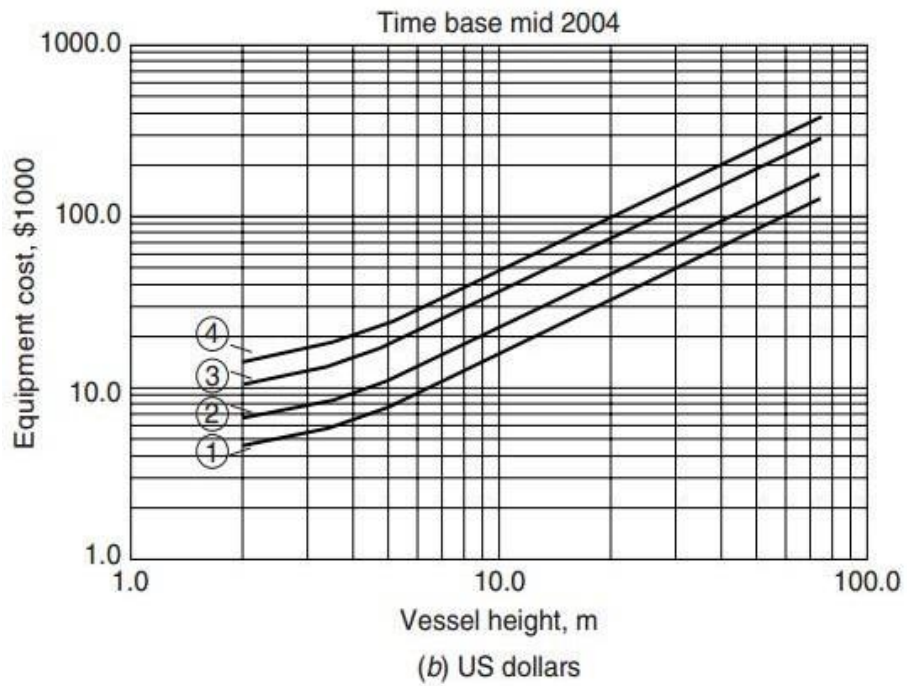


| Materials | | Pressure factors | | Type factors | |
|----------------|-----------------|------------------|--------|------------------|--------|
| Shell | Tubes | | | | |
| ① Carbon steel | Carbon steel | 1-10 bar | × 1.0 | Floating head | × 1.0 |
| ② C.S. | Brass | 10-20 | × 1.1 | Fixed tube sheet | × 0.8 |
| ③ C.S. | Stainless steel | 20-30 | × 1.25 | U tube | × 0.85 |
| ④ S.S. | S.S. | 30-50 | × 1.3 | Kettle | × 1.3 |
| | | 50-70 | × 1.5 | | |

Figure 5 S&T Heat Exchanger Cost Graph (Coulson Vol 6, 1999)

Graph 2

Cost of columns is estimated using this graph.



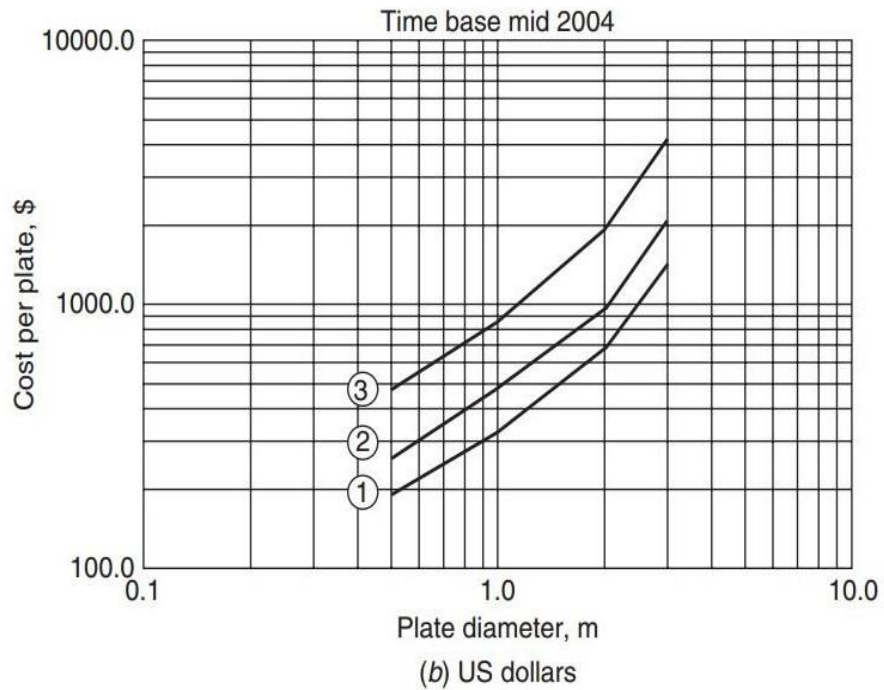
| Diameter, m | | Material factors | Pressure factors |
|-------------|-------|------------------|------------------|
| ①—0.5 | ③—2.0 | C.S. × 1.0 | 1–5 bar × 1.0 |
| ②—1.0 | ④—3.0 | S.S. × 2.0 | 5–10 × 1.1 |
| | | Monel × 3.4 | 10–20 × 1.2 |
| | | S.S. clad × 1.5 | 20–30 × 1.4 |
| | | Monel × 2.1 | 30–40 × 1.6 |
| | | clad | 40–50 × 1.8 |
| | | | 50–60 × 2.2 |

Temperature up to 300°C

Figure 6 Pressure Vessel Cost Graph (Coulson Vol 6, 1999)

Graph 3

Cost of plates is estimated using this graph.



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

Figure 7 Column Plate Cost Graph (Coulson Vol 6, 1999)

Table 17 Column Packing Cost (Coulson Vol 6, 1999)

| Size, mm | Cost | | |
|-----------------------------|------------------|----------------------|------------|
| | £/m ³ | (\$/m ³) | |
| | 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.

| 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 | $\times 1.7$ for ss |
| Compressors | | | | | | |
| Centrifugal | driver | 20-500 | 1160 | 1920 | 0.8 | electric, |
| | power, kW | | | | | max. press. |
| Reciprocating | | | 1600 | 2700 | 0.8 | 50 bar |
| Conveyors | | | | | | |
| Belt | length, m | 2-40 | | | | |
| 0.5 m wide | | | 1200 | 1900 | 0.75 | |
| 1.0 m wide | | | 1800 | 2900 | 0.75 | |
| Crushers | | | | | | |
| Cone | t/h | 20-200 | 2300 | 3800 | 0.85 | |
| Pulverisers | kg/h | | 2000 | 3400 | 0.35 | |
| Dryers | | | | | | |
| Rotary | area, m ² | 5-30 | 21,000 | 35,000 | 0.45 | direct |
| Pan | | 2-10 | 4700 | 7700 | 0.35 | gas fired |
| Evaporators | | | | | | |
| Vertical tube | area, m ² | 10-100 | 12,000 | 20,000 | 0.53 | carbon steel |
| Falling film | | | 6500 | 10,000 | 0.52 | |
| Filters | | | | | | |
| Plate and frame | area, m ² | 5-50 | 5400 | 8800 | 0.6 | cast iron |
| Vacuum drum | | 1-10 | 21,000 | 34,000 | 0.6 | carbon steel |
| Furnaces | | | | | | |
| Process | | | | | | |
| Cylindrical | heat abs, kW | 10^3-10^4 | 330 | 540 | 0.77 | carbon steel |
| Box | | 10^3-10^5 | 340 | 560 | 0.77 | $\times 2.0$ ss |
| Reactors | | | | | | |
| Jacketed, agitated | capacity, m ³ | 3-30 | 9300 | 15,000 | 0.40 | carbon steel |
| | | | 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 | $\times 2$ for |
| cone roof | | 50-8000 | 1400 | 2300 | 0.55 | stainless |

Table 18 Other Equipment Cost Estimates (Coulson Vol 6, 1999)

Rich Solution Exchanger (E-523)

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 Total Cost = \$ 16250

Lean Solution Cooler (E-524)

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 Total Cost = \$ 20625

Condenser (E-522)

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 Total Cost = \$ 17125

Reboiler (E-521)

Shell and tube heat exchanger of

kettle type. Heat Transfer Area =

28.6 m²

Bare Equipment cost = \$ 16750 (Using Graph

1) Pressure factor = 1.25

Type factor = 1.3

Cost = Bare equipment cost * Pressure factor * Type factor Cost = \$ 34,125

Purge Gas Absorber (C-523)

Height of Column = 9.28 m Column's Diameter = 0.600 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

Cost of Packing = \$3620

Total cost of absorption column = \$ 35,800

Off Gas Absorber (C-522)

Height of Column = 11.24 m Column's Diameter = 0.48 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

Cost of Packing = \$ 6180

Total cost of absorption column = \$ 12900

Distillation Column (C-521)

Height of Column = 11.34 m Column's Diameter = 0.62 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 = \$ 36295

Total purchase cost of major equipment items (PCE) = \$ 184,820

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) = \$ 443,568

f10 Design and Engineering 0.30

f11 Contractor's Fee none (unlikely to be used)

f12 Contingencies 0.10

Fixed capital = 439850 (1 + 0.3 + 0.1) = \$ 620,995

Investment Needed for Plant (rounded) = \$ 620,000

Working capital (5% of fixed capital) = \$ 31,050

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

Total Investment Required = \$ 652,000

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 = \$ 47,819

2. Steam required = 71.5 kgmol/hr = 1287

kg/hr = 1.287 t/hr Steam cost = 1.287t * 12\$/t =

\$ 22,857/hr

Variable cost = TVC = \$ 70,676

Fixed Costs

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

4. Operating labor, minimal required for operation, say \$3,000 /year

5. Supervision, no additional supervision would be needed

6. Plant overheads (50% of operating labor) = C8 = \$ 1500
7. Laboratory (take as 30% of operating labor) = C9 = \$ 900
8. Capital charges (6% of fixed capital) = C10 = \$ 36947
9. Insurance (1% of fixed capital) = C11 = \$ 6158
10. Local taxes are neglected
11. Royalties not applicable

$$\text{Total fixed cost} = \text{C5} + \text{C6} + \text{C8} + \text{C9} + \text{C10} + \text{C11} = \text{TFC} = \$ 79,294$$

12. Sales expense not applicable
13. General overheads not applicable
14. Research and development not applicable

$$\text{Annual operating cost, rounded} = \text{TVC} + \text{TFC} = \$ 150,000$$

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)

- 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 8 HYSYS 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-Plöcker, MBWR, **Peng- Robinson**, PR-Twu, PRSV, Sour SRK, Sour PR, SRK, SRK-Twu, or Zudkevitch-Joffe

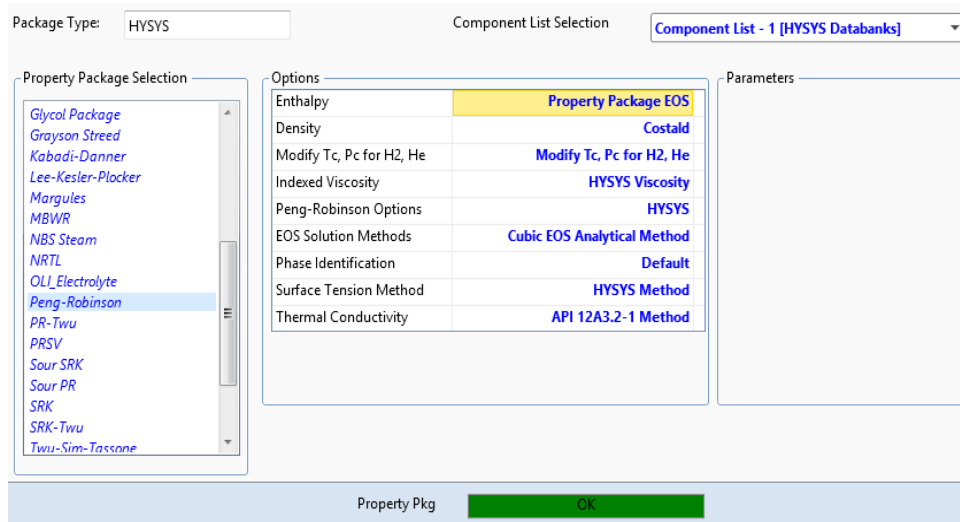


Figure 9 HYSYS Property Package

8.2 Absorber 1

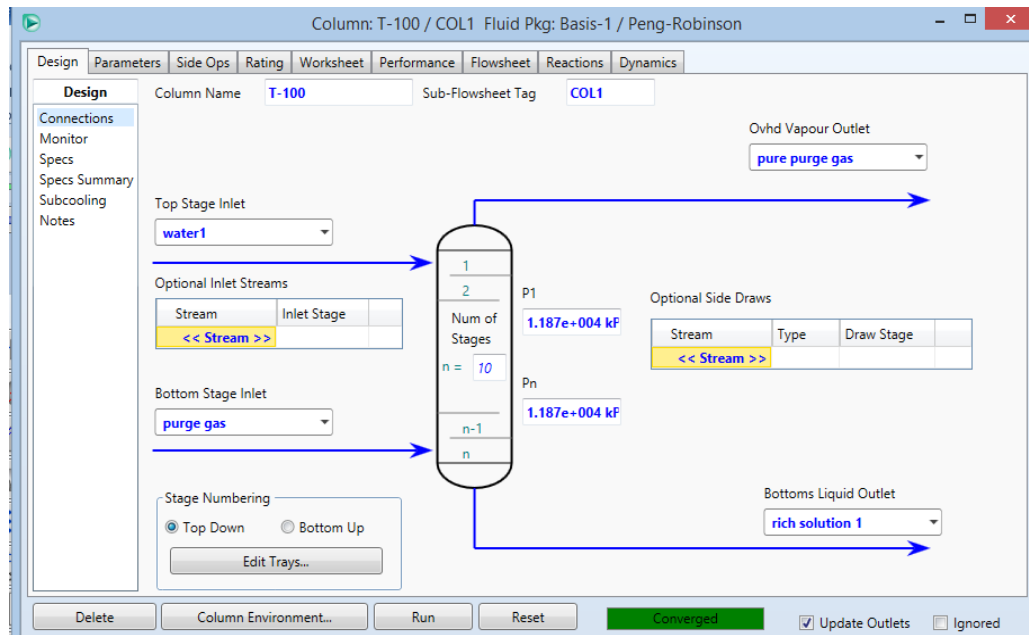


Figure 10 Simulation of Absorber 1

8.3 Absorber 2

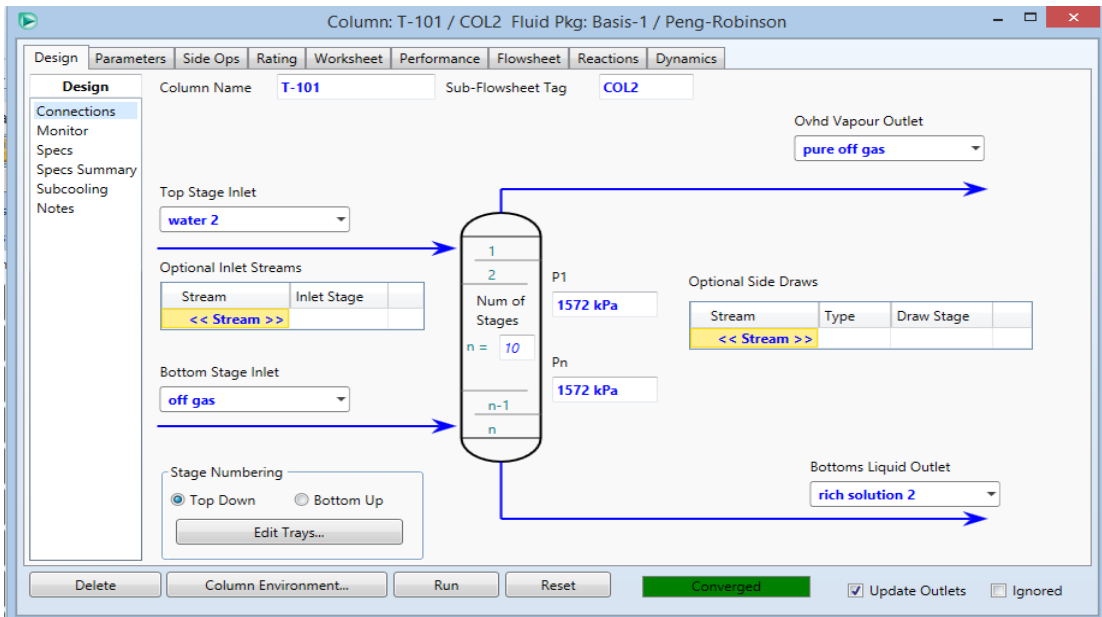


Figure 11 Simulation of Absorber 2

8.4 Distillation

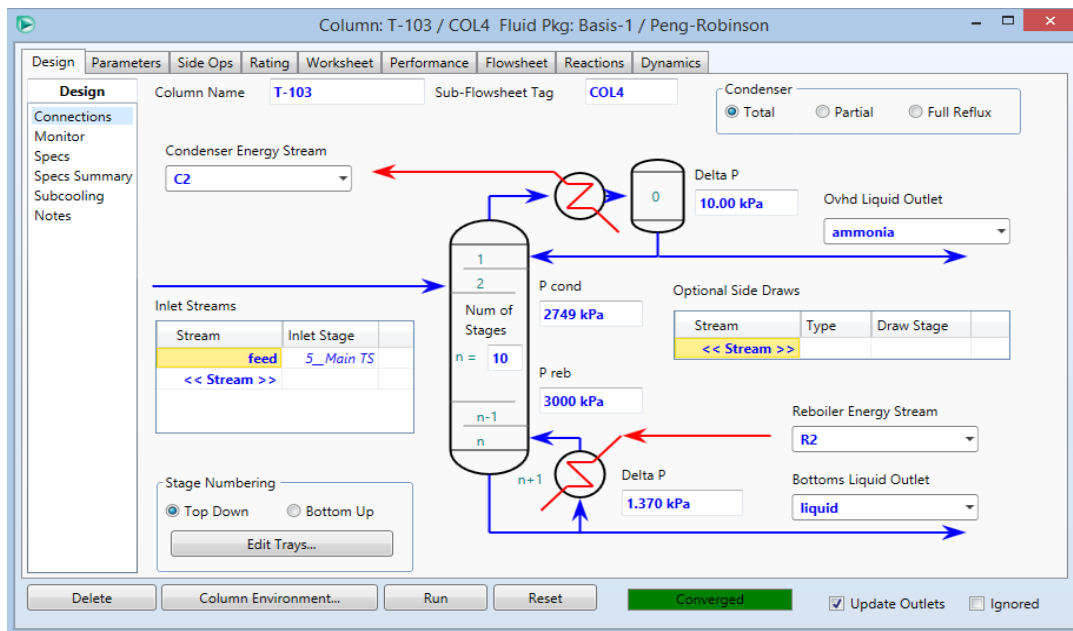


Figure 12 Simulation of Distillation Column

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:

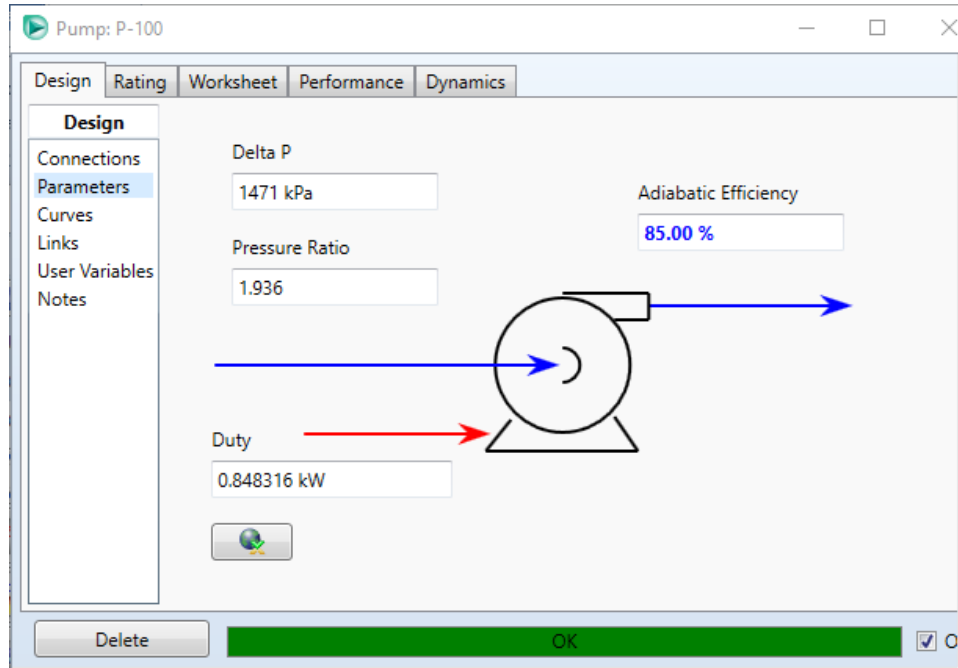


Figure 13 Simulation of Pump 1

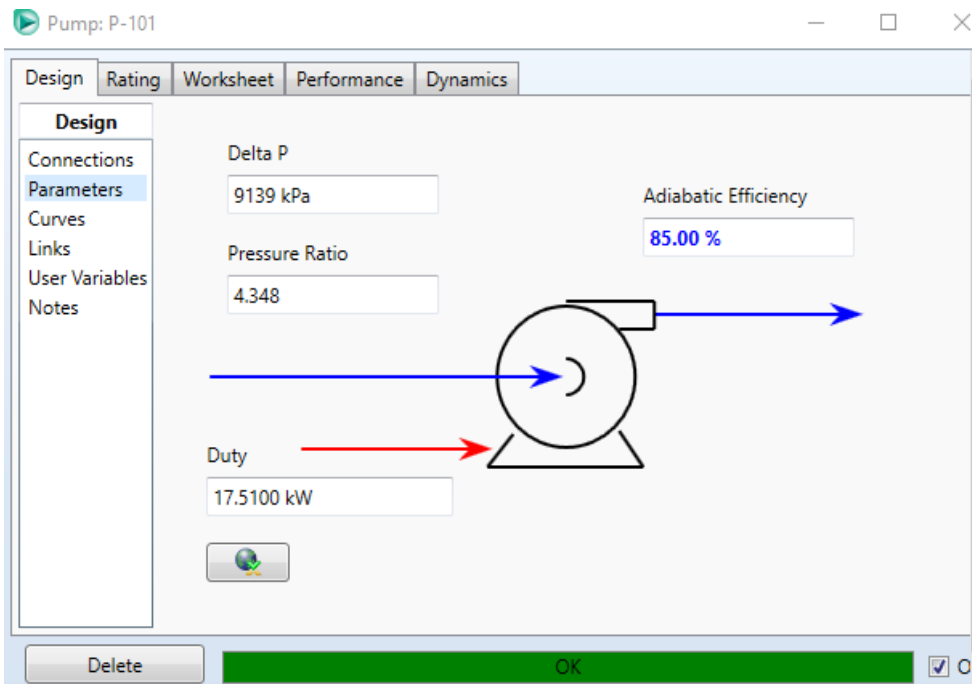


Figure 14 Simulation of Pump 2

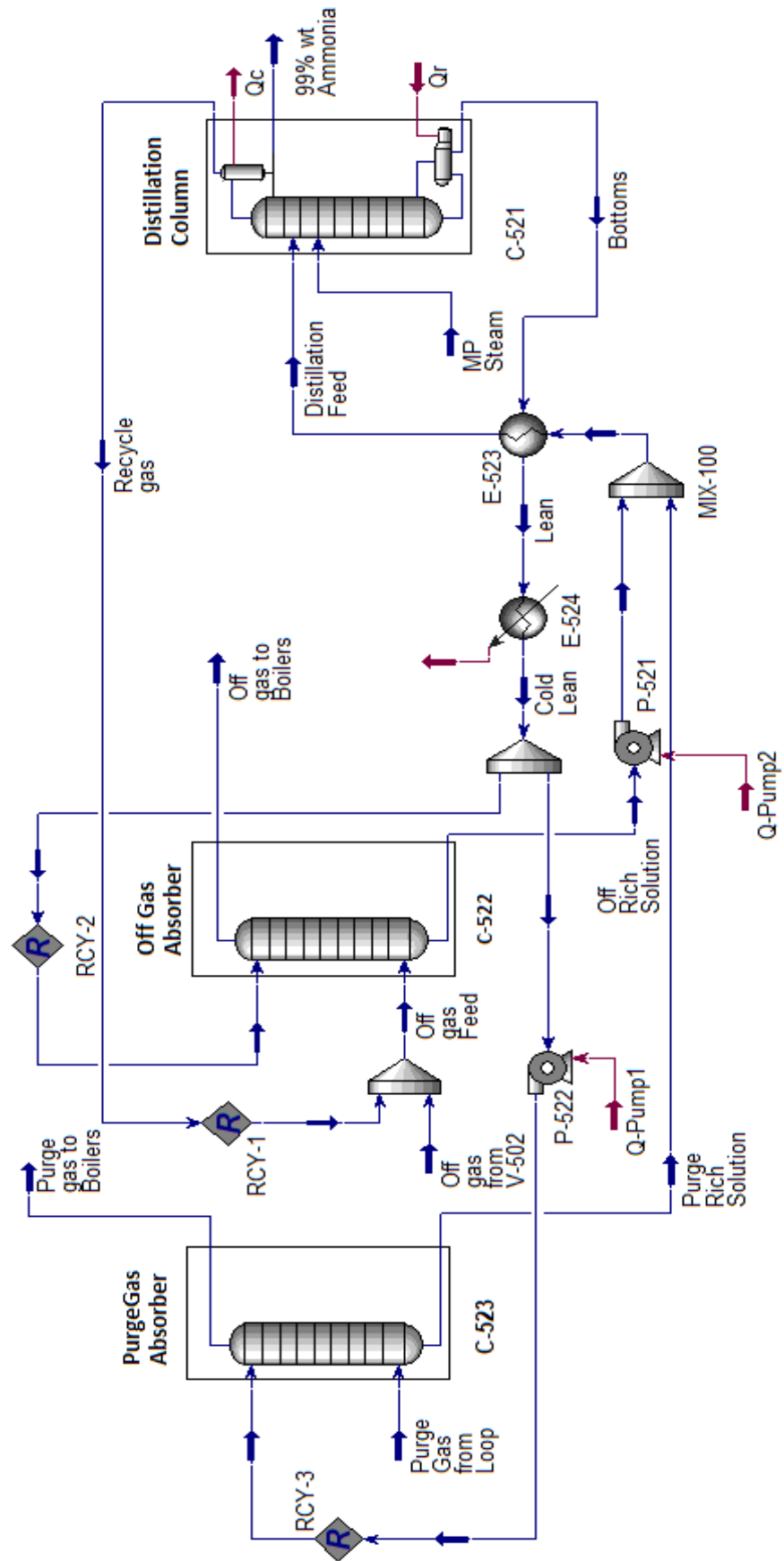


Figure 15 Aspen HYSYS Simulation Model

Chapter 9: Hazard and Operability (HAZOP) Study

Equipment: Absorber

| Guide Word | Deviation | Cause | Consequence & Action |
|--|-----------|-------------------------------|--|
| Line 1: Intentions 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 | | | |
| Less | Flow | Reflux Control Valve Failure | Ammonia product is not purified and contains water as well. Composition analyzer must be used for product. |
| More | Flow | Reflux Control Valve Failure | Required product purification is not achieved. Composition analyzer must be used for product. |
| No | 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 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.[20] 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.[20]

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. (Nitsche, 2017) [12]

Make up steam is added in the distillation column as some water is lost along with distillate (ammonia product).[20] 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.

Chapter 11: Energy Optimization

After obtaining complete design and operating parameters of the plant. We can work towards an optimized solution that could possibly optimize the reboiler duty for the Distillation Column C-523. The target was to achieve a compromise of ammonia production and steam requirement for maximized profitability and energy savings. There are many statistical techniques that could work on this situation but would require extensive multi-dimensional equations and solving derivative equations for functions of ammonia production and reboiler steam duty. For which we need to evaluate the various variables that comprise the numerical function of these responses. Another approach is the factorial optimization that is the essence of Response Surface Methodology (RSM). It is a statistical optimization technique that targets an experimental response with the simultaneous variation of multiple factors. RSM gives us an accurate approximation of optimum factor values that optimize a response.[4] We have employed the technique for the optimization of reboiler duty for the distillation column of ARU.

11.1 Design of Experiments:

Design of experiments, referred to as DOE, is a systematic approach to understanding how process and product parameters affect response variables. The experiments suggest which process variables are contribute to changes in the response variable. For DOE, we have used the software of Design Expert 11, in which we have to design a case where different process variables of the ARU are selected as independent factors.

11.2 Aspen HYSYS Simulation:

Rigorous column modelling and simulation using Aspen HYSYS were done after obtaining preliminary design and material balance. The Shortcut column Analysis was done to verify the results of the preliminary design and get an idea of the favorable parameters for the distillation column such as the minimum reflux ratio, minimum number of stages, optimum feed point entry and optimum reflux ratio.[14]

The feed is specified through the material balance calculations. The internals of the column are specified through the pressure temperature data. The key component

fractions at top and bottom calculated through material balance of outlet streams along with purity and recovery allowances.

We obtained the following results through the Shortcut Analysis.

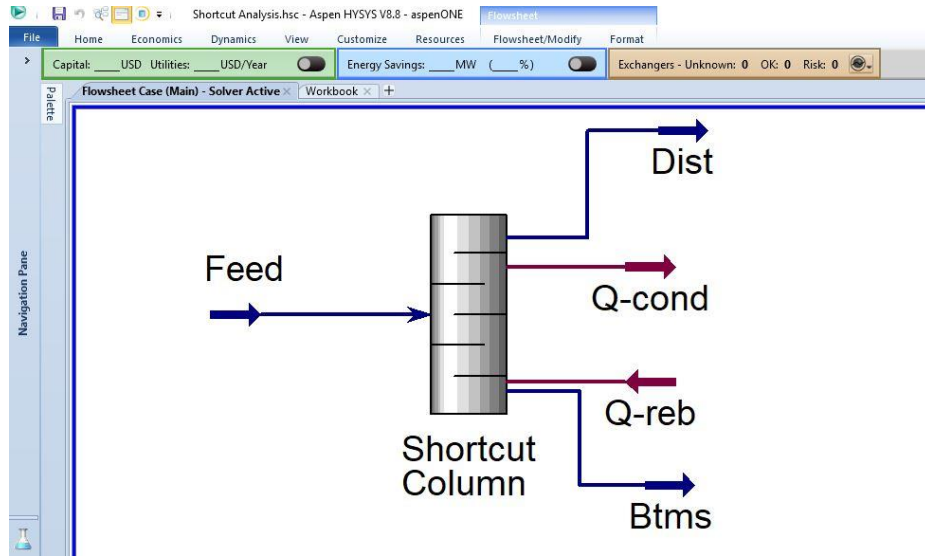


Figure 16 Shortcut Column Simulation

The optimum Reflux ratio as suggested by Richardson Coulson for economic column erection and operation is at most 1.5 times the minimum Reflux ratio.

| Components | |
|-------------------------|----------------|
| Light Key in Bottoms | Ammonia 0.0015 |
| Heavy Key in Distillate | H2O 0.0085 |

| Pressures | |
|--------------------|--------------|
| Condenser Pressure | 2749.121 kPa |
| Reboiler Pressure | 2749.121 kPa |

| Reflux Ratios | |
|-----------------------|-------|
| External Reflux Ratio | 1.700 |
| Minimum Reflux Ratio | 1.140 |

Figure 17 Shortcut Column Results

Based on these results we try to perform the rigorous simulation of the Column C-521 in HYSYS as observed in the Simulation Chapter. Reboiler duty at design case was evaluated which was to be optimized using a DOE and Simulation based approach.

11.3 Response Surface Methodology:

Using the DOE software Design Expert, a 3 factor RSM was performed against the reboiler duty as response of the experiments. The factors selected for the experiments were the following:

1. Reflux Ratio of Column (continuous)
2. Feed Composition NH₃ (continuous)
3. Feed Point plate from bottom (discrete)

Following were the ranges for the three factors:

Table 19 RSM Factor Ranges

| Factor | Lower Limit | Upper Limit |
|-------------------------|--------------------|--------------------|
| Reflux Ratio | 1.14 | 3.0 |
| Feed Composition | 0.10 | 0.40 |
| Feed Point | 7 | 13 |

According to the ranges defined into the software, optimization requires the responses on various experiments or runs of the factors. These factors are then run in Aspen HYSYS to evaluate the reboiler duty. The responses for 25 cases as required by Design Expert were recorded and fed to the software. The software analyzed the data and recommended the calculations mode for the RSM analysis. After complete analysis, we are provided with 3D graph for the multi factor optimization.

Experiments Summary from Design Expert:

Table 20 RSM Run Results

| RUN | Factor 1: Reflux Ratio | Factor 2: Feed Position | Factor 3: Composition NH₃ | Response Reboiler Duty Gcal/h |
|------------|-------------------------------|--------------------------------|---|--------------------------------------|
| 1 | 3 | 7 | 0.4 | 0.5895 |
| 2 | 1.15 | 10 | 0.290267 | 0.4121 |
| 3 | 1.15 | 13 | 0.1 | Unconverged |
| 4 | 1.15 | 7 | 0.1 | Unconverged |
| 5 | 2.1675 | 10 | 0.1 | 1.132 |

| | | | | |
|-----------|---------|----|----------|-------------|
| 6 | 1.779 | 13 | 0.4 | 0.3782 |
| 7 | 1.15 | 13 | 0.4 | 0.2718 |
| 8 | 2.334 | 10 | 0.292 | 0.6083 |
| 9 | 3 | 13 | 0.4 | Unconverged |
| 10 | 1.7975 | 13 | 0.2125 | 0.6402 |
| 11 | 3 | 7 | 0.175 | 0.9024 |
| 12 | 2.334 | 10 | 0.292 | 0.6083 |
| 13 | 1.15 | 10 | 0.290267 | 0.4112 |
| 14 | 1.15 | 9 | 0.2695 | 0.449 |
| 15 | 3 | 12 | 0.186208 | 0.8831 |
| 16 | 3 | 11 | 0.208802 | 0.8496 |
| 17 | 2.1675 | 10 | 0.1 | 1.132 |
| 18 | 1.65875 | 7 | 0.175 | 0.6736 |
| 19 | 1.6125 | 7 | 0.4 | 0.3503 |
| 20 | 2.001 | 11 | 0.4 | 0.4157 |
| 21 | 1.7975 | 13 | 0.2125 | 0.6402 |
| 22 | 3 | 13 | 0.1 | 1.323 |
| 23 | 1.7975 | 13 | 0.2125 | 0.6402 |
| 24 | 1.63822 | 7 | 0.4 | 0.3547 |
| 25 | 3 | 8 | 0.4 | 0.5838 |

11.4 Optimization Results

The results for the RSM analysis on Design Expert 11 represented in a 3D-plot

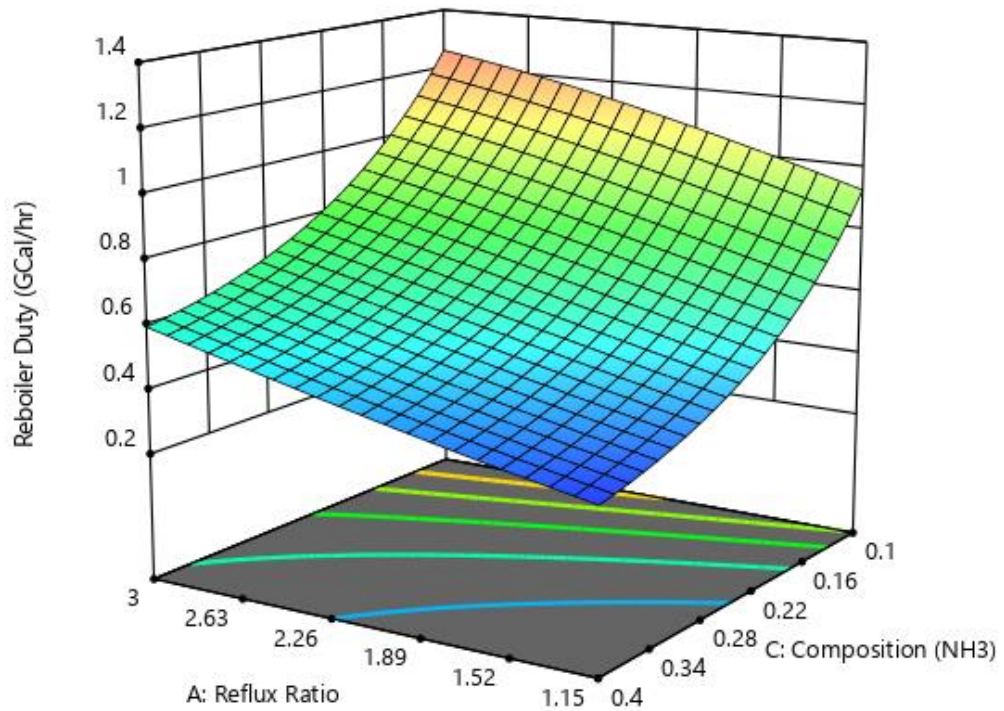


Figure 18 RSM 3D Surface Plot

It was possible to incur the following conclusions from the 3D response surface:

1. Reboiler Duty is a direct function of the reflux ratio. Keeping the reflux as low as possible is required to lower the Steam consumption; while considering the same level of purity and capacity is achieved verified through simulations.
2. Keeping the NH_3 concentration in feed a maximum to achieve better separation and a lower value of reboiler duty for the same performance.
3. Introducing the feed at a higher plate to match the concentration of the tray.

The following conditions for the 3 factors was achieved at which the HYSYS simulation would give an optimum reboiler duty, while keeping the same level of production and purity and while keeping the exit purge gases containing essentially no Ammonia to result in NO_x

| Factor | Optimized Solution |
|-------------------------|---------------------------|
| Reflux Ratio | 1.7 |
| Feed Composition | 0.25 |
| Feed Point | 13 |

Through the analysis performed the ARU reboiler duty was statistically reduced to 40% of the designed case requirements. Which would result in a considerable amount of savings in terms of operating costs.

Table 21 ARU Duties & Steam Requirements

| Purge Gas Flowrates | Reboiler Duty (Gcal/h) | Steam requirement (kg/h) |
|--|-------------------------------|---------------------------------|
| Design Case (22499 Nm³/h) | 1.48 | 3650 |
| Current Case (11000 Nm³/h) | 0.95 | 2354 |
| Optimized case | 0.577 | 1418 |

Considering an industrial quoted rate of MP steam generation from natural gas at Rs. 2,550 per ton. We can calculate the savings as;

- Steam cost before: PKR 53.2 M / year
- Steam cost after: PKR 32.0 M / year
- **Total Savings: PKR 21.2 Million/year**

Conclusion

Using rigorous HYSYS simulations and 3-factor Response Surface Methodology, the ARU's steam requirement was optimized to 1.42t/h which is a 40% reduction in energy requirement. The optimized parameters recommended in this study neither require any investment nor requires any changes made to the design of the currently installed ARU at Plant-1. This would mean, that apart from the recovered Ammonia that is contributing towards the extra earnings, the savings in the form of lesser energy requirements are also contributing towards profitability. The project is completely suggestive of an economically beneficial, efficient, safe and absolutely sustainable solution to the underlying industrial problem; which is the essence of Chemical Engineering.

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