Recovery of Free Ammonia from Dirty Cooling Water at Effluent Treatment Plant

By

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Certificate

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Dedication

This project is dedicated to our parents, without whom we would not have been able to achieve all that we have achieved today. Their dedication and hard work to provide us with the best of facilities is the most commendable and we shall forever be grateful.

This project is also dedicated to SCME and our teachers who have worked tirelessly to provide us with the best education which made us capable of working towards this achievement.

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Nomenclature

C_p	Heat Capacity
ΔT_m	Log Mean Temperature Difference
m	Meter
mm	Millimeter
cm	Centimeter
m^2	Meter square
m^3	Meter cube
kJ	Kilo Joule
kg	Kilogram
h	Hour
gmol	gram mole
L	Length
do	Outer diameter of tubes
di	Internal diameter of tubes
G	Mass Flow Rate
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
°C	Centigrade
K	Kelvin
P	Pressure
ρ	Density
kJ/kg	Kilo Joule per kilogram
U	Overall Heat Transfer Coefficient
D_s	Shell Diameter
I_b	Baffle Spacing
F_t	Friction Factor
H_2	Hydrogen
Re	Reynolds Number
ΔP	Pressure Difference
PCE	Purchase Cost of Equipment

h_i	Internal Pipe Heat Transfer Coefficient
O.D	Outer diameter
Pr	Prandtl Number
Nu	Nusselt Number
d_e	Equivalent diameter
do	Tube Outside Diameter
F_{LG}	Flow Parameter
Y	Capacity Parameter
F_ρ	Density Parameter
F_η	Viscosity Parameter

Abstract

Ammonia is a highly irritating gas which is corrosive in nature and causes perilous effects both on environment and human health if its amount increases the safe limit. Upon contact with humans by either inhalation or ingestion, it causes damage to skin, eyes and the respiratory system. It causes eutrophication, soil acidification and disturbs the eco-system if discarded in environment. Therefore, its removal from the effluent water streams of all industries is a pressing need.

Our aim is to reduce the amount of Ammonia from 800 to 20 ppm in the effluent water stream of NP Plant of Pak-Arab Fertilizers. By doing so, the effluent water composition matches the standards set by the Environment Protection Agency. Firstly, the dissolved Phosphate and Fluoride impurities are removed by the addition of CaO in water, settling of hence formed insoluble salts takes place in a set of clarifiers, sludge thickener and filter press are installed to dewater the sludge and obtain a solid cake. A heat exchanger is employed to increase the temperature of supernatant water stream, in order to meet the desired stripper conditions. An air stripper, with counter-current flow type, reduces the amount of ammonia in effluent water to 20 ppm. After that, the stripped ammonia moves to an absorber column in which Sulphuric acid is used as an absorbent, resulting in a lean solution of Ammonium Sulphate. The effluent air stream from absorber column has 12 ppm of Ammonia. Hence, by the use of this Ammonia removal and recovery process the amount of ammonia in both effluent air and water streams meets the restriction of EPA and falls well within the safe limits.

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1 Introduction

1.1 Background

Fertilizer industry is an integral part of Pakistan's economy. It is estimated that fertilizers industry accounts for approximately 1.2% of worlds energy, thus there is no room for compromising on efficiency of production keeping in consideration the increasing energy crisis and availability of limited fuel resources. The products of Fertilizer Industry generally include ammonia and urea whereas the basic raw materials required are hydrocarbon fuel, air and steam. Nitrogen is taken from Air whereas Hydrogen and Carbon dioxide are obtained by reforming reaction of Hydrocarbons using steam. In the fertilizer industry, once the process has been completed, some ammonia, along with other reacting components, is removed with the effluent streams. These impurities are then processed and removed in the effluent treatment plant. However, some industries have overlooked the EPA standard of 20 ppm ammonia limit and thus this ammonia is disposed-off into the environment. This ammonia is found to be an environmental hazard as well as being a wasted resource within the industry.

The effluent treatment plant is equipped to deal with impurities such as hydrogen fluoride (HF) and phosphorus pentaoxide(P_2O_5). Water carrying these impurities is treat with calcium oxide to for insoluble salts that are dried and removed as sludge after passing through various thickeners and filters.

Although these plants are proficient at the removal of these impurities, ammonia passes through untouched and poses as an environmental hazard due to the plant's inability to deal with it. Ammonia, when in quantities greater than the EPA specified 20 ppm, can act as a severe hazard to health. It causes irritation on contact due to its basic nature and may damage the lungs on inhalation.

1.2 Statement of the Problem

The objective of our project is the removal and recovery of ammonia that is present in the effluent coming from the NP plant at Pak-Arab Fertilizers. Our project has three major amendments that are being made into the Effluent Treatment. A **heat exchanger** to allow for our stream to reach the operating temperature for the **stripper** which is then used to separate the ammonia from the water which is finally compressed towards the **absorber** where it is chemically absorbed into H_2SO_4 to form $(NH_4)_2SO_4$ as our final product.

1.3 Purpose of the Study

The losses of ammonia experienced through the effluent treatment plant may not be significant to the operation, but their significance lies in the danger they exhibit to the environment. 800 ppm of ammonia was dumped per day through the ETP whilst the Environmental Protection Agency (EPA) has an allowance of 20 ppm.

Ammonia affects humans as well as aquatic life. Most people are exposed to ammonia from breathing its gas or vapors. It is lighter than air and will rise, so that generally it does not settle in low-lying areas. However, in the presence of moisture, ammonia can form vapors that are heavier than air. These vapors can spread along the ground or other low-lying areas. When ammonia enters the body as a result of breathing, swallowing or skin contact, it reacts with water to produce ammonium hydroxide. This chemical is very corrosive and damages cells in the body on contact. Ammonia is corrosive. The severity of health effects depends on the route of exposure, the dose and the duration of exposure. Exposure to high concentrations of ammonia in air causes immediate burning of the eyes, nose, throat and respiratory tract and can result in blindness, lung damage or death. Inhalation of lower concentrations can cause coughing, and nose and throat irritation. Swallowing ammonia can cause burns to the mouth, throat and stomach. Skin or eye contact with concentrated ammonia can also cause irritation and burns.

When ammonia is present in water at high enough levels, it is difficult for aquatic organisms to sufficiently excrete the toxicant, leading to toxic buildup in internal tissues and blood, and potentially death. Environmental factors, such as pH and temperature, can affect ammonia toxicity to aquatic animals.

Therefore, the release of such quantity of ammonia into the environment is unacceptable and must be treated.

1.4 Definitions of the Considered Terms

1.4.1 Biological Nitrification and Denitrification:

The biological conversion of ammonium to nitrate nitrogen is called Nitrification. Nitrification is a two-step process. Bacteria known as Nitrosomonas, converts ammonia and ammonium to nitrite. Next, bacteria called Nitrobacter finishes the conversion of nitrite to nitrate. The reactions are generally coupled and proceed rapidly to the nitrate form; therefore, nitrite levels at any given time are usually low.

The biological reduction of nitrate (NO_3) to nitrogen gas (N_2) by facultative heterotrophic bacteria is called Denitrification. “Heterotrophic” bacteria need a carbon source as food to live. “Facultative” bacteria can get their oxygen by taking dissolved oxygen out of the water or by taking it off of nitrate molecules

1.4.2 Ion Exchange Resins:

It is possible to use natural resins, zeolites (Chabazite, Modernite, Clinoptilolite), or synthetic resins, the latter being preferred for reasons of durability and ease of implementation. These, of cationic, set primarily calcium and magnesium ions, and, to a lesser extent, ammonium ions. Their use is therefore justified in the context of a softening treatment. Then they allow the removal of ammonia and possibly even some cationic micropollutants. This process is based on the ammonium ion exchange in the water with sodium ions contained in the resin.

1.4.3 Breakpoint Chlorination:

Breakpoint chlorination is the point where the demand for chlorine has been fully satisfied in terms of chlorine addition to the water.

When chlorine is added to water, a reaction is produced in the compounds present in the water. These compounds utilize the chlorine, resulting in zero chlorine residual. Once chlorine has been added to water, it is consumed by a type of chemical reaction that has a net effect of increasing the chlorine concentration. For a typical addition of chlorine, the reaction rate instantly increases and reduces the chlorine concentration. This is because chlorinated compounds acquire more chlorine. The pace at which the chlorine atoms are added is comparatively slow, but the rates can be faster because chlorinating increases the reactivity. Once almost all of the chlorine reactions are accomplished then adding more chlorine leads to permanent residual chlorine.

1.4.4 Air Stripping and Absorption:

Air stripping is a technology in which volatile compounds are scrubbed off water to air. Typically, air stripping takes place in a packed tower (known as an air stripper) or an aeration tank. The "air stripper" includes a spray nozzle at the top of the tower. It sprays groundwater that has been pumped to the surface over the packing in the column. As the water descends, air is forced up through the column, stripping off the volatile compounds.

The volatile compound in this case is the ammonia that is in the water. Packing or baffles within the tower increase the surface area of the contaminated water that is exposed to air, thus maximizing the amount of volatilization.

Absorption is a process in which atoms, molecules or ions enter some bulk phase liquid or solid material. This is a different process from adsorption, since molecules undergoing absorption are taken up by the volume, not by the surface (as in the case for adsorption). A more general term is sorption, which covers absorption, adsorption, and ion exchange. Absorption is a condition in which something takes in another substance. Ammonia is difficult to remove though physical absorption hence chemisorption is used in which H_2SO_4 is used to react with ammonia removed from the stripper. This reaction produces $(NH_4)_2SO_4$ which is a fertilizer.

1.5 Contribution of the Study

Our study has ensured that the fertilizer industry prospers from their waste streams. This study has shown that waste ammonia can be used to produce 4000 mol/h of $(NH_4)_2SO_4$

which is a substantial amount and can be used effectively. The process has been designed and explained intricately so as to ensure that the industry may easily prosper from this study.

1.6 Organization of the Remainder of the Study

The authors have done their best in organizing every single detail in the process designed. The contents of this report should be sufficient for anyone looking to implement this process on a practical level. The specifications of the design are precise and some extra margins for contingencies have been considered throughout the designing phase of the project.

2 Literature Review

2.1 Process Description

The effluent treatment plant was installed to treat impurities such as P_2O_5 and HF but as the processes in the industry evolved, the impurities increased and ammonia was also incorporated to the stream owing to leakages and losses. The plant, however, was not equipped to remove ammonia. The Phosphate and Fluoride impurities are formed as a result of dissolution of Rock Phosphate in Nitric acid during the production of NitroPhosphate Fertilizers.

Initially, a flash mixing tank is used to mix the effluent from the NP Plant with CaO which is used to treat P_2O_5 and HF. The mixture is then sent into a reaction clarifier. This clarifier allows the CaO to react and form insoluble salts i.e. $CaHPO_4$ and CaF_2 which are then removed and sent towards a sludge thickener. The remaining mixture is pumped forward to the 2nd Reaction Clarifier. Before this stream enters the clarifier, CO_2 is injected into the mixture. This CO_2 reacts with the excess CaO to form $CaCO_3$. The insoluble solids formed from reaction are also sent into the sludge thickener.

The sludge thickener allows the solids to settle whilst removing the water. Sludge thickener can be of Rotary drum type. The sludge is further pushed into a filter press which produces a sludge cake that may be further used. The filter press is of Plate and Frame Type.

The initial stream, now free of salts, then enters a heat exchanger which heats the stream to $60^{o}C$ so that it may be prepared for the stripper. The stripper and absorption method was solely chosen on the basis of the product produced which fulfills the recovery of ammonia in this system. The rest of the above mentioned techniques were proficient at ammonia removal but lacked quality recovery products that may be used.

2.2 Process Flow Diagram

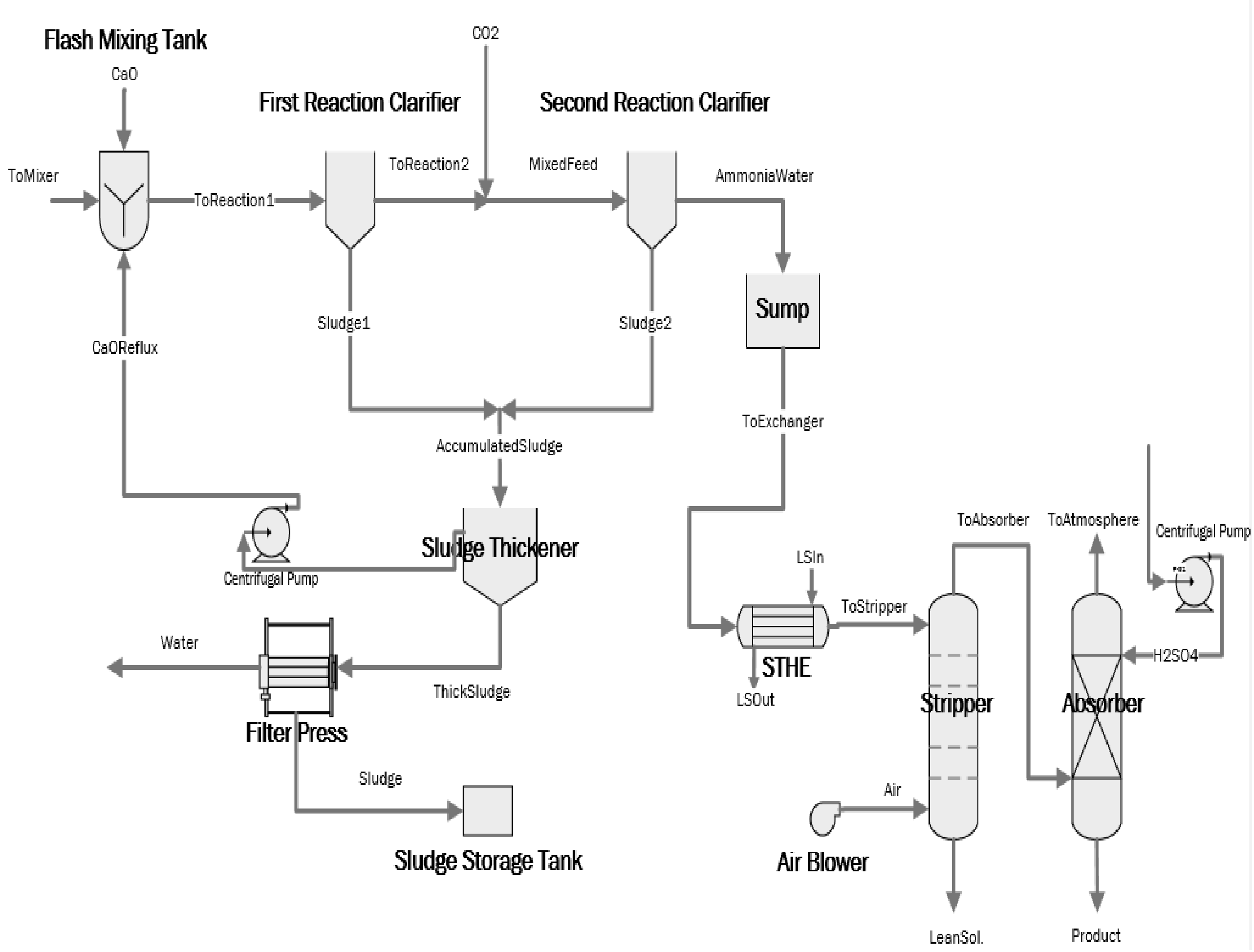


Figure 1: Process Flow Diagram

2.3 Ammonia Removal & Recovery Process

The basic aim of the project is the removal of ammonia from the water that comes after the NP plant and then to recover that ammonia by reacting it with a suitable acid to form a fertilizer which would be useful for the industry. There are several methods which could be used for the removal and recovery of ammonia but the most reliable and adoptable method on industrial scale is the Air or Steam Stripping process. Other methods like biological nitrification and denitrification break-point chlorination are lab scale methods which could not be used on industrial scale efficiently, hence the method of steam or air stripping must be considered.

2.3.1 Choice of Air Stripping

The reason of choosing air stripping instead of steam stripping is that ammonia is already present in water if we choose steam stripping that ammonia will be recovered from the water but it will again be present in the similar medium after that steam would be condensed. Hence the problem of removing ammonia from the water stream shall remain the same. This is why steam stripping is not considered for this process and Air stripping is chosen.

2.3.2 Air Stripping

Air stripping is a process in which air is made to contact with the water which contains a solute which is required to be stripped off. Air is passed from the bottom of the column and the water coming from the top is passed through the distribution plate which converts the water in droplets so that when the rising air contacts with the falling droplets the surface tension would be minimum and the rate of mass transfer of the solute present in water to the clean air would be maximum. This is the general principle on which the air stripping column works. Air stripping is a controlled process for selected ammonia removal and it is unaffected by the toxic compounds. The factors which increase the efficiency of stripping process are the liquid loading rate height and diameter of the column and the stripping factor.

2.3.3 Why Absorption is Required

After the ammonia is removed from the water it is transferred in the gaseous medium where if it is discharged in atmosphere would disobey the environmental standards, for example in our process 800 ppm of ammonia is present initially and after stripping the amount that would be removed should be greater than 750 ppm so it means that 750 ppm of ammonia would at least be present in air after the stripping process and if that air is discharged in atmosphere it would affect the environment in a negative way because the environmental standards allow us to discharge less than 40 ppm of ammonia in air.

2.4 Absorption

The discharge of the stripping column which is in the gaseous phase is used as the inlet of the absorber column. The gaseous medium consists of ammonia which needs to be recovered and it is added in the column from the bottom of the vessel. To recover the ammonia an absorbent is required, selection of which depends upon the affinity of the absorbent for the solute present in the gaseous phase. Several absorbents could be used like HNO_3 H_2SO_4 MDEA etc. and but we have chosen H_2SO_4 as our absorbent. Our absorption process involves a chemical reaction in which our absorbent reacts with ammonia to form a lean solution of ammonium sulphate. Like other processes in absorbers this one is of a different kind because it is not a physical absorption process, hence the usual design equations can't be used for this and a chemisorption process needs to be followed. In general H_2SO_4 reacts with NH_3 to form ammonium sulphate and the discharge air from the absorber column contains less than 40 ppm of ammonia.

2.4.1 Ammonium Sulphate:

As said before ammonium sulphate will be obtained as a by-product in the ammonia removal and recovery process. It is mainly used as fertilizer (N=21%) for field and leaf fertilizing. Other uses are

- Water Treatment
- Fermentation
- Fire Proofing
- Manufacture of viscose rayon
- Tanning
- Food additive

2.5 pH Adjustment in Process

Ammonia nitrogen exists in aqueous solution as either ammonium ion or ammonia, according to the following equilibrium reaction:

$$NH_4^+ \Leftrightarrow NH_{3(aq)} + H^+$$

According to this reaction, the presence of ammonia as free or dissolved form depends on the pH. If pH is high the equilibrium moves towards the forward direction and most of the ammonia exists in free form (NH_3).

In the flash mixing tank CaO is added, this leads to the increment in the amount of hydroxide ions present in our process stream, which then increases the pH and the ammonia which was initially present in dissolved ammonium ion form, gets converted to Ammonia.

We can see from this graph that at higher temperature and at higher pH most of ammonia is in free-state so for our stripping process we have chosen 60˚C and the pH is 11.

At the chosen temperature and pH value ammonia stripping is most favorable. We are capable of reducing ammonia amount to 20 ppm in the effluent liquid stream. Stripping is favorable at a higher temperature and a lower pressure but the other factor which controls stripping in our case is the value of pH. At a lower value of pH the ammonia is present in dissolved form which is not easily stripped out and hence resulting in lowering the stripping factor. When the stripping factor is lowered we have to increase it by increasing the liquid loading rate which adds to the cost. Hence an optimum pH is required in the stripping column which should keep the stripping factor greater than 1 and that pH has a value between 9-11. In our case the value of pH is 11 which is maintained by adding CaO. Hence for the removal of ammonia from 800 ppm to 20 ppm in the stripping column we are operating on a ph of 11 and at a temperature of 60˚C. Further increasing the temperature doesn’t affect the value of free ammonia present in the system hence as a result of which we don’t further increase the temperature of our column.

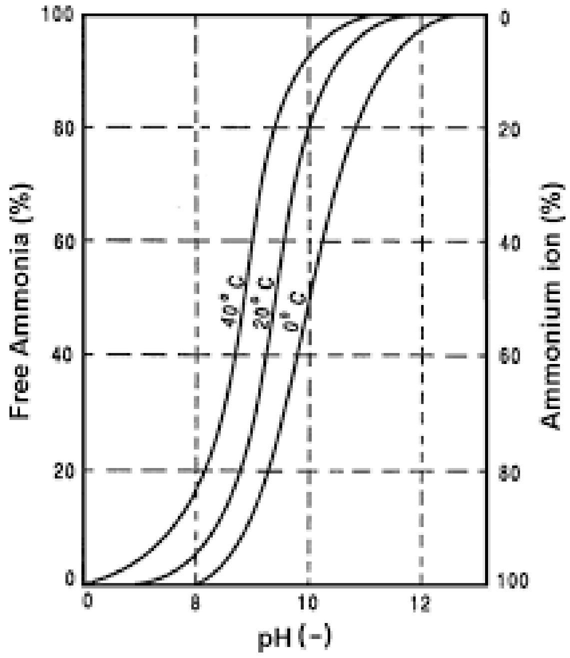


Figure 2: pH effect on Ammonia Conc.

2.6 Packing in Absorber and Stripper Column

Packings are basically used to increase the contact time and surface area for both fluids and have low pressure drop as compared to the trays and plates.

There are basically two types of arrangement available for packing:

1. Structured packing
2. Random packing

2.6.1 Structured Packing

These packings are usually thin corrugated metal gauzes which are placed in a manner that forces the fluids to take complex paths while passing through the packed bed, in this way it creates a large surface area for contact between different phases. These are used for cases in which exceptionally low pressure drop and high flow rates are required and when diameter of packing is greater than 3 in.

It can be made of different materials like

1. Metal
2. Plastic
3. Wire gauzes

2.6.1.1 Advantages

- Pressure drop is lower as compared to random packing
- It has higher capacity than random packing
- Efficiency is higher
- Liquid hold up is low

2.6.1.2 Disadvantages

- High cost
- Higher sensitivity to mal-distribution of liquids

2.6.2 Random Packing

This term is used for specially designed materials which are dumped into the column randomly to increase the surface area and contact time between the different phases. These are made up of different materials like

- Plastic
- Ceramic
- Metal

2.6.2.1 Advantages

- High pressure applications
- Low cost
- High liquid loads
- Ease of storage and replacement

2.6.2.2 Disadvantages

- Pressure drop is slightly higher than stacked packing
- Efficiency is slightly lower than stacked packing

2.7 Selection of Packing

Packing is generally selected on the basis of following factors

- Cost
- Temperature and pressure requirement of the system
- Corrosion resistance required
- Allowable pressure drop
- Capacity of loading and Efficiency required

2.7.1 Best Packing Option for our System

Figure 3: Pall Rings

Structured packing is used to enhance the capacity and efficiency of column. As we are designing a new column without any serious pressure drop constraint we are selecting **random packing** due to its economic efficiency. In random packing we selected **Pall rings,** which are basically an advanced form of Raschig rings. It has cylindrical dimension similar to raschig rings but has two lines of punched holes along with fingers or webs which are turned into inner side of cylinder. Pall rings provide high capacity and low pressure drop. This structure increases the performance in terms of throughput and pressure drop. Material of random packing was selected on the basis of our system temperature, pressure requirement and cost of material which resulted in metal type and specifically **stainless steel** to avoid any corrosion and because of our system.

2.7.1.1 Advantages of Stainless Steel Pall Rings

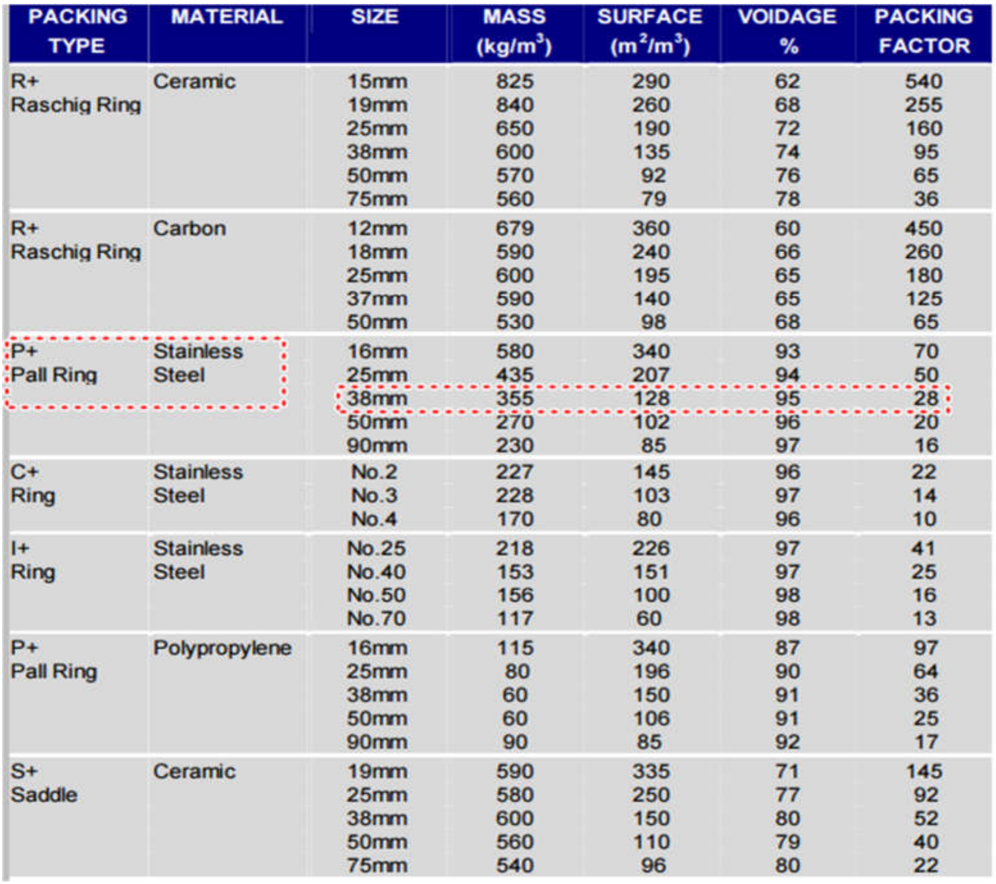
- Better liquid/gas distribution
- High capacity and low pressure drop

- High mechanical strength
- High temperature applications
- Probability of breakage is very low
- Wetting is easy
- High resistance to fouling

2.7.1.2 Specifications of Pall Rings

Given is the table, which includes the specifications of different types of packings along with the specifications of pall rings.

As the nominal diameter increases the pressure drop decreases but efficiency and surface area decreases so we selected pall rings with **50 mm diameter** for our design because larger nominal diameter will reduce efficiency and lower will increases pressure drop so we selected the diameter of 50 mm in between the diameters available.



PACKING TYPE	MATERIAL	SIZE	MASS (kg/m^3)	SURFACE (m^2/m^3)	VOIDAGE %	PACKING FACTOR
R+ Raschig Ring	Ceramic	15mm	825	290	62	540
		19mm	840	260	68	255
		25mm	650	190	72	160
		38mm	600	135	74	95
		50mm	570	92	76	65
		75mm	560	79	78	36
R+ Raschig Ring	Carbon	12mm	679	360	60	450
		18mm	590	240	66	260
		25mm	600	195	65	180
		37mm	590	140	65	125
		50mm	530	98	68	65
P+ Pall Ring	Stainless Steel	16mm	580	340	93	70
		25mm	435	207	94	50
		38mm	355	128	95	28
		50mm	270	102	96	20
		90mm	230	85	97	16
C+ Ring	Stainless Steel	No.2	227	145	96	22
		No.3	228	103	97	14
		No.4	170	80	96	10
I+ Ring	Stainless Steel	No.25	218	226	97	41
		No.40	153	151	97	25
		No.50	156	100	98	16
		No.70	117	60	98	13
P+ Pall Ring	Polypropylene	16mm	115	340	87	97
		25mm	80	196	90	64
		38mm	60	150	91	36
		50mm	60	106	91	25
		90mm	90	85	92	17
S+ Saddle	Ceramic	19mm	590	335	71	145
		25mm	580	250	77	92
		38mm	600	150	80	52
		50mm	560	110	79	40
		75mm	540	96	80	22

Figure 4: Packing Table

3 Material Balance

3.1 Flash Mixing Tank

Stream 1: To Mixer:

Name	Mass flow rate (kg/day)
P_2O_5	2395
HF	554
H_2O	4080×10^3
Dissolved NH_3	3252

Table 1: Stream 1

Total mass flow rate of To Mixer: 4086.2×10^3 kg/d

Stream 2: CaO Reflux:

Name	Mass flow rate (kg/day)
CaO	446.5
Water	720×10^3

Table 2: Stream 2

Total mass flow rate of CaO Reflux: 720.447×10^3 kg/d

Stream 3: Stream CaO:

Name	Mass flow rate (kg/day)
CaO	8115
Water	216×10^3

Table 3: Stream 3

Total mass flow rate of CaO stream: 224.12×10^3 kg/d

Applying Mass Balance equation:

Total mass flow rate in – Total mass flow rate out + Generation = Accumulation

Assuming the process is occurring at steady state, the accumulation term becomes zero.

Hence, Input = Output

In the first unit, that is the flash mixing tank, there are three inlet streams and the composition of the outlet stream is found by adding the moles of each component in the three inlet streams.

To mixer + CaO Reflux + CaO = ToReaction1

$4086.2\text{x}10^3 + 720446.5 + 224.12\text{x}10^3 = 5030.8\text{x}10^3$ kg/day

Applying P_2O_5 balance:

P_2O_5 in stream 1 + P_2O_5 in stream 2 + P_2O_5 in stream 3 = P_2O_5 in stream 4

2395 + 0 + 0 = 2395 kg/day

Applying HF balance:

HF in stream 1 + HF in stream 2 + HF in stream 3 = HF in stream 4

554 + 0 + 0 = 554 kg/day

Applying CaO balance:

CaO in stream 1 + CaO in stream 2 + CaO in stream 3 = CaO in stream 4

0 + 446.5 + 8115 = 8561.5 kg/day

Applying H_2O balance:

H_2O in stream 1 + H_2O in stream 2 + H_2O in stream 3 = H_2O in stream 4

$4080\text{x}10^3 + 720\text{x}10^3 + 216\text{x}10^3 = 5{,}016\text{x}10^3$ kg/day

Dissolved NH_3 balance:

Dissolved NH_3 in stream 1 + Dissolved NH_3 in stream 2 + Dissolved NH_3 in stream 3 = Dissolved NH_3 in stream 4

3252 + 0 + 0 = 3252 kg/day

Hence, composition of **ToReaction1 stream** (Stream 4):

Name	Mass flow rate (kg/d)
P_2O_5	2395
HF	554
H_2O	$5,016x10^3$
Dissolved NH_3	3252
CaO	8651.5

Table 4: Stream 4

Total mass flow rate of ToReaction1: $5030.853x10^3$ kg/d

3.2 First Reaction Clarifier

ToReaction1 (stream 4) = ToReaction2 (stream 5) + Sludge 1 (stream 9)

In first reaction clarifier, three reactions are taking place. First is the reaction of HF with CaO to produce CaF_2. The conversion of reaction given is 92.8%. So we calculate the amount of CaO used using stoichiometric mole balance (2:1) and similarly, the amount of CaF_2 produced is calculated as well. Second reaction is of P_2O_5 with water to produce H_3PO_4. Conversion of reaction is 98% so amount of H_3PO_4 produced is found by the stoichiometric mole balance (1:2). Third reaction is of H_3PO_4 with CaO to produce $CaHPO_4$, CaO used and $CaHPO_4$ produced are calculated similarly. Unreacted amount of CaO is calculated by subtracting the amount of CaO used in two reactions from its amount in the inlet stream.

Reactions occurring in First Reaction Clarifier:

Compound	Molar Mass kg/kmol
CaO	56
HF	20
CaF_2	78
H_2O	18
P_2O_5	142
H_3PO_4	98
$CaHPO_4$	136

Table 5: First Clarifier

1. **$CaO + 2HF \rightarrow CaF_2 + H_2O$**

   Conversion given = 92.8%

Moles of HF= 554 kg/ 20 kg/kmol = 27.7 mols

CaF_2 produced from 27.7 moles of HF = 27.7/2 = 13.85 mols

Mass of CaF_2 = 13.85 x 0.928 x 78

= 1002.5 kg/day

CaO used in reacting with 27.7 moles of HF = (27.7 x 0.928)/2 = 12.853

Mass of CaO used = 12.853 x 56 = 719.76 kg/day

H_2O produced from 27.7 moles of HF = (27.7 x 0.928 x 18)/2 = 231.35 kg/day

2. **$P_2O_5 + 3 H_2O \rightarrow 2 H_3PO_4$**

   Conversion = 99.2 %

Moles of P_2O_5 = 2395 kg/ 142 kg/kmol = 16.866 mols

H_3PO_4 produced from 16.866 moles of P_2O_5 = (2x16.866x0.992x98) = 3279.33 kg/day

H_2O used in reacting with 16.866 mols of P_2O_5 = 3x16.866x0.992x18 = 903.5 kg/day

3. $CaO + H_3PO_4 \rightarrow CaHPO_4 + H_2O$

C $aHPO_4$ produced from 33.46 moles of H_3PO_4 = 33.46x136 = 4550.9 kg/day

CaO used in reacting with 33.46 moles of H_3PO_4 = 33.46x56 = 1873.9 kg/day

H_2O produced from 33.46 moles of H_3PO_4 = 33.46x18= 602.33 kg/day

Unreacted CaO = CaO in – CaO reacted
= (8561.3) – (719.76+1873.9)
= 5967.64 kg/day

Water left = water in – water reacted + water produced
= 5,016x10^3 – 903.5 + 231.35 + 602.33
= 5,015.9x10^3 kg/day ≈ 5,016x10^3 kg/day

P_2O_5 left = P_2O_5 in - P_2O_5 reacted ∵ conversion = (mass in – mass out)/ mass in
= 2395 – (2395x0.992)
= 20 kg/day

HF left = HF in - HF reacted ∵ conversion = (mass in – mass out)/ mass in
= 554 – (554x0.928)
= 40 kg/d

There are two outlet streams from first reaction clarifier namely, Sludge1 and ToReaction2. ToReaction2 stream contains water from which insoluble P_2O_5 and CaF_2 are removed. Sludge1 stream contains these insoluble impurities which have settles down under the action of gravity.

It is required to recycle 7.5% CaO from first reaction clarifier to flash mixing tank so 7.5% of the unreacted CaO leaves in the stream sludge1 and the remaining CaO goes to the second reaction clarifier.

CaO in Sludge1 = 446.5 kg/d

CaO in ToReaction2 = 5967.64 – 446.5

= 5521.14 kg/d

Water entrained with slurry is also 7.5% of the total inlet water, so the amount calculated is,

$= 0.75 \times 5016 \times 10^3$

$= 381.2 \times 10^3$ kg/d

Hence, water leaving in stream ToReaction2

$= 5016 \times 10^3 - 381.2 \times 10^3$

$= 4635 \times 10^3$ kg/d

Hence, composition of **ToReaction2 stream** (Stream 5):

Name	Mass flow rate (kg/day)
P_2O_5	20
HF	40
H_2O	$4,635 \times 10^3$
Dissolved NH_3	3252
CaO	5521.14

Table 6: Stream 5

Total mass flow rate of ToReaction2: 4643.8×10^3 kg/d

Composition of **Sludge1 stream** (Stream 9):

Name	Mass flow rate (kg/day)
$CaHPO_4$	4550.9
CaF_2	1002.5
H_2O	381.2×10^3

CaO	446.5

Table 7: Stream 9

Total mass flow rate of Sludge2: 387.2x10^3 kg/d

3.3 Second Reaction Clarifier

Now, the ToReaction2 stream enters the second reaction clarifier, where CO_2 is added to remove the surplus amount of CaO which remained unreacted. The unreacted CaO from first reaction clarifier is completely converted to $CaCO_3$ in second reaction clarifier using CO_2. First, we calculate the amount of CO_2 required for complete conversion of 98.59 kmol/d of CaO. Then the amount of $CaCO_3$ produced is found, as there is a 1:1 mol ratio of CaO and $CaCO_3$ so same amount of $CaCO_3$ is produced. This $CaCO_3$ alongwith 7.5% pf water leaves as sludge2.

Overall material balance of second reaction clarifier shows that the amount of CaO is completely converted to $CaCO_3$ while others remain the same.

$CaO + CO_2 \rightarrow CaCO_3$

Amount of CO_2 required = (5521.14 x 44)/ 56

= 4338 kg/d

Hence, composition of **CO_2 stream** (stream 6)

Name	Mass flow rate (kg/day)
CO_2	4338
H_2O	1,320x10^3

Table 8: Stream 6

Total mass flow rate of CO2 stream: 1324.338x10^3 kg/d

This stream adds to the ToReaction2 stream and forms a new stream, MixedFeed. The composition of **MixedFeed stream (**stream 7) then becomes,

Name	Mass flow rate (kg/day)
P_2O_5	20
HF	40
H_2O	$5954.5x10^3$
Dissolved NH_3	3252
CaO	5521.14
CO_2	4338

Table 9: Stream 7

Total mass flow rate of MixedFeed stream: $967.7x10^3$ kg/d

To find amount of $CaCO_3$ produced we use stoichiometric mole balance and find out that amount of $CaCO_3$ produced will be,

= (5521.14x100)/56

= 9859.2 kg/d

Now, the insoluble $CaCO_3$ formed will settle down and leave in the stream Sludge2 alongwith an entrainment of 7.5% of water.

Water in sludge2 = 0.075 x $5954.5x10^3$

= $446.5x10^3$ kg/d

Hence, composition of **Sludge2 stream** (stream 10)

Name	Mass flow rate (kg/day)
$CaCO_3$	9859.2
H_2O	$446.5x10^3$

Table 10: Stream 10

Total mass flow rate of Sludge2 stream: $456.4x10^3$ kg/d

Remaining water leaves in the supernatant stream, which is AmmoniaWater. It is calculated by subtracting the amount of water leaving in sludge2 stream from the total inlet amount of water.

Water in AmmoniaWater Stream = $5954.5 \times 10^3 - 446.5 \times 10^3$

$= 5508 \times 10^3$ kg/d

Composition of **AmmoniaWater Stream** (stream 8)

Name	Mass flow rate (kg/day)
P_2O_5	20
HF	40
H_2O	5508×10^3
Dissolved NH_3	3252

Table 11: Stream 8

Total mass flow rate of AmmoniaWater Stream: 5511.3×10^3 kg/d

3.4 Sludge Thickener

Streams sludge1 and sludge2 combine and enter in the sludge thickener as AccumulatedSludge stream. Here, dehydration of sludge takes place, water is separated from sludge. Thick, viscous sludge is obtained from the less viscous sludge.

Composition of **AccumulatedSludge Stream** (stream 11)

Name	Mass flow rate (kg/day)
$CaHPO_4$	4550.9
CaF_2	1002.5
H_2O	827.7×10^3
CaO	446.5
$CaCO_3$	9859.2

Table 12: Stream 11

Total mass flow rate of AccumlatedSludge Stream: 3388.11×10^3 kg/d

Sludge thickener provides the necessary residence time for the sludge to settle down and water to be separated. The water separated from sludge is recycled back to the flash mixing tank.

The composition of the reflux stream, CaOReflux, is known:

CaO	446.5
Water	720×10^3

Table 13: Reflux Stream

Total mass flow rate of CaO Reflux: 720446.5 kg/d

Hence, the composition of ThickSludge stream can be found out by subtracting the values of CaOReflux from AccumlatedSludge Stream.

Amount of water in ThickSludge stream = $827.7 \times 10^3 - 720 \times 10^3$
= 107.7×10^3 kg/d

The composition of the **ThickSludge stream** (stream 13):

Name	Mass flow rate (kg/day)
$CaHPO_4$	4550.9
CaF_2	1002.5
H_2O	107.7×10^3
$CaCO_3$	9859.2

Table 14: Stream 13

Total mass flow rate of ThickSludge Stream: 123.113×10^3 kg/d

3.5 Filter Press

Here, complete separation of water from sludge takes place. The ThickSludge Stream enters the filter press and by mechanical action, the water present in it is completely separated and solid cake of the sludge is obtained.

Water leaving in **Water** Stream (stream 14) = 107.7x10^3 kg/d

Composition of **Sludge** stream (stream 15):

Name	Mass flow rate (kg/day)
$CaHPO_4$	4550.9
CaF_2	1002.5
$CaCO_3$	9859.2

Table 15: Stream 15

Total mass flow rate of Sludge Stream: 15.413x10^3 kg

3.6 Heat Exchanger

Material Balance for heat exchanger is simple. Amount of material in remains equal to amount of materials out as no reaction takes place and there is no accumulation or generation.

Composition of **Shell side inlet Stream** (stream 8)

Name	Mass flow rate (kg/day)
P_2O_5	20
HF	40
H_2O	5508x10^3
Dissolved NH_3	3252

Table 16: Shell Inlet

Total mass flow rate of AmmoniaWater Stream: 5511.3×10^3 kg/d

Material in = Material out

So, composition of **Shell side outlet stream** (ToStripper):

Name	Mass flow rate (kg/day)
P_2O_5	20
HF	40
H_2O	5508×10^3
Dissolved NH_3	3252

Table 17: Shell Outlet

Total mass flow rate of ToStripper Stream: 5511.3×10^3 kg/d

Composition of **Tube Side inlet Fluid** (LS_{in}):

Name	Mass flow rate (kg/day)
L.S	5.4336×10^6

Table 18: Tube Inlet

The tube side fluid is the heating medium. It releases heat which is absorbed by the AmmoniaWater stream. Its composition remains the same whereas its heat energy changes.

So, L.S in = L.S out

Composition of **Tube Side outlet Fluid** (LS_{out}):

Name	Mass flow rate (kg/day)
L.S	5433.6×10^3

Table 19: Tube Outlet

3.7 Stripper

Following notations are used in the material balance calculation across the stripper:

L_{in} = Amount of liquid at the inlet of stripper
X_{in} = Mole ratio of solute to solute free liquid at inlet
L_{out} = Amount of liquid at outlet
X_{out} = Mole ratio of solute to solute free liquid at outlet
G_{in} = Amount of air at the inlet of stripper
Y_{in} = Mole ratio of solute to solute free air at inlet
G_{out}= Amount of air at the outlet
Y_{out} = Mole ratio of solute to solute free air at outlet

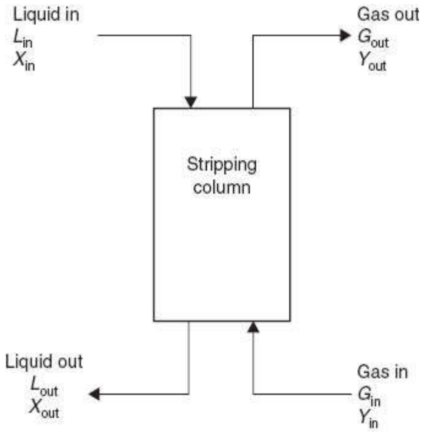


Figure 5: Stripper

Conditions at Inlet:

Component	Molar flow(mol/h)	Mol solute/ mol of solute free substance
Water	1.27×10^7	
Ammonia	7972	
Xin		0.0006
Air	?	
Yin		0

Table 20: Stripper Inlet

In order to find the molar flow rate of air we must apply the mass balance across the stripper. The equation for mass balance across the stripper is given below

$$L_{in}X_{in} + G_{in}Y_{in} = L_{out}X_{out} + G_{out}Y_{out}$$

The left-hand side values are known except Gin

For the left hand-side we know the value of X_{out} must be less than **20ppm** as per our environmental standards.

For our own convenience we assume that $\mathbf{L_{in} = L_{out}}$ **and** $\mathbf{G_{in} = G_{out}}$**.**

Now to obtain $\mathbf{Y_{out}}$ we apply the henry law to calculate the partial pressure which our ammonia would exert in the gaseous phase upon mass transfer

For $\mathbf{Y_{out}}$

$\mathbf{Pi = H \times X_{in}}$ (Henry law)

The value of H (Henry constant) at 60(C) = 3.23 bar
$X_{in} = 0.0006$
Hence $P_i = 0.0019$ bar
$Y_{out} = P_i/P$
In our case, P = 1 bar
So $\mathbf{Y_{out} = 0.0019}$

By using the mass balance-equation the amount of flow rate of air required for ammonia stripping = $\mathbf{3.8 \times 10^6}$

This is the value obtained by applying henry law, hence this is at equilibrium. We multiply this value by a factor of 1.5 to get the actual value of G.

$\mathbf{G_{actual} = 6x10^6}$ **mol/hr**

3.8 Absorber

Since chemisorption is involved in our absorber, the mass balance is applied with the help of chemical equation.

Composition of Absorber feed (ToAbsorber Stream)

Name	Mass flow rate (mol/hr)

Air	$6x10^6$
Ammonia	7371
Sulphuric Acid	?
Water	?

Table 21: Absorber Inlet

The amount of H_2SO_4 required is obtained via a chemical equation.

$$\mathbf{H_2SO_4 + 2NH_3 \rightarrow (NH_4)_2SO_4}$$

According to the above chemical equation for absorption to take place adequately pH should be less (acidic) all the time hence H_2SO_4 should be present in the system throughout. At a pH of 5 more than 99% of ammonia is recovered through absorption. Hence 0.01% is discharged in atmosphere along with the air.

Composition of Absorber outlet:

Name	Mass flow rate (mol/hr)
Air	$6x10^6$
Ammonia	0.01x7371=73.71
Ammonium Sulphate	?
Water	?

Table 22: Absorber Outlet

Amount of ammonium sulphate formed will also be calculated by using the above equation.

For H_2SO_4

Mole balance is applied between NH_3 and H_2SO_4

For 2 moles of Ammonia, moles of **H_2SO_4** required = 1
For 1 mole of Ammonia, moles of **H_2SO_4** required = ½

For 7371 moles of Ammonia, moles of $\mathbf{H_2SO_4}$ required = ½ x 7371= **3686** (mol/hr)
Note: Since the system is to remain acidic throughout we multiply these moles of H_2SO_4 by 1.2 so,

Actual H_2SO_4 = 1.2 x 3686 = 4424 (mol/hr)

For $(NH_4)_2SO_4$

Mole balance is applied between NH_3 and $(NH_4)_2SO_4$

2 moles of NH3 produce moles of $(NH_4)_2SO_4$ = 1
1 mole of NH3, produce mole of $(NH_4)_2SO_4$ = 1/2
7298 moles of NH produce moles of $\mathbf{(NH_4)_2SO_4}$ = 0.5 x 7298= **3649 (mol/hr)**

In grams per hour: 3649x132 = $4.8\text{x}10^5$ (g/hr)

For H_2O

T(˚C)	g$(NH_4)_2SO_4$/100g of H_2O	% by Weight
30	78	43
40	81	44
60	88	46
80	95.3	48

Table 23: Water vs Temperature

According to the above table, at 60 ˚C 88grams of $(NH_4)_2SO_4$ dissolves per 100 grams of water. Since our operating temperature is around 55 ˚C, we will use 84 g instead of 88 g.

84 g $(NH_4)_2SO_4$ requires H_2O = 100
1 g $(NH_4)_2SO_4$ requires H_2O = 100/84
$4.8\text{x}10^5$ g $(NH_4)_2SO_4$ require H_2O = (100/84)x$4.8\text{x}10^6$
= $5.7\text{x}10^5$ (g/hr) = $\mathbf{3.1x10^4}$ **(mol/hr)**

Hence, flow rate of water = $\mathbf{3.1x10^4}$ **(mol/hr)**

4 Energy Balance

4.1 Energy Balance Formulae

Equation used for applying energy balance across the system is as under:

$$\mathbf{(Enthalpy + K.E + P.E + Q)_{in} = (Enthalpy + K.E + P.E + W)_{out}}$$

Firstly, the Q of all the streams was calculated using $mCp\Delta T$. Molar flow rates of all the streams are known. Water treatment occurs at room temperature so 30° Celsius temperature is used. Reference Temperature used is 273 K. for the calculation of heat capacity of each stream we used equation

$$\mathbf{Cp = a + bT + cT^2 + dT^3}$$

The values for a, b, c, d are taken from the book, Chemical Properties Handbook for each component. MATLAB code was devised for the Specific Heat integration. It helps to integrate the value of heat capacity between given temperature ranges. Once, the a, b, c, d co-efficient for each component are listed then simply by inserting the temperatures the value of C_p can be obtained.

The code used was:

```
a=input('temperature of the stream:')
b=input('reference temperature:')
Cpwater=33.933*(t-y)-8.4186*10^-3*(t^2-y^2)/2+2.9906*10^-5*(t^3-y^3)/3-1.7825*10^-
8*(t^4-y^4)/4+3.6934*10^-12*(t^5-y^5)/5
Cpammonia=33.573*(t-y)-1.2581*10^-2*(t^2-y^2)/2+8.906*10^-5*(t^3-y^3)/3-
7.1783*10^-8*(t^4-y^4)/4+1.8569*10^-11*(t^5-y^5)/5
```

Similarly, the code for other components (CaO, $CaHPO_4$, CaF_2, HF, P_2O_5, $CaCO_3$) was written and using MATLAB we found the heat capacity value for each component.

For, determining the heat capacity value of the stream, we used the equation:

$$\mathbf{Cp_{mixture} = x_aCp_a + x_bCp_b + x_cCp_c + \ldots..}$$

So, for any stream firstly we found the C_p values for all the components and then multiplied it with the molar fraction of each component. They are then added to find the C_p for the stream.

4.2 Flash Mixing Tank

In Mixing tank, we neglected the kinetic and potential energy changes between the inlet and outlet streams and the shaft work done by the mixing impeller. Hence energy balance equation reduced to $\mathbf{mH_{in} = mH_{out}}$ **+ heat of mixing.**

$$\mathbf{Q_1 + Q_2 + Q_3 = Q_4}$$

$$\mathbf{Q = m\, C_p\, \Delta T}$$

Composition of stream 1:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
P_2O_5	2395	16.866
HF	554	27.7
H_2O	4080×10^3	226.7×10^3
Dissolved NH_3	3252	191.29

Table 24: Composition of Stream 1

$m = 226.9 \times 10^3$ kmol/d

For stream 1,

$m = 226.9 \times 10^3$ kmol/d

$C_p = 1896.45$ J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_1} = 226.9 \times 10^3 \times 1896.45 \times (303 - 273)$

$= 1.29\text{x}10^{10}$ kJ/d

$= 5.379\text{x}10^{8}$ kJ/hr

Composition of stream 2:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
CaO	446.5	7.97
Water	$720\text{x}10^3$	$40\text{x}10^3$

Table 25:Composition of Stream 2

$m = 40.008\text{x}10^3$ kmol/d

For stream 2,

$m = 40.008\text{x}10^3$ kmol/d

$C_p = 1897.67$ J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

Q_2 = $40.008\text{x}10^3 \text{ x } 1897.67 \text{ x } (303 – 273)$

$= 2.2777 \text{ x}10^9$ kJ/d

$= 94.9\text{x}10^6$ kJ/hr

Composition of stream 3:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
CaO	8115	144.91
water	$216\text{x}10^3$	$12\text{x}10^3$

Table 26: Composition of Stream 3

$m = 12.1449\text{x}10^3$ kmol/d

For stream 3,

$m = 12.1449\text{x}10^3$ kmol/d

C_p = 1998.73 J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_3} = 12.1449\text{x}10^3 \text{ x } 1998.73 \text{ x } (303 - 273)$

$= 7.28 \text{ x}10^8 \text{ kJ/d}$

$= 30.3\text{x}10^6 \text{ kJ/hr}$

So,

$\mathbf{Q_1} + \mathbf{Q_2} + \mathbf{Q_3} = 5.379\text{x}10^8 + 94.9\text{x}10^6 + 30.3\text{x}10^6$

$= 6.631\text{x}10^8 \text{ kJ/hr}$

→ $\mathbf{Q_4} = 6.631\text{x}10^8 \text{ kJ/hr}$

Composition of stream 4:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
P_2O_5	2395	16.366
HF	554	27.7
H_2O	$5,016\text{x}10^3$	$27.87\text{x}10^3$
Dissolved NH_3	3252	191.29
CaO	8651.5	152.88

Table 27: Composition of Stream 4

m = $279.1\text{x}10^3$ kmol/d

To countercheck if the energy balance calculation is correct or not, $\mathbf{Q_4}$ was calculated using its own values of mass flow rate, specific heat and temperature difference and the value came out to be the same. Calculations shown as under,

For stream 4,

m = $279.1\text{x}10^3$ kmol/d

$C_p = 1898.9$ J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_4} = 279.1\text{x}10^3 \text{ x } 1898.9 \text{ x } (303 - 273)$

$= 1.589 \text{ x}10^{10}$ kJ/d

$= 6.624\text{x}10^8$ kJ/hr ≈ $6.63\text{x}10^8$ kJ/hr

This value is equal to that calculated by applying energy balance equation hence, energy balance is correct.

4.3 First Reaction Clarifier

In first reaction chamber three reactions are taking place. So, their heat of reactions is calculated using the equation (sum of heat of formation of products)- (sum of heat of formation of the reactants). These are exothermic reactions. Hence, their heat is also added in the heat of the sludge1 and ToReaction2 streams to calculate the outlet Q. And the energy difference of inlet and outlet is negligible. The temperature is not increased by the exothermic reactions as the concentrations are too small compared to the water concentration.

$\mathbf{Q_4 = Q_5 + Q_9 + Q_{reactions}}$

Here, following three exothermic reactions are taking place

$$CaO + 2HF \rightarrow CaF_2 + H_2O \ (\Delta H_{rxn} = -290.37 \text{ kJ/mol})$$

$$P_2O_5 + 3H_2O \rightarrow 2\ H_3PO_4 \ (\Delta H_{rxn} = -176.78 \text{ kJ/mol})$$

$$CaO + H_3PO_4 \rightarrow CaHPO_4 + H_2O \ (\Delta H_{rxn} = -207.7 \text{ kJ/mol})$$

The heat of reaction for each of the above reactions was calculated using equation:

$$\mathbf{\Delta H_{rxn} = \sum \Delta H_{formation}\ (products) - \sum \Delta H_{formation}\ (reactants)}$$

Heat of formation of each component was:

Component	**$\Delta H_{formation}$ (kJ/mol)**
CaO	-634.9
HF	-273.3
CaF_2	-1186.07
H_2O	-241.82
P_2O_5	-1519.82
H_3PO_4	-1277
$CaHPO_4$	-1833.8

Table 28:Heat of Formations

Composition of stream 5:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
P_2O_5	20	0.14
HF	40	2
H_2O	$4,635 \times 10^3$	257.5×10^3
Dissolved NH_3	3252	191.29
CaO	5521.14	98.59

Table 29: Composition of Stream 5

$m = 257.79 \times 10^3$ kmol/d

For stream 5,

$m = 257.79 \times 10^3$ kmol/d

$C_p = 1896.5$ J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_5}$ = 257.79x10^3 x 1896.5 x (303 – 273)

= 1.4667x10^{10} kJ/d

= 6.11x10^8 kJ/hr

Composition of stream 9:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
$CaHPO_4$	4550.9	33.46
CaF_2	1002.5	12.853
H_2O	381.2x10^3	21.178x10^3
CaO	446.5	7.97

Table 30: Composition of Stream 9

m = 21.232x10^3 kmol/d

For stream 9,

m = 21.232x10^3 kmol/d

C_p = 1895.76 J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_9}$ = 21.232x10^3 x 1895.76 x (303 – 273)

= 1.2675x10^9 kJ/d

= 50.3x10^6 kJ/hr

Equation used for calculating the Enthalpy of reaction is,

$$\mathbf{Q_{reaction} = \Delta H_{rxn} \text{ x moles of product produced}}$$

→ $Q_{reaction,\,1}$ = 290.37 kJ/mol x 12.853 kmol/d

= 3732.13x10^3 kJ/d

= 1.555x10^5 kJ/hr

→ $Q_{reaction, 2}$ = 176.78 kJ/mol x 33.46 kmol/d

= 5915×10^3 kJ/d

= 2.464×10^5 kJ/hr

→ $Q_{reaction, 3}$ = 207.7 kJ/mol x 33.46 kmol/d

= 6949.6×10^3 kJ/d

= 2.896×10^5 kJ/hr

Hence,

$\mathbf{Q_{reaction}}$ = $1.555 \times 10^5 + 2.464 \times 10^5 + 2.896 \times 10^5$

= 691500 kJ/hr

Now,

$\mathbf{Q_5 + Q_9 + Q_{reactions}}$

= 6.11×10^8 kJ/hr + 50.3×10^6 kJ/hr + 691500 kJ/hr

= 6.62×10^8 kJ/hr

≈ 6.624×10^8 kJ/hr = $\mathbf{Q_4}$

So, energy is balanced in the input and output streams.

Now, CO_2 stream (6) combines with the stream 5 and forms stream 7, the energy for that stream is calculated by:

$\mathbf{Q_5 + Q_6 = Q_7}$

Composition of stream 7:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
P_2O_5	20	0.14
HF	40	2
H_2O	5954.5×10^3	330.8×10^3

Dissolved NH_3	3252	191.29
CaO	5521.14	98.59
CO_2	4338	98.59

Table 31: Composition of Stream 7

$m = 331.196\text{x}10^3$ kmol/d

For stream 7,

$m = 331.196\text{x}10^3$ kmol/d

$C_p = 1899.25$ J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_7} = 331.196\text{x}10^3 \text{ x } 1899.25 \text{ x } (303 - 273)$

$= 1.887\text{x}10^{10}$ kJ/d

$= 7.86\text{x}10^8$ kJ/hr

4.4 Second Reaction Clarifier

$\mathbf{Q_7 = Q_8 + Q_{10} + Q_{reaction}}$

Here, following exothermic reaction is taking place

$$CaO + CO_2 \rightarrow CaCO_3 \ (\Delta H_{rxn} = -178.49 \text{ kJ/mol})$$

The heat of reaction was calculated using equation:

$$\Delta H_{rxn} = \sum \Delta H_{formation} \text{ (products)} - \sum \Delta H_{formation} \text{ (reactants)}$$

Heat of formation of each component:

Component	**$\Delta H_{formation}$ (kJ/mol)**
CaO	-634.9

CO_2	-393.51
$CaCO_3$	-1207.6

Table 32: Component Heat of Formation

→ ΔH_{rxn} = -178.49 kJ/mol

Composition of stream 8:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
P_2O_5	20	0.14
HF	40	2
H_2O	5508×10^3	306×10^3
Dissolved NH_3	3252	191.29

Table 33: : Composition of Stream 8

m = 306.193×10^3 kmol/d

For stream 8,

m = 306.193×10^3 kmol/d

C_p = 1897.59 J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_8}$ = $306.193 \times 10^3 \times 1897.59 \times (303 - 273)$

= 1.74×10^{10} kJ/d

= 7.263×10^8 kJ/hr

Composition of stream 10:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
$CaCO_3$	9859.2	98.592

H_2O	$446.5\text{x}10^3$	$24.8056\text{x}10^3$

Table 34: : Composition of Stream 10

m = $24.904\text{x}10^3$ kmol/d

For stream 10,

m = $24.904\text{x}10^3$ kmol/d

C_p = 1908.68 J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_{10}} = 24.904\text{x}10^3 \text{ x } 1908.68 \text{ x } (303 - 273)$

$= 1.426\text{x}10^9$ kJ/d

$= 5.94\text{x}10^7$ kJ/hr

Equation used for calculating the Enthalpy of reaction is,

$$Q_{reaction} = \Delta H_{rxn} \text{ x moles of product produced}$$

→ $Q_{reaction}$ = 178.49 kJ/mol x 98.592 kmol/d

$= 17597.7\text{x}10^3$ kJ/d

$= 7.3324\text{x}10^5$ kJ/hr

Now,

$\mathbf{Q_8 + Q_{10} + Q_{reaction}}$

$= 7.263\text{x}10^8$ kJ/hr + $5.94\text{x}10^7$ kJ/hr + $7.3324\text{x}10^5$ kJ/hr

$= 7.864\text{x}10^8$ kJ/hr

≈ $7.86\text{x}10^8$ kJ/hr = $\mathbf{Q_7}$

So, energy is balanced in the input and output streams of second reaction clarifier.

Streams 9 and 10 (sludge1 and sludge2) combine to form stream 11 which enters the sludge thickener.

→ $\mathbf{Q_{11} = Q_9 + Q_{10}}$

$= 50.3\text{x}10^6$ kJ/hr + $5.94\text{x}10^7$ kJ/hr

$= 1.097\text{x}10^8$ kJ/hr

4.5 Sludge Thickener

The water present in accumulated sludge stream is separated in sludge thickener.

$\mathbf{Q_{11} = Q_{12} + Q_{13}}$

Composition of stream 12:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
CaO	446.5	7.97
Water	$720\text{x}10^3$	$40\text{x}10^3$

Table 35: : Composition of Stream 12

m = $40.008\text{x}10^3$ kmol/d

For stream 12 which is equal to stream 2,

m = $40.008\text{x}10^3$ kmol/d

C_p = 1897.67 J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_{12}} = 40.008\text{x}10^3 \text{ x } 1897.67 \text{ x } (303 - 273)$

$= 2.2777 \text{ x}10^9$ kJ/d

$= 94.9\text{x}10^6$ kJ/hr

Composition of stream 13:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
$CaHPO_4$	4550.9	33.46
CaF_2	1002.5	12.853
H_2O	107.7×10^3	5983.33
$CaCO_3$	9859.2	98.592

Table 36: Composition of Stream 13

m = $6.128 \times 10^3 \times 10^3$ kmol/d

For stream 13,

m = 6.128×10^3 kmol/d

C_p = 1892 J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_{13}}$ = $6.128 \times 10^3 \times 1892 \times (303 - 273)$

= 3.478×10^8 kJ/d

= 1.449×10^7 kJ/hr

$\rightarrow \mathbf{Q_{12} + Q_{13}} = 94.9 \times 10^6$ kJ/hr + 1.449×10^7 kJ/hr

= 1.0966×10^8 kJ/hr

≈ 1.097×10^8 kJ/hr = $\mathbf{Q_{11}}$

Hence, energy of the inlet and outlet streams of Sludge Thickener are equal.

4.6 Filter Press

Here, complete removal of water from sludge takes place and solid cake is obtained.

$\mathbf{Q_{13} = Q_{14} + Q_{15}}$

Composition of stream 14:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
Water	107.7×10^3	5983.33

Table 37: Composition of Stream 14

m = 5.983×10^3 kmol/d

For stream 14,

m = 5.983×10^3 kmol/d

C_p = 2275 J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_{14}} =$ $5.983 \times 10^3 \times 2275 \times (303 - 273)$

$= 4.1 \times 10^8$ kJ/d

$= 1.449 \times 10^7$ kJ/hr

Composition of stream 15:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
$CaHPO_4$	4550.9	33.46
CaF_2	1002.5	12.853
$CaCO_3$	9859.2	98.592

Table 38: Composition of Stream 15

m = 144.91 kmol/d

For stream 15,

m = 144.91 kmol/d

C_p = 1692.44 J/mol K

inlet temperature = 30°C = 303K

reference temperature = 0°C = 273K

Hence,

$\mathbf{Q_{15}}$= 144.91x 1692.44 x (303 – 273)

= 7.357x10^6 kJ/d

= 3.0655x10^5 kJ/hr

$\mathbf{Q_{14}}$ + $\mathbf{Q_{15}}$ = 1.449x10^7 kJ/hr + 3.0655x10^5 kJ/hr

= 1.45x10^7 kJ/hr

≈ 1.449x10^7 kJ/hr = $\mathbf{Q_{13}}$

Hence, it is proved that energy in the inlet and outlet streams remains constant.

4.7 Heat Exchanger

Stream 8, AmmoniaWater, then goes to the heat exchanger. For air stripping we need temperature of 60 degree Celsius. So Q_{tube} is found using $mC_p\Delta T$ where ΔT is (60-30). This Q_{tube} is equal to Q_{shell}. For heating L.S is used whose temperature is 135 degrees Celsius. Its temperature drops to 85 degrees Celsius so the mass flow rate is the only unknown and it is found out by rearranging the equation.

Composition of Tube side stream (8):

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
P_2O_5	20	0.14
HF	40	2
H_2O	5508x10^3	306x10^3
Dissolved NH_3	3252	191.29

Table 39: Composition of Tube Side Stream

m = 306.193x10^3 kmol/d

For tube side fluid,

m = 306.193x10^3 kmol/d

C_p = 2635 J/mol K

inlet temperature = 30°C = 303K

outlet temperature = 60°C = 333K

Hence,

$\mathbf{Q_{tube}} = 306.193\text{x}10^3 \text{ x } 2635 \text{ x } (333 – 303)$

$= 2.42\text{x}10^{10}$ kJ/d

$= 1.01\text{x}10^9$ kJ/hr

Energy balance on heat exchanger tells that energy released by hot fluid is absorbed by the cold fluid, hence

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

L.S is the hot fluid which is present in shell. Its inlet temperature is 180°C and outlet temperature is 147°C. Heat capacity is 1606 J/mol.K. so,

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

$= m \textbf{ x } 1606 \text{ x } (453 – 420)$

$= Q_{tube}$

$\rightarrow \mathbf{m} = (1.01\text{x}10^9)/(1606 \text{ x } 30)$

= 12,577.83 kmol/hr

$= 5.4336\text{x}10^6$ kg/d

Hence, mass flow rate of L.S in shell is $5.4336\text{x}10^6$ kg/d.

Composition of Shell side stream:

Name	Mass flow rate (kg/d)	Molar flow rate (kmol/d)
L.S	$5.4336\text{x}10^6$	12577.83

Table 40: Composition of Shell Side Stream

4.8 Stripper

Water system

Molar flow (mol/hr)	$1.27\text{x}10^7$
C_p water (kJ/molK)	0.07524
T_1 (˚C)	60
T_2 (˚C)	?

Table 41: Water System

$$\mathbf{Q_{water} = m \text{ x } C_p \text{ x } (T_1 - T_2)}$$

Air System

Molar flow (mol/hr)	$6\text{x}10^6$
Cp (kJ/molK)	0.02917
T_1 (˚C)	25
T_2 (˚C)	?

Table 42: Air System

$$\mathbf{Q_{air} = mC_p(T_2 - T_1)}$$

According to the law of conservation of energy, $Q_{air} = Q_{water}$ and the outlet temperature of both the streams leaving the stripper would be equal to the temperature at which the stripper is operating so by taking T2 common and equating the equations we get:

$$\mathbf{T_2 = ((m_{water}Cp_{water}T_{1water}) + (m_{air}Cp_{air}T_{1air}))/((Cp_{water}T_{1water}) + (Cp_{air}T_{1air}))}$$

By putting the values in the above equation, we get

$\mathbf{T_2 = 54.53}$ **˚C**

$Q_{water} = mC_p(T_2 - T_1)$

$= 1.27\text{x}10^7 \text{ x } 0.075 \text{ x } (60 – 54.53) =$ $\mathbf{5.23 \text{ x } 10^6}$ **(kj/hr)**

$Q_{air} = mC_p(T_2 - T_1)$

$= 6\text{x}10^6 \text{ x } 0.0291 \text{ x } (54.53 – 25) =$ $\mathbf{5.23 \text{ x } 10^6}$ **(kj/hr)**

$$\mathbf{\rightarrow Q_{water} = Q_{air}}$$

5 Design

5.1 Heat Exchanger Design:

Exchanger type: Shell and tube Heat Exchanger

Shell Passes = 1

Tube Passes = N_p = 2

Shell Side Fluid: L.S Steam

Tube Side Fluid: Ammonical Water

Shell inlet temperature = T_1 = 180 ^oC

Shell outlet temperature = T_2 = 147 ^oC

Tube inlet temperature = t_1 = 30 ^oC

Tube outlet temperature = t_2 = 60 ^oC

C_p of LS Steam = 2.17 kJ/kgmol. ^oC

C_p of ammonical water = 4.18 kJ/kgmol. ^oK

Molar flow rate of shell side = 5.433×10^6 kg/d

Molar flow rate of tube side = 63500 kg/sec

$Q_{tube} = m.C_{p.}\Delta T_m$

= 63500 x 4.18 x (60 - 25)

= 2580.569 kW

$Q = U.A.\Delta T_m$

$\Delta T_m = F_t.LMTD$

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

$= 64.48\ ^\circ C$

For the Calculation of Ft:

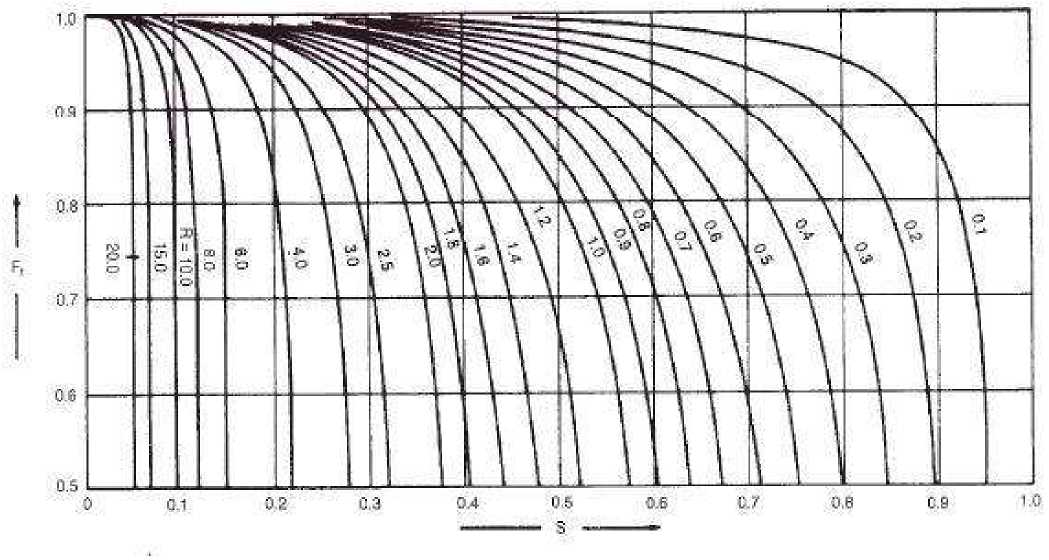


Figure 6: Ft Factor Graph

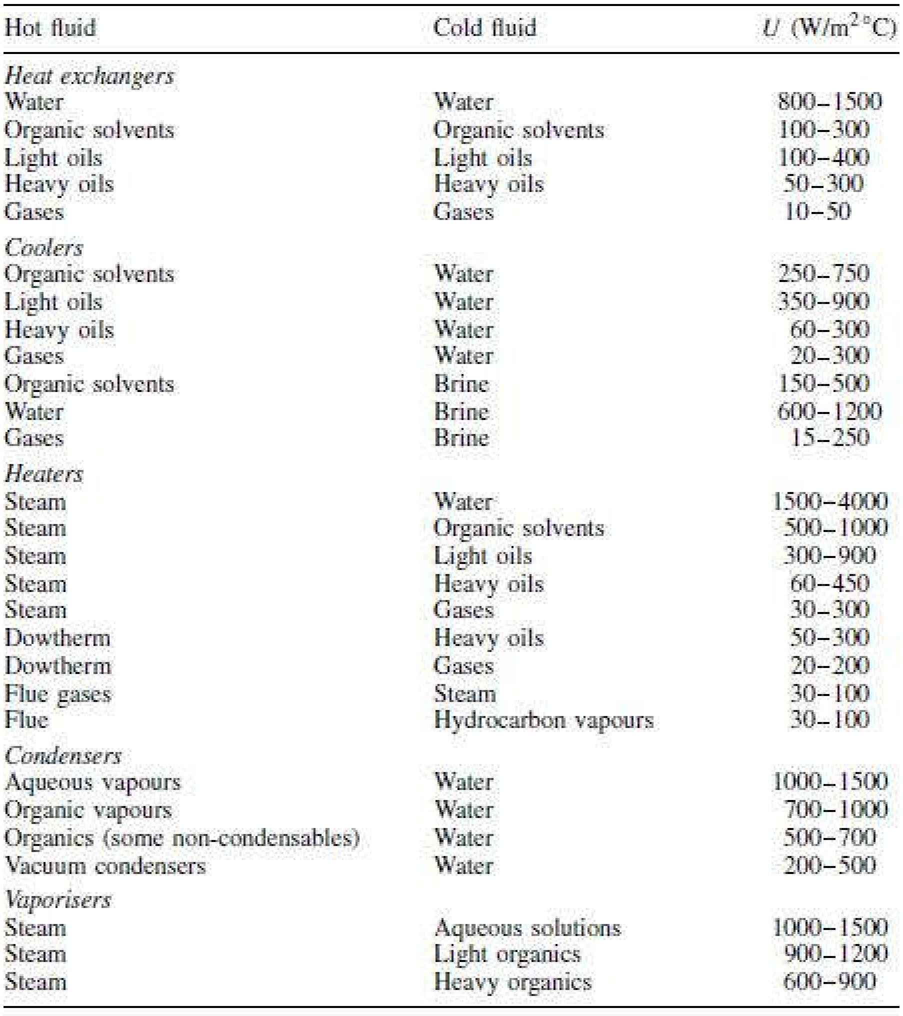
From Figure, for 1 shell pass and 2 tube passes using values of S and R, we get $F_t = 0.93$

$\Delta T_m = F_t.LMTD$

$\Delta T_m = 0.93x64.48 = 60.38\ ^o C$

Overall heat transfer coefficient for this system is assumed using the table given below. For ammonical water system the range of overall heat transfer coefficient is between 800-1500 $W/m^2.\ ^oC$, so we have chosen the value of 1000 $W/m^2.\ ^oC$.

$U = 1000\ W/m^2.\ ^oC$



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

Figure 7: Overall Heat Transfer Co-efficient for different compounds

Heat transfer area = $Q/(U\Delta T_m)$

= 2580.569/1000x62.62

= 36.63 m^2

Assumptions: **We chose 30 mm outer diameter, 26 mm inner diameter and 4.88 m long tubes Inner diameter of tube = di = 0.026 m**

Outer Diameter of tube = do = 0.03 m

Length of tube = L = 3.66 m

Area of one tube = π x L x d = π x 3.66 x 0.02 = 0.46 m^2

Number of tubes = Heat transfer area/ area of one tube = 36.63/0.46 = 80

Use triangular pitch,

Pitch = pt = 1.25 x do = 1.25 x 0.03 = 37.5 mm

Bundle diameter = $D_b = d_o(N_t/K_1)^{1/n1}$

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

K_1 = 0.249, n_1 = 2.209

D_b= 410.98 mm

Clearance for shell diameter = 0.088 m (assumption)

Shell Diameter = Bundle Diameter + Clearance

Shell Diameter = D_o = 410 + 88 = **490.9 mm**

Baffle spacing = l_B = 0.5 x Do = 0.5 x 512.9 = 256 mm

No of baffles = length of tube/baffle spacing= 4.88/0.256 = 19

5.1.1 Tube Side Co-efficient:

Mean temperature= $(t_1+t_2)/2$ = 45 °C

Density of solution = ρ = 980 kg/m^3

Viscosity of solution = μ = 0.00089 kg/m.sec

Heat capacity of solution = C_p = 4180 J/kg °C

Thermal conductivity of solution = k = 0.6182 W/m. °C

Tube cross sectional area = $\pi di^2/4$ = 531 mm^2

Tubes per pass = total tubes / 2 = 80/2 = 40

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

Total flow area = 0.000531 x 40

Total flow area = 0.021 m^2

Mass velocity = Mass flow rate / flow area = 2966 kg/sec.m^2

Linear velocity = u_t = (mass velocity / density) = 2966 / 980 = 3.026 m/sec

Reynolds Number in tubes = Ret = (densityxvelocityxdiameter)/(viscosity) = 77128.67

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

$J_H = 0.024$

Prandtl number = (Cpu)/k = 6.017

$h_i d_i/k = J_h \text{ x } Re \text{ x } Pr^{0.33} x(u/u_w)^{0.14}$

h_i = 89526.08 W/m^2.C

5.1.2 Shell-Side Co-Efficient

Mean temperature= $(T_1+ T_2)/2$ = 110°C

Density of solution = ρ = 1.496 kg/m^3

Viscosity of solution = μ = 0.000013 kg/m sec

Heat capacity of solution = Cp = 2177 J/kg^o C

Thermal conductivity of solution = K = 0.0288 W/m.^{o} C

Cross flow area = $(p_t-d_o)Ds\ Lb / p_t$ = 0.263m^2

Eq diameter = $(1.10/d_0) \text{ x } (pt^2 - 0.917\ d_0^2)$ = 21.3015 mm

Mass velocity = Mass flow rate / flow area = 100.29 kg/sec.m^2

Linear velocity = u_t = (mass velocity / density) = 1.08 m/sec

Reynolds Number in tubes = Ret = (density*velocity*diameter)/(viscosity) = 160629.09

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

J_n = 0.18

Prandtl number = (Cpu)/k = 1.005

$h_i d_i/k$ = k x jn x Re x $Pr^{0.33}$/de

h_s = 39160 W/m^2.C

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

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

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

Thermal conductivity of stainless steel = 17 W/m.° C

$1/U_o$ =0.001026

U_o = 975 W/m^2.C

5.1.3 Pressure Drop

5.1.3.1 Tube Side Pressure Drop

Reynolds number = 77128.7

From graph between Re and J_f,

Friction factor = J_f = 0.0024

Figure 8: Pressure Drop Across Heat Exchanger

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

ΔP = 0.897 kPa. (Acceptable range)

5.1.3.2 Shell Side Pressure Drop

Linear velocity = 1.06 m/s

Friction factor = J_f = 0.0043

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

ΔP = 3.65 kPa (acceptable range)

5.2 Stripper Design

Material of construction is “carbon steel” because less amount of ammonia is present in the system, if we chose stainless steel that would be very costly.

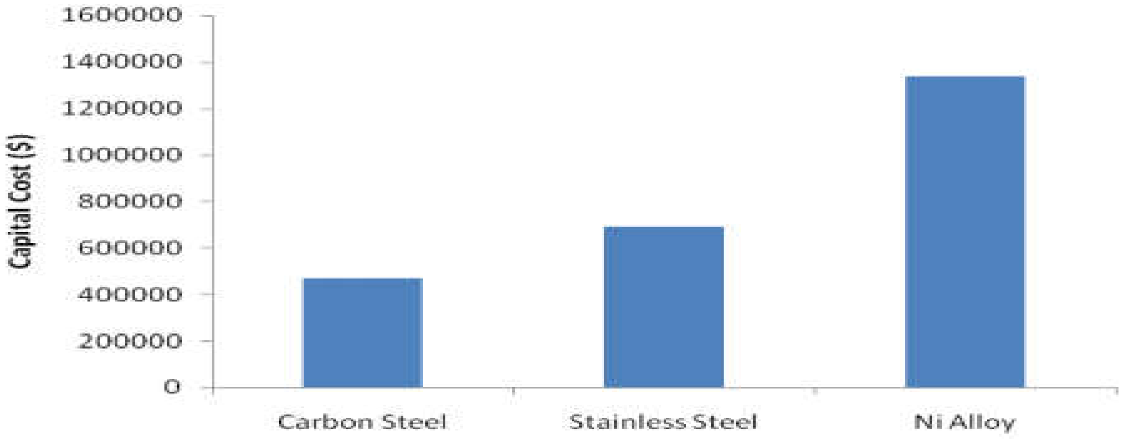


Figure 9: Cost Comparison of Different Materials

Material of carbon steel chosen is **A-53 A** because it is capable of handling large shear stress with little deformation upto 345°C.

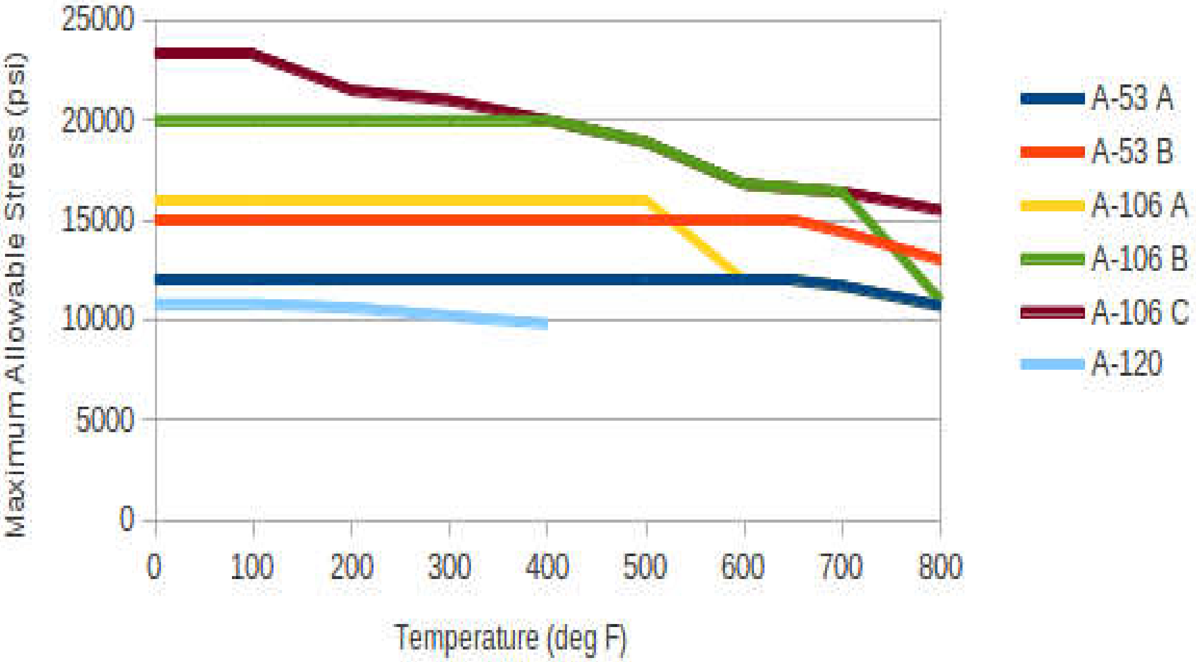


Figure 10: Stress vs Temperature Graph

5.2.1 Packing

We are using Pall rings of stainless steel type as our packing. The arrangement of our packing is random instead of structured because as we are designing a new column we have to keep the economic balance too. The advantages of random packing are:

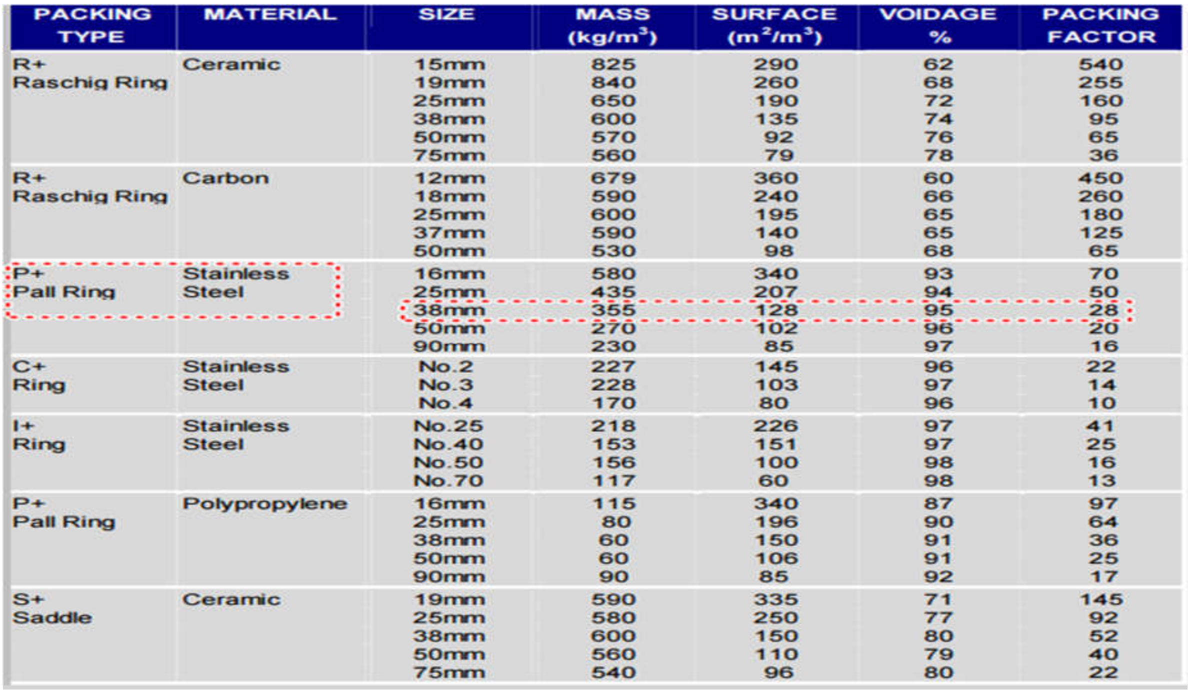
- Low cost
- High liquid loads
- Ease of replacement and maintenance
- High pressure applications

We are using Pall rings of stainless steel type because of the following conditions:

- It provides high capacity and low pressure-drop
- High mechanical strength
- High temperature applications
- Wetting is easy
- Probability of breakage is low

5.2.2 Specification of Pall Rings

In the table given below we are provided with the specifications of different types of packing materials. As the diameter would increase the pressure drop would decrease but the specific area required to provide greater retention time would decrease. Hence, we have chosen 50 mm as our diameter to compensate between the pressure drop and retention time.



PACKING TYPE	MATERIAL	SIZE	MASS (kg/m³)	SURFACE (m²/m³)	VOIDAGE %	PACKING FACTOR
R+ Raschig Ring	Ceramic	15mm	825	290	62	540
		19mm	840	260	68	255
		25mm	650	190	72	160
		38mm	600	135	74	95
		50mm	570	92	76	65
		75mm	560	79	78	36
R+ Raschig Ring	Carbon	12mm	679	360	60	450
		18mm	590	240	66	260
		25mm	600	195	65	180
		37mm	590	140	65	125
		50mm	530	98	68	65
P+ Pall Ring	Stainless Steel	16mm	580	340	93	70
		25mm	435	207	94	50
		38mm	355	128	95	28
		50mm	270	102	96	20
		90mm	230	85	97	16
C+ Ring	Stainless Steel	No.2	227	145	96	22
		No.3	228	103	97	14
		No.4	170	80	96	10
I+ Ring	Stainless Steel	No.25	218	226	97	41
		No.40	153	151	97	25
		No.50	156	100	98	16
		No.70	117	60	98	13
P+ Pall Ring	Polypropylene	16mm	115	340	87	97
		25mm	80	196	90	64
		38mm	60	150	91	36
		50mm	60	106	91	25
		90mm	90	85	92	17
S+ Saddle	Ceramic	19mm	590	335	71	145
		25mm	580	250	77	92
		38mm	600	150	80	52
		50mm	560	110	79	40
		75mm	540	96	80	22

Figure 11: Packing Rings' Table

5.2.3 Calculation of Stripper Design

For the stripper column now we will calculate the diameter and height required for the column.

5.2.4 Diameter Calculations

Liquid Mass Flow rate= m_L= 228600 Kg/h

Density of Liquid= $\rho_L = 980\ kg/m^3$

Density of Water= $\rho_{L,H2O} = 999.5\ kg/m^3$

Feed Air Flow Rate = m_G= 130500 kg/h

Density of Air = $\rho_G = 1.18\ kg/m^3$

Average dynamic viscosity of Liquid= $\eta_L = 0.000466$ Pa.s

Fraction of Flooding = f = 0.7

We know that,

$\mathbf{F_{LG} = (m_L\sqrt{\rho_G}) / (m_G\sqrt{\rho_L})}$

Putting values → $F_{LG} = 0.06$

$\mathbf{Y = \exp[-3.7121 - 1.0371 \ln F_{LG} - 0.1501 (\ln F_{LG})^2 - 0.007544 (\ln F_{LG})^3]}$

Putting values → $Y = 0.17$

$\mathbf{F_\rho = -0.8787 + 2.6776 (\rho_{L,H2O}/\rho_L) - 0.6313 (\rho_{L,H2O}/\rho_L)^2}$

Putting values → $F_\rho = 1.18$

$\mathbf{F_\eta = 0.96\, \eta_L^{0.19}}$

Putting values → $F_\eta = 0.22$

Packing factor chosen for the pall rings **is $F_p =$ 27 (1/m)**

All of the above parameters are calculated for finding the flooding velocity $\mathbf{W_{G,max}}$

$\mathbf{Y = (W_{G,max}^2 F_p \rho_G F_\rho F_\eta) / (g\, \rho_{L,H2O})}$

Putting values → $W_{G,max} = 14.5$ (m/s)

Now the diameter of the column is a function of flooding velocity. Hence for diameter:

$\mathbf{D = 2/\sqrt{\pi} \times \sqrt{(m_G)/(\rho_G f W_{G,max})}}$

Putting values → **D= 1.8 m**

5.2.4.1 Column Height Calculations

Height of the column is the function of number of transfer units (NTU) and height of the transfer units (HTU).

NTU is calculated as a function of stripping factor which has the following formula:

NTU$=1/(1-S^{-1})$ **x** $\ln[(1-S^{-1})(x_{in}/x_{out})$

Where S is known as the stripping factor which determines the efficiency of the column

$S = KV/L = 1.15$

$K = H/P = 3.38$

L is the liquid flow rate and V is the Air flow rate.

Putting the values, we get **NTU = 12**

Total Height (Z):
HTU is obtained from the table given below

Z = HTU x NTU
= 0.6 x 12
= 7.2 (m)

This is the packed height of column. In order to accommodate the demister pads and liquid distributors in the column we add 1.25 m on both top and bottom of column. So overall height (Z):

$Z = \text{Packed Height} + H_{min}$
= 7.2 + 2
= 9.7 meters

Packing Size (mm)	HTU (m)
25	0.3 - 0.6
38	0.5 - 0.75
50	0.6 - 1

Table 43: Packing vs Heat Transfer Units

5.2.5 Column Internals

Distribution Plates:

Volumetric Flowrate of condensate = V = 233 m^3/hr
Diameter = D = 1.8 m
Area = A = $\pi D^2/4$ = 84.88 m^2

Specific Liquid Load = V/A = 9.59 m^3/m^2h

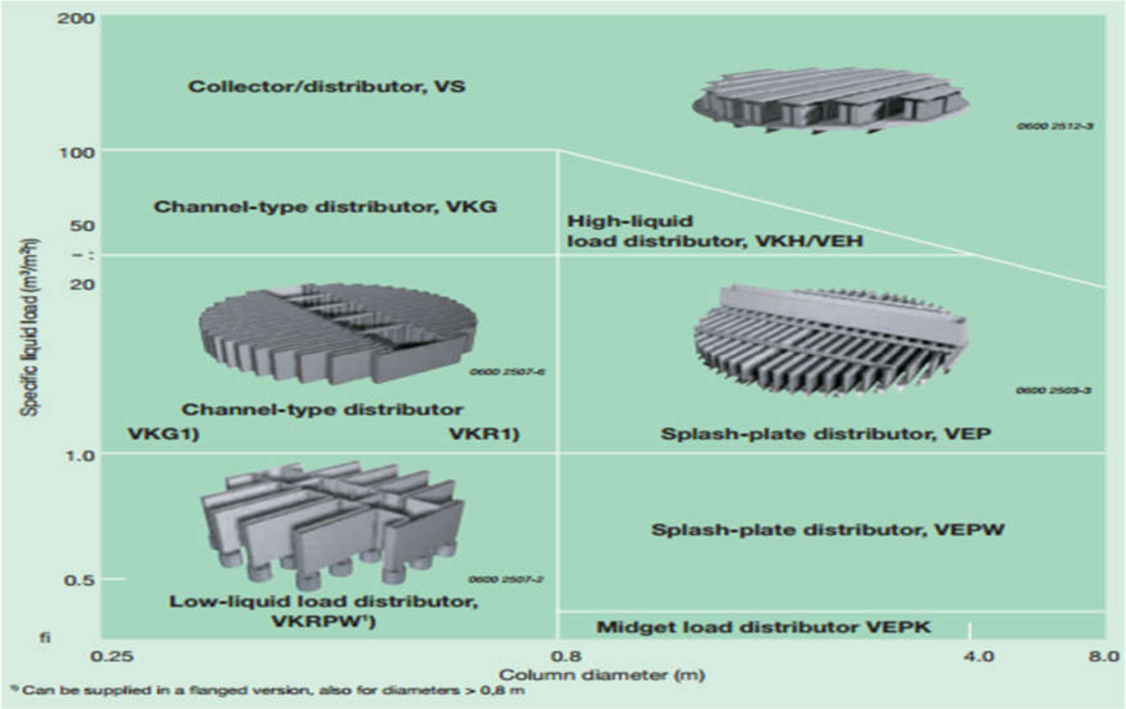
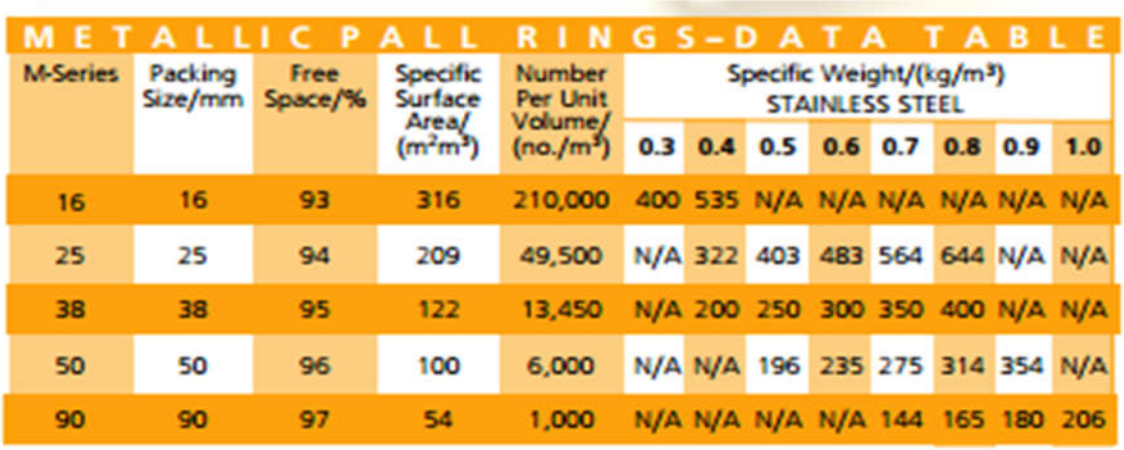


Figure 12: Distributor Types

Channel Type distributor VKH/VEH would be installed according to the above graph. These distributors are really important as we convert water in the droplet form which then contacts with the air and high mass transfer takes place due to minimum surface tension.

Number of Packings:

Diameter = D = 1.8 m
Packed Height = H = 7.2 m
Volume = $\pi D^2h/4$ = 18.31 m^3



METALLIC PALL RINGS-DATA TABLE

M-Series	Packing Size/mm	Free Space/%	Specific Surface Area/ (m^2m^3)	Number Per Unit Volume/ (no./m^3)	Specific Weight/(kg/m^3) STAINLESS STEEL							
					0.3	0.4	0.5	0.6	0.7	0.8	0.9	1.0
16	16	93	316	210,000	400	535	N/A	N/A	N/A	N/A	N/A	N/A
25	25	94	209	49,500	N/A	322	403	483	564	644	N/A	N/A
38	38	95	122	13,450	N/A	200	250	300	350	400	N/A	N/A
50	50	96	100	6,000	N/A	N/A	196	235	275	314	354	N/A
90	90	97	54	1,000	N/A	N/A	N/A	N/A	144	165	180	206

Figure 13: Pall Ring Data

From Table for Metal Pall Rings of Dia 50 mm

Number per unit volume = 6000 no/m^3

Thus number of Packings = **109874**

5.2.6 Total Number of Stages Obtained:

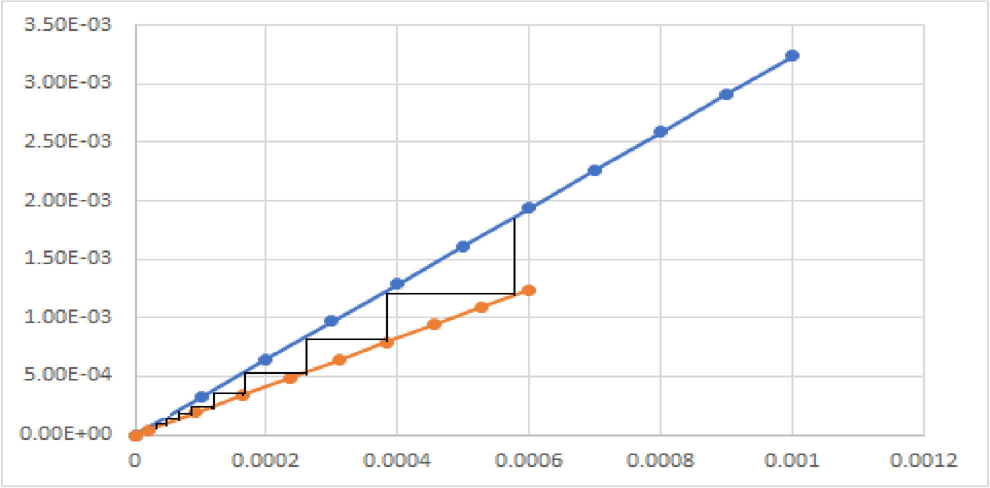


Figure 14: Ammonia in Air vs Ammonia in Liquid Phase

x-axis represents the amount of ammonia in liquid phase

y-axis represents the amount of ammonia in Air

When the graph was plotted b/w these two variables the number of stages were calculated to be **8**

Total stages = 8.

5.3 Absorber Design

Material of construction is "stainless steel" because the system is acidic, if we chose carbon steel the maintenance cost will be very high.

Material of carbon steel chosen is **A-53 A** because it is capable of handling large shear stress with little deformation upto 345C.

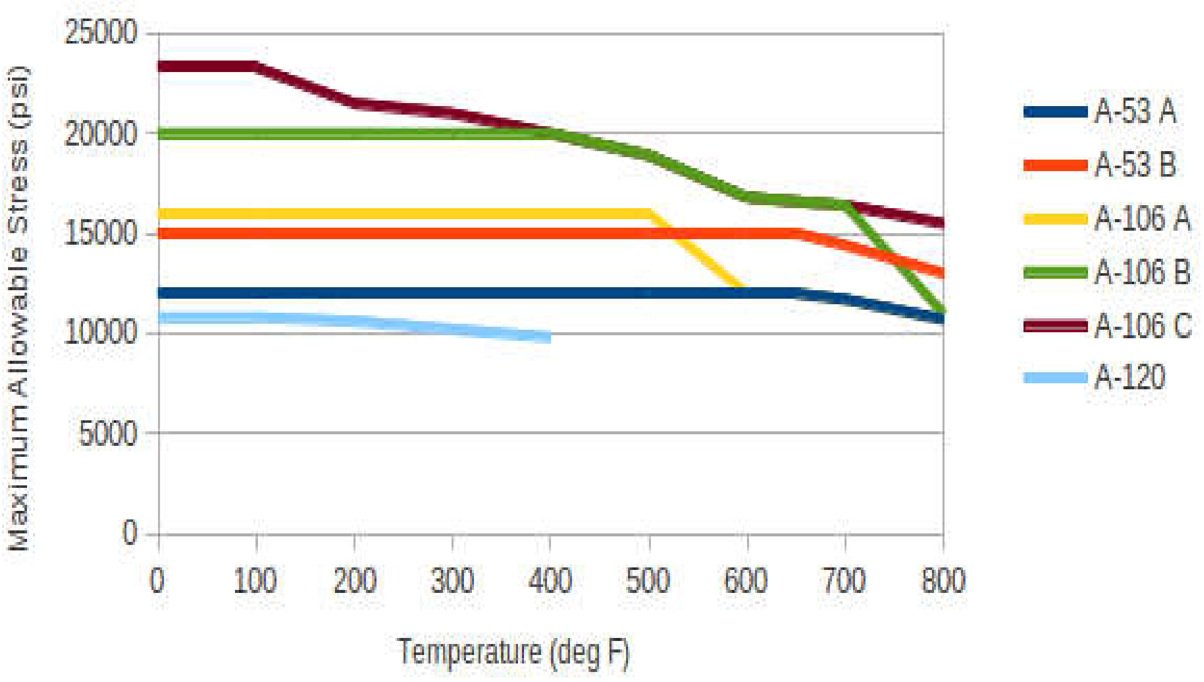


Figure 15: Stress vs Temperature

5.3.1 Packing

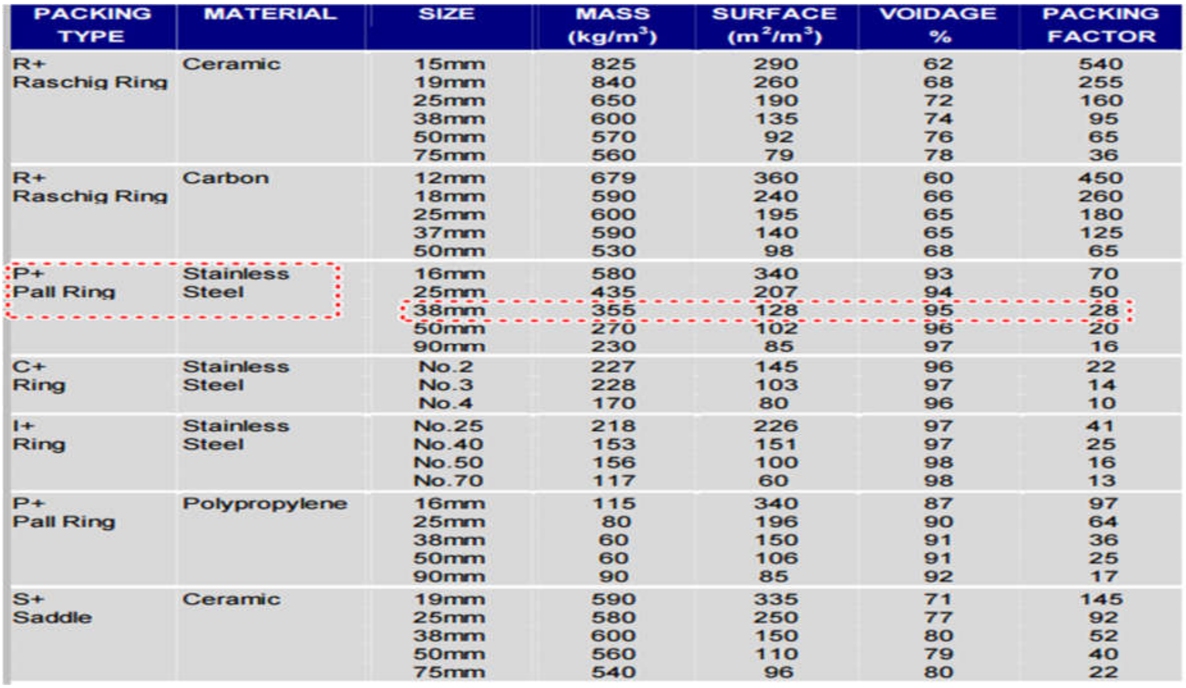
Since the absorption column involves a chemical reaction hence a packing which can withstand a chemical reaction in an acidic medium must be used. According to Koch-Glich manufacturers Flexiring random packing is most suitable in chemical reaction applications. Hence, we are using these in our absorption column. Also, the specifications of flexi rings are similar to that of pall rings.

We are using Flexiring of stainless steel type because of the following conditions:

- They are industrially equivalent to pall rings
- They provide high liquid hold up along with high residence time unlike other packing.
- Since the liquid hold up is high hence the absorption efficiency is higher
- Wetting is easy
- Probability of breakage is low

5.3.2 Specification of Flexi-ring:

We would read flexiring as equivalent to pall rings because on industrial scale for chemical reactions pall rings are replaced by flexirings but the specifications remain the same. Size of 50mm is chosen for absorber.



PACKING TYPE	MATERIAL	SIZE	MASS (kg/m^3)	SURFACE (m^2/m^3)	VOIDAGE %	PACKING FACTOR
R+ Raschig Ring	Ceramic	15mm	825	290	62	540
		19mm	840	260	68	255
		25mm	650	190	72	160
		38mm	600	135	74	95
		50mm	570	92	76	65
		75mm	560	79	78	36
R+ Raschig Ring	Carbon	12mm	679	360	60	450
		18mm	590	240	66	260
		25mm	600	195	65	180
		37mm	590	140	65	125
		50mm	530	98	68	65
P+ Pall Ring	Stainless Steel	16mm	580	340	93	70
		25mm	435	207	94	50
		38mm	355	128	95	28
		50mm	270	102	96	20
		90mm	230	85	97	16
C+ Ring	Stainless Steel	No.2	227	145	96	22
		No.3	228	103	97	14
		No.4	170	80	96	10
I+ Ring	Stainless Steel	No.25	218	226	97	41
		No.40	153	151	97	25
		No.50	156	100	98	16
		No.70	117	60	98	13
P+ Pall Ring	Polypropylene	16mm	115	340	87	97
		25mm	80	196	90	64
		38mm	60	150	91	36
		50mm	60	106	91	25
		90mm	90	85	92	17
S+ Saddle	Ceramic	19mm	590	335	71	145
		25mm	580	250	77	92
		38mm	600	150	80	52
		50mm	560	110	79	40
		75mm	540	96	80	22

Figure 16: Packing Data

5.3.3 Calculations of Absorber Design

Now, we will calculate the height and diameter of our absorber column.

5.3.4 Diameter Calculation:

Liquid Mass Flow rate= m_L= 5.7×10^5 Kg/h
Density of Liquid= $\rho_L = 980$ kg/m^3
Density of Water= $\rho_{L,H2O} = 999.5$ kg/m^3
Feed Air Flow Rate = m_G= 130500 kg/h
Density of Air = $\rho_G = 1.18$ kg/m^3
Average dynamic viscosity of Liquid= $\eta_L = 0.000466$ Pa.s
Fraction of Flooding = f = 0.7

We know that,

$$\mathbf{F_{LG} = (m_L\sqrt{\rho_G}) / (m_G\sqrt{\rho_L})}$$

Putting values → $F_{LG} = 0.15$

$$\mathbf{Y= \exp[- 3.7121 - 1.0371 \ln F_{LG} - 0.1501 (\ln F_{LG})^2 - 0.007544 (\ln F_{LG})^3]}$$

Putting values → Y= 0.101

$$\mathbf{F_\rho = - 0.8787 + 2.6776 (\rho_{L,H2O}/\rho_L) - 0.6313 (\rho_{L,H2O}/\rho_L)^2}$$

Putting values → $F_\rho = 1.196$

$$\mathbf{F_\eta = 0.96\, \eta_L^{0.19}}$$

Putting values → $F_\eta = 0.22$

Packing factor chosen for the pall rings is, $\mathbf{F_p = 28\ (1/m)}$

All of the above parameters are calculated for finding the flooding velocity, $\mathbf{W_{G,max}}$

$$\mathbf{Y= (W_{G,max}^2\, F_p\, \rho_G\, F_\rho\, F_\eta) / (g\, \rho_{L,H2O})}$$

Putting values → $W_{G,max} = 10.86$ (m/s)

Now the diameter of the column is a function of flooding velocity hence for dia;

$$\mathbf{D = 2/\sqrt{\pi} \times \sqrt{(m_G)/(\rho_G\, f\, W_{G,max})}}$$

Putting values → **D= 2.14 m**

5.3.4.1 Column Height Calculations

By keeping the pH low, the absorption of NH_3 to form NH_4^+ will be irreversible and the equilibrium line will be almost horizontal and separation very favorable

For the special case of irreversible absorption:
$m = 0$
$x_{in} = 0$ (no ammonia in inlet)

$$NTU = 1/(1-mV/L) \times Ln[(1-mV/L) \times (Y_{IN}-mx_{out}/Y_{OUT}-mx_{in}) + mV/L]$$
$$= \mathbf{\ln Y_{in}/Y_{out}}$$

$Y_{in} = 1.23 \times 10^{-3}$

$Y_{out} = 1.23 \times 10^{-5}$

NTU = 4.6 ≈ 5

Total Height (Z):
HTU is obtained from the table given below

Packing size (mm)	HTU(m)
25	0.3-0.6
38	0.5-0.75
50	0.6-1

Table 44: Packing Size vs Heat Transfer Units

Z = HTU x NTU
= 1 x 5
= 5 meters

This is the packed height of column. In order to accommodate the demister pads and liquid distributors in the column we add 1.25m on both top and bottom of column.

So overall height (Z),

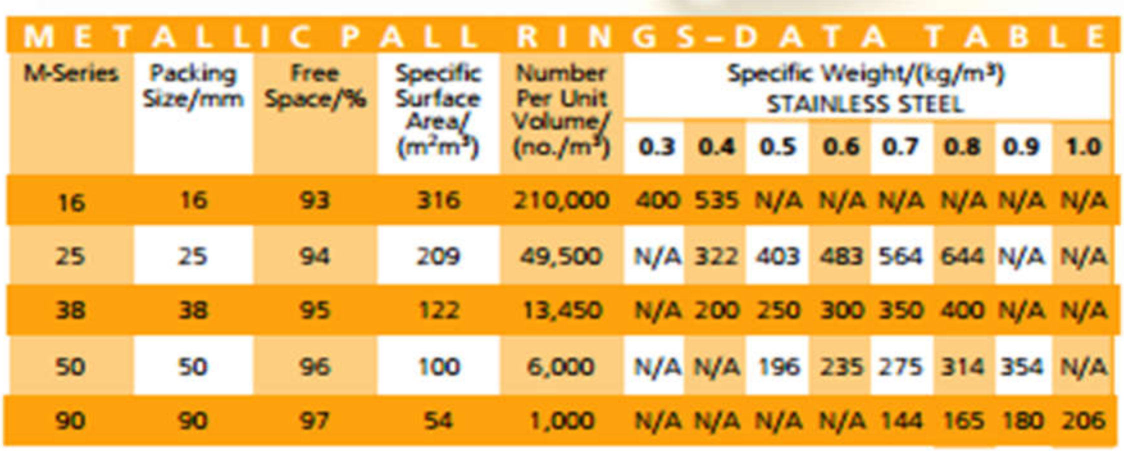
Z = Packed Height + H_{min}

= 5 + 2.5

= 7.5 meters

5.3.5 Number of Packings

Diameter = D = 2.14 m
Packed Height = H = 5 m
Volume = $\pi D^2h/4 = 15.7\ m^3$



METALLIC PALL RINGS-DATA TABLE

M-Series	Packing Size/mm	Free Space/%	Specific Surface Area/ (m^2m^3)	Number Per Unit Volume/ (no./m^3)	Specific Weight/(kg/m^3) STAINLESS STEEL							
					0.3	0.4	0.5	0.6	0.7	0.8	0.9	1.0
16	16	93	316	210,000	400	535	N/A	N/A	N/A	N/A	N/A	N/A
25	25	94	209	49,500	N/A	322	403	483	564	644	N/A	N/A
38	38	95	122	13,450	N/A	200	250	300	350	400	N/A	N/A
50	50	96	100	6,000	N/A	N/A	196	235	275	314	354	N/A
90	90	97	54	1,000	N/A	N/A	N/A	N/A	144	165	180	206

Figure 17: Pall Rings' Data

From Table for Metal Rings of Dia 50 mm
Number per unit volume = 6000 no/m^3
Thus number of Packings = **94200**

6 Costing

First of all the cost of all the major equipment is calculated. By using the area of heat exchanger, heights for absorber and stripper we find out the cost of equipment from the graphs which are shown below. This gives a preliminary cost estimate for the equipment. The base date is mid-2004, and the prices are accurate within ±25 per cent. The graphs are taken from Richardson and Coulson's Chemical Engineering Design Vol. 6.

6.1 Purchase Cost of Major Equipment Items

6.1.1 Cost of Heat Exchanger

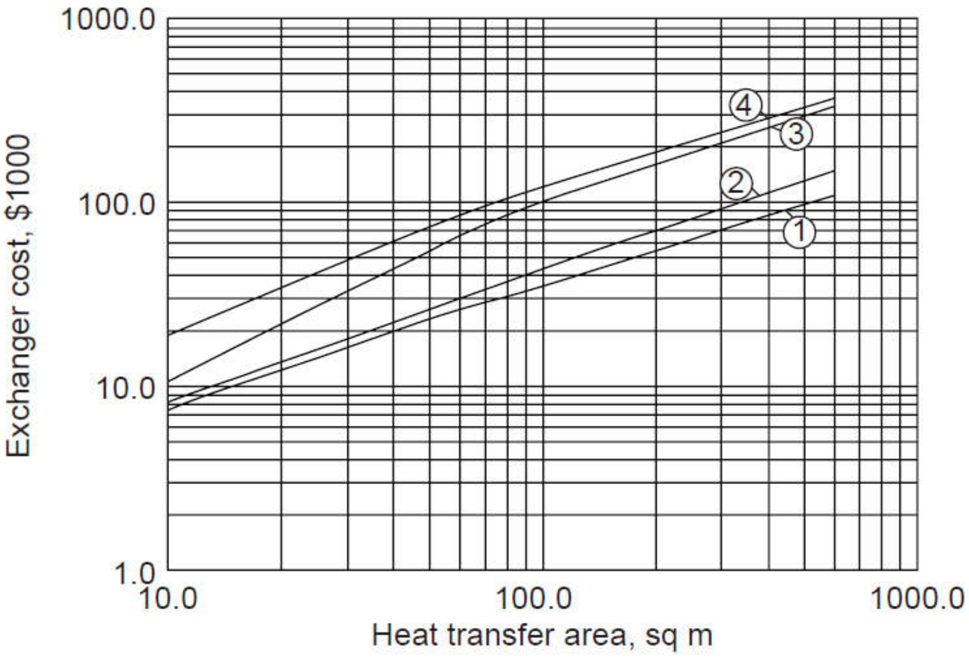
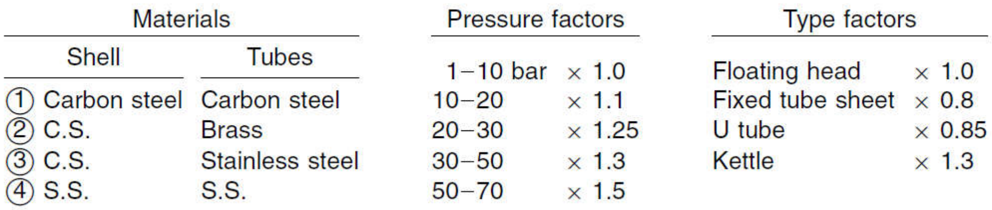


Figure 18: Exchanger Cost vs Heat Tranfer Area



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 19: Costing Factors

Purchase cost = Bare cost from graph x Material factor x Pressure factor

Heat Exchanger is a Shell and Tube type with floating head.

Area of Heat Exchanger = 36.63 m^2

Material of construction of Shell = Carbon Steel

Material of construction of Tubes = Stainless Steel

Pressure factor = 1

Type factor = 1

Bare cost from graph = 38,000 $

Cost of Heat Exchanger = Bare cost from graph x Pressure factor x Type factor

= 38,000 x 1 x 1

= 38,000 $

Cost of Columns:

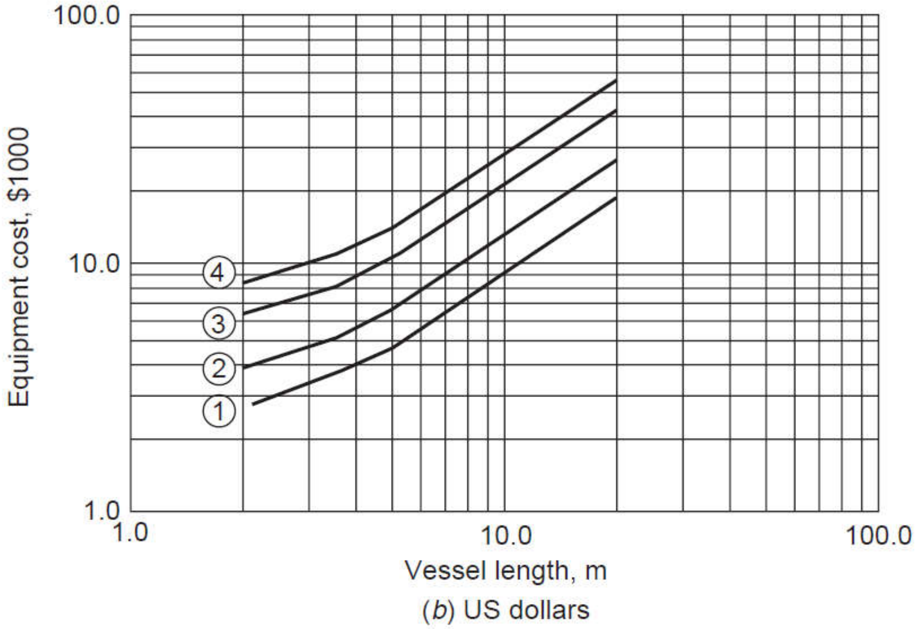
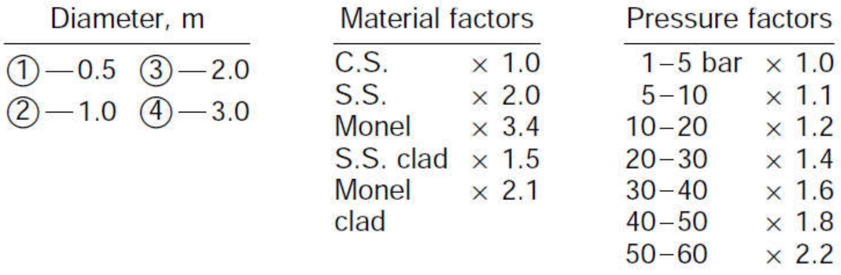


Figure 20: Stripper Equipment Cost vs Vessel Length

Cost of Absorber and Stripper columns

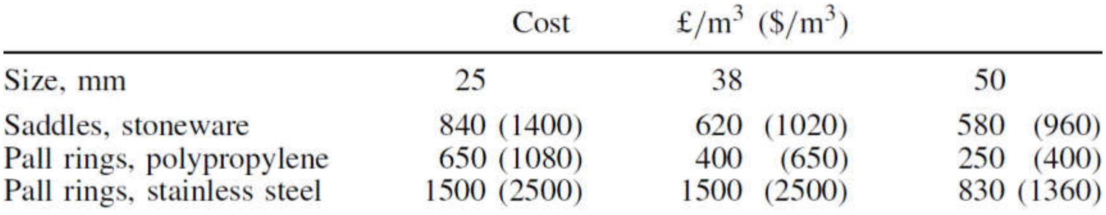


Diameter, m	
①—0.5	③—2.0
②—1.0	④—3.0

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

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

Figure 21: Costing Factors' Table



	Cost £/m^3 (\$/m^3)		
Size, mm	25	38	50
Saddles, stoneware	840 (1400)	620 (1020)	580 (960)
Pall rings, polypropylene	650 (1080)	400 (650)	250 (400)
Pall rings, stainless steel	1500 (2500)	1500 (2500)	830 (1360)

Figure 22: Packing Cost

Purchase cost = Bare cost from figure x Material factor x Pressure factor

6.1.2 Cost of Stripper column

Height of stripper column = 9.7 meters

Packed Height of column = 7.2 meters

Diameter of stripper = 1.8 meters

Material of construction of stripper = Carbon Steel

Bare vessel cost from graph = 35,000 $

Pressure factor = 1

Material factor = 1

Purchase cost = Bare Cost from figure x Material factor x Pressure factor
= 35,000 x 1 x 1
= 35,000 $

Packing cost:

Pall Rings made of stainless steel are used as packing material. Their size chosen is 50 mm.

Cost = 1360 $\$/m^3$

Volume occupied by packing = $\pi D^2 h/4$ = 18.31 m^3

Cost of packing = 18.31 x 1360
= 24,901.6 $

Total Cost of column = 35,000 $ + 24,901.6 $

= 59,901.6 $

6.1.3 Cost of Absorber Column

Height of Absorber column = 7.5 meters

Packed Height of Absorber = 5 meters

Diameter of Absorber = 2.14 meters

Material of construction of Absorber = Stainless Steel

Bare vessel cost from graph = 25,000 $

Material factor = 2

Pressure factor = 1

Purchase cost = Bare Cost from figure x Material factor x Pressure factor

= 25,000 x 2 x 1

= 50,000 $

Packing cost:

Pall Rings made of stainless steel are used as a packing material. Their size chosen is 50 mm.

Cost = 1360 $/m^3

Volume occupied by packing = $\pi D^2 h/4$ = 17.89 m^3

Cost of packing = 17.89 x 1360
= 24,330.4 $

Total Cost of column = 50,000 $ + 24,330.4 $

= 74330.4 $

6.1.4 Cost of Air Blower

Purchase cost of Air Blower which blows air at the rate of $6x10^6$ mol/hr from internet:

Type = Centrifugal

Capacity = $6x10^6$ mol/hr

Cost = 22,310 $

→ **Cost of Purchasing Major Equipment Items (PCE):**

Hence, the total purchase cost of the major equipment items used for Ammonia removal and recovery process is:

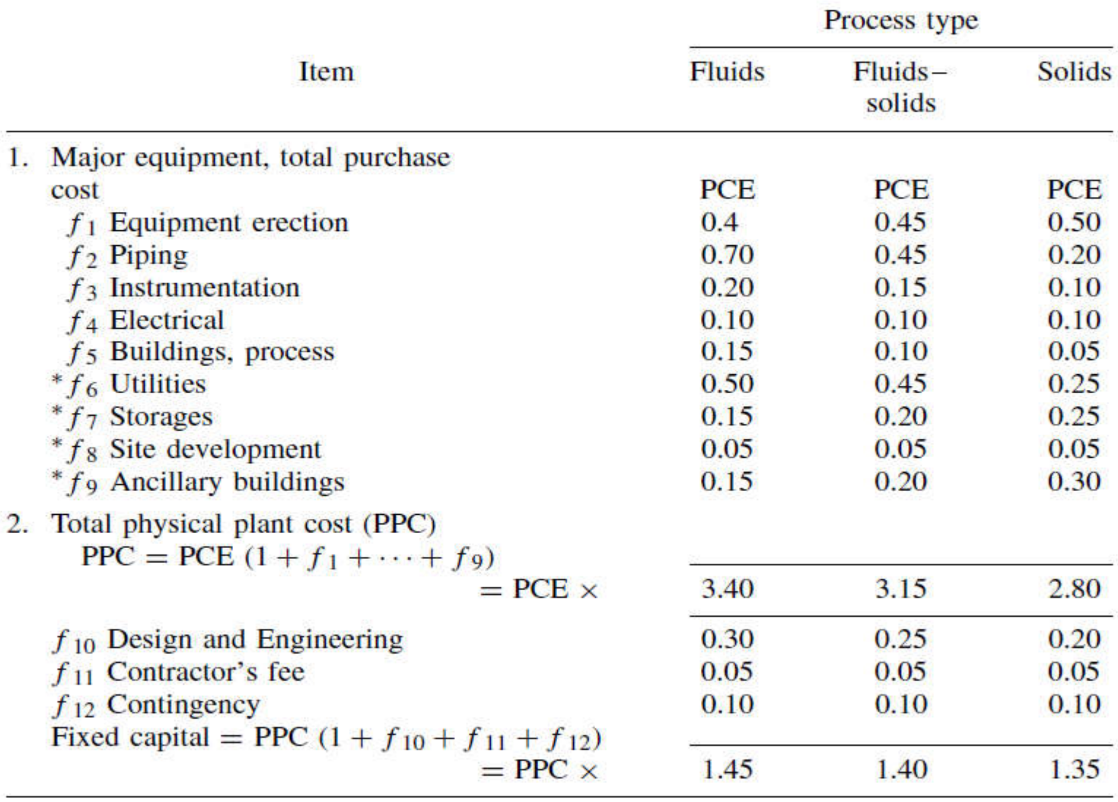
Sr. no	Equipment	Cost ($)
1.	Heat Exchanger	38,000
2.	Stripper	59,901.6
3.	Air Blower	22,310
4.	Absorber	74330.4

Table 45: Purchase Cost of Major Equipment

PCE = $ 194,542

6.2 Estimation of Fixed Capital Cost

Now, for the determination of **Fixed Capital Cost** for the system the following table is used.



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

Figure 23: Lang Factors

Since, our process is Fluids Type as the system processes Ammonia, Water and Air streams so from the table the values of Lang factors chosen are stated below:

f_1 Equipment Erection	0.4
f_2 Piping	0.7
f_3 Instrumentation	0.2
f_4 Electrical	0.1
f_5 Buildings	0.5
f_6 Utilities	0.5
f_7 Storages	0.15
f_8 Site Development	0.05
f_9 Ancilliary Buildings	Not applicable

Figure 24: Selected Lang Factors for PPC

PCE = $ 194,542

Total Physical Plant Cost = PPC = PCE (1 + f_1 + f_2 + f_3 + …. + f_9)

= 194,542 x (1 + 0.4 + 0.7 + 0.2 + 0.1 + 0.5 + 0.15 + 0.05)

= 603,080.2 $

f_{10} Design and Engineering	0.3
f_{11} Contractor's fee	0.05
f_{12} Contingencies	0.1

Figure 25: Selected Lang Factors for Fixed Capital

Fixed Capital = F.C = PPC (1 + f_{10} + f_{11} + f_{12})

= 603,080.2 x (1 + 0.3 + 0.05 + 0.1)

= 874,466.29 $

Working Capital is 5% of the Fixed capital.

Hence, **Working Capital** = 0.05 x 874,466.29

= 43,723.3 $

So,

Total Investment Required for the Project = Fixed Capital + Working Capital

= 874,466.29 + 43,723.3

= 918,189.6 $

6.3 Estimation of Direct Production Cost

6.3.1 Variable Costs

Attainment = 96%

= 0.96 x 365 x 24 = 8409.6 hr/year

1. Miscellaneous Materials = 10% of Maintenance Cost
   = 0.1 x 43,723.3 $
   = 4,372.33 $
2. Steam (L.S) = 12$/t x 5.4336x$10^6$ kg/d x 1t/1000kg
   = 65,203.2 $
3. Shipping and Packaging = Not applicable

Hence,

Total variable cost = 4,372.33 $ + 65,203.2 + 0

= 69,575.53 $$

6.3.2 Fixed Costs

1. Maintenance Cost = 5% of Fixed Capital
   = 0.05 x 874,466.29
   = 43,723.3 $
2. Operating Labour = 28,000 $
3. Supervision = Not applicable
4. Plant Overheads = 50% of Operating Labor
   = 0.5 x 28,000
   = 14,000 $
5. Laboratory = 20% of Operating Labor
   = 0.2 x 28,000
   = 5,600 $
6. Capital Charges = 6% of Fixed Capital

= 0.06 x 653,661.12

= 39,219.67 $

7. Insurance = 1% of Fixed Capital

= 0.01 x 653,661

= 6,536.6112 $

Hence,

Total Fixed cost = 43,723.3 + 28,000 + 14,000 + 5,600 + 39,219.67 + 6,536.6

= 137,079.57 $

Direct Production Cost = Total Fixed Cost + Total Variable Cost

= 137,079.57 $ + 69,575.53 $

= 206,655.1 $

Sales Expense = Not Applicable

General Overhead = Not Applicable

Research and Development = Not Applicable

Therefore,

Annual Operating cost = Direct Production Cost

= 206,655.1 $

6.4 Revenue Generated

Sale of Ammonium Sulphate = 60 $/t

Ammonium Sulphate produced = $4.8\text{x}10^5$g/hr = $4.8\text{x}10^2$ kg/hr

So, cost of ammonium sulphate sold in a year = 4.8x102 x 1t/1000kg x 60 x 8409.6 hr/yr

= 282,562.6 \$/year

Water discharged from filter press and stripper = $107.7 \times 10^3 + 304.8 \times 10^3$ kg/d

= 412.5×10^3 kg/d

It is sold at 1.2 \$/tonne, so

= 412.5×10^3 x 1t/1000kg x 1d/24hr x 8409.6hr/yr x 1.2 \$/tonne

= 173,448 \$/year

Hence,

Total revenue generated = 282,562.6 + 173,448

= 456,010.56 \$/year

6.5 Profit

Profit = Total Revenue Generated - Production Cost

= 456,010.56 - 206,655.1

= 249,355.5 \$/year

6.6 Rate of Return

Rate of return = Profit / Initial Investment

= (249,355.5 / 918,189.6)x100

= 27.2%

6.7 Payback Period

Payback period is reciprocal of rate of return.

Hence,

Payback period = 100 / 27.2

= 3.68 ≈ 3.7

So, payback period is 3 years and 7 months approximately. As it is less than five years it means that we'll reach at breakeven in 3 years and seven months approximately and profit will be attained after this time. So, it is profitable to undertake this project.

7 Simulation

7.1 Aspen-Plus Model

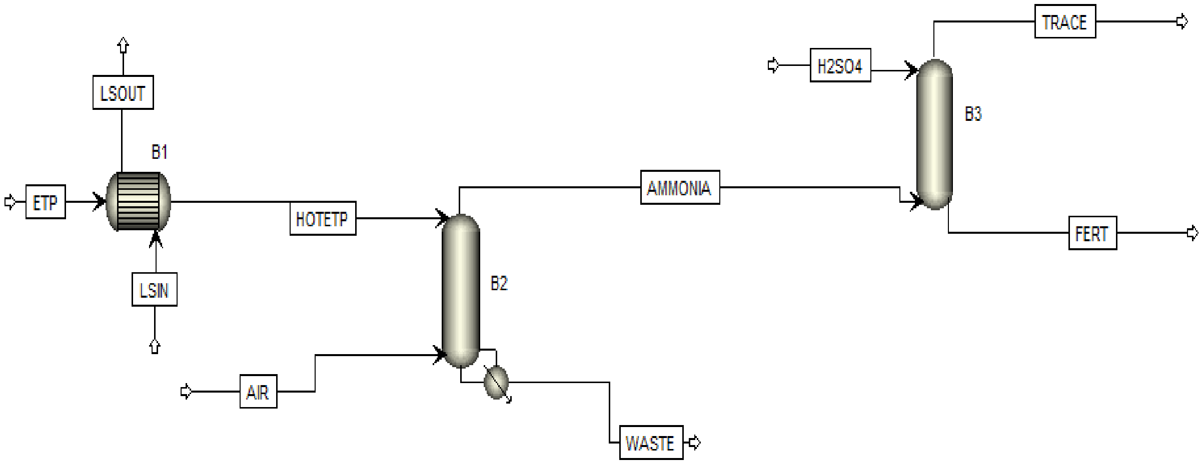
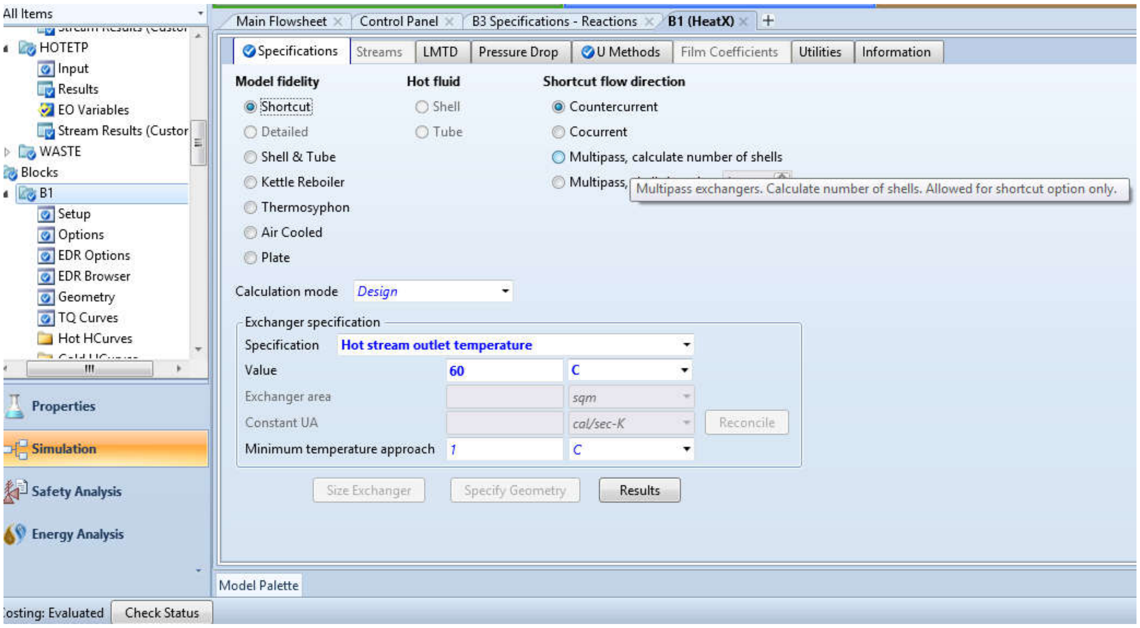


Figure 26: Aspen Plus Simulation Model

This model shows the ammonia removing section that is proposed to be added to the effluent treatment plant. After the water has been treated from the P_2O_5 and HF, it is sent into the heat exchanger where it is heated to 60^{o}C by high pressure steam. It then enters the stripper where the ammonia is separated from the water using air at ambient temperature. The water is then disposed away and the ammonia stream is further sent into the absorber where it experiences chemisorption in H_2SO_4 to produce ammonium sulfate, which is an important product and is removed at the bottom of the column whilst the trace amounts of unreacted ammonia, sulfuric acid and water are removed from the top.

7.2 Heat Exchanger (B1)



Model fidelity
- ● Shortcut
- ○ Detailed
- ○ Shell & Tube
- ○ Kettle Reboiler
- ○ Thermosyphon
- ○ Air Cooled
- ○ Plate

Hot fluid
- ○ Shell
- ○ Tube

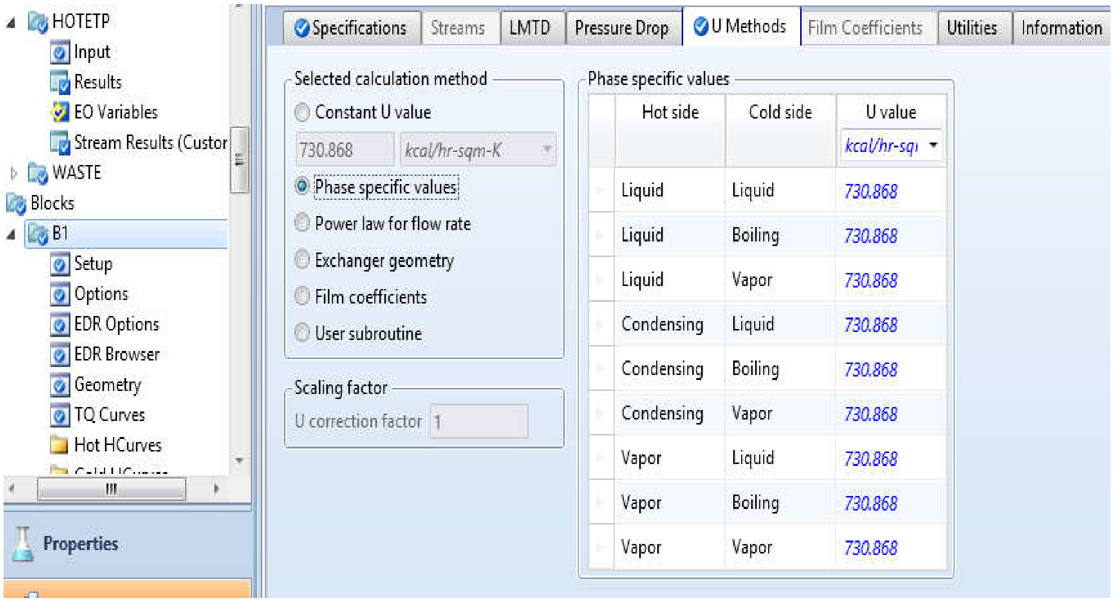
Shortcut flow direction
- ● Countercurrent
- ○ Cocurrent
- ○ Multipass, calculate number of shells
- ○ Multipass,

Multipass exchangers. Calculate number of shells. Allowed for shortcut option only.

Calculation mode: Design

Exchanger specification
- Specification: Hot stream outlet temperature
- Value: 60 C
- Exchanger area: sqm
- Constant UA: cal/sec-K
- Minimum temperature approach: 1 C

Figure 27: Heat Exchanger



Selected calculation method
- ○ Constant U value: 730.868 kcal/hr-sqm-K
- ● Phase specific values
- ○ Power law for flow rate
- ○ Exchanger geometry
- ○ Film coefficients
- ○ User subroutine

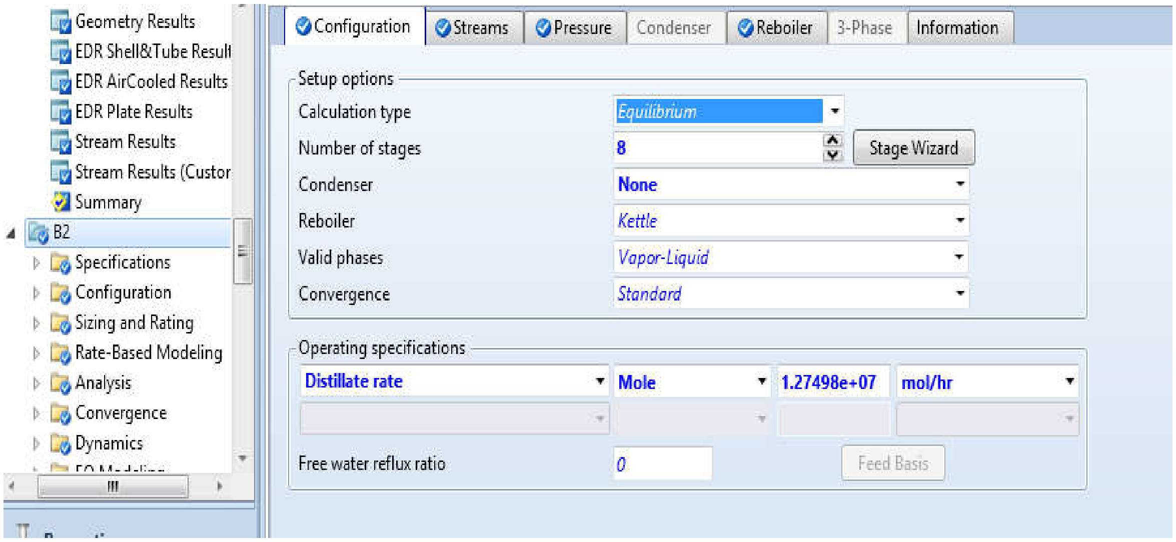
Scaling factor
- U correction factor: 1

Phase specific values

Hot side	Cold side	U value kcal/hr-sq
Liquid	Liquid	730.868
Liquid	Boiling	730.868
Liquid	Vapor	730.868
Condensing	Liquid	730.868
Condensing	Boiling	730.868
Condensing	Vapor	730.868
Vapor	Liquid	730.868
Vapor	Boiling	730.868
Vapor	Vapor	730.868

Figure 28: Heat Exchanger

7.3 Stripper (B2)



Setup options

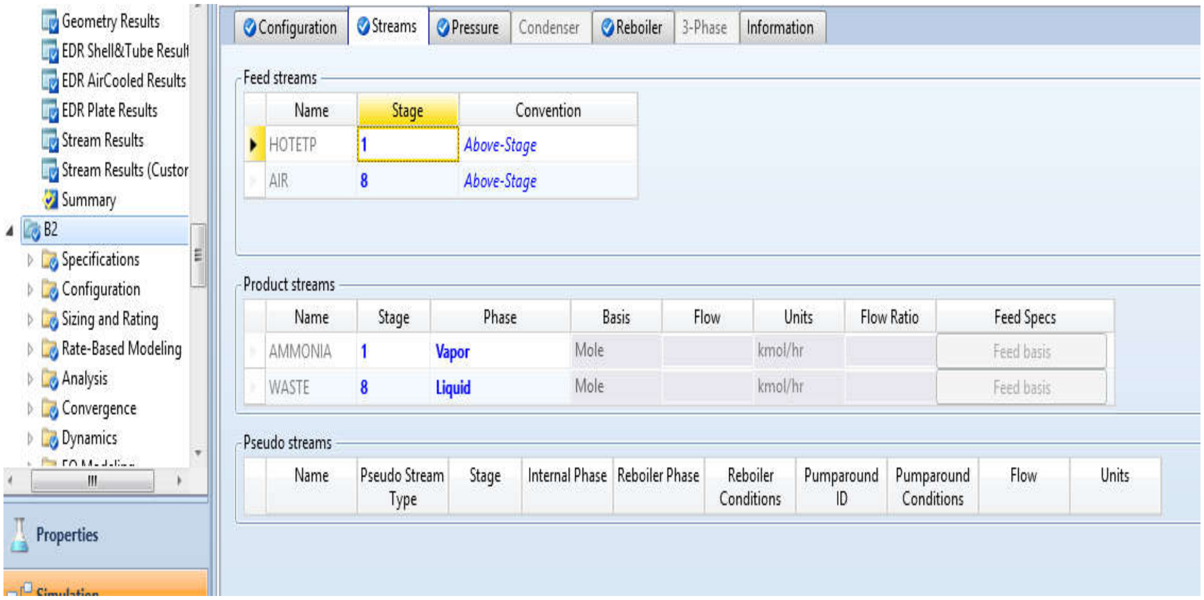
Calculation type: Equilibrium
Number of stages: 8
Condenser: None
Reboiler: Kettle
Valid phases: Vapor-Liquid
Convergence: Standard

Operating specifications

Distillate rate | Mole | 1.27498e+07 | mol/hr

Free water reflux ratio: 0

Figure 29: Stripper



Feed streams

Name	Stage	Convention
HOTETP	1	Above-Stage
AIR	8	Above-Stage

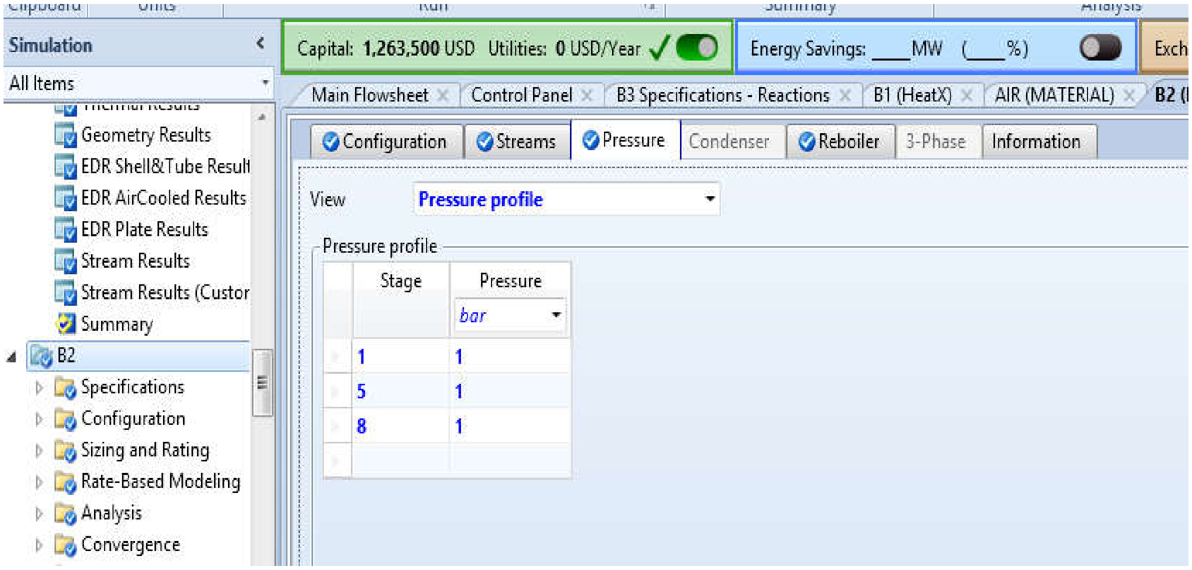
Product streams

Name	Stage	Phase	Basis	Flow	Units	Flow Ratio	Feed Specs
AMMONIA	1	Vapor	Mole		kmol/hr		Feed basis
WASTE	8	Liquid	Mole		kmol/hr		Feed basis

Pseudo streams

Name	Pseudo Stream Type	Stage	Internal Phase	Reboiler Phase	Reboiler Conditions	Pumparound ID	Pumparound Conditions	Flow	Units

Figure 30: Stripper



Capital: 1,263,500 USD Utilities: 0 USD/Year

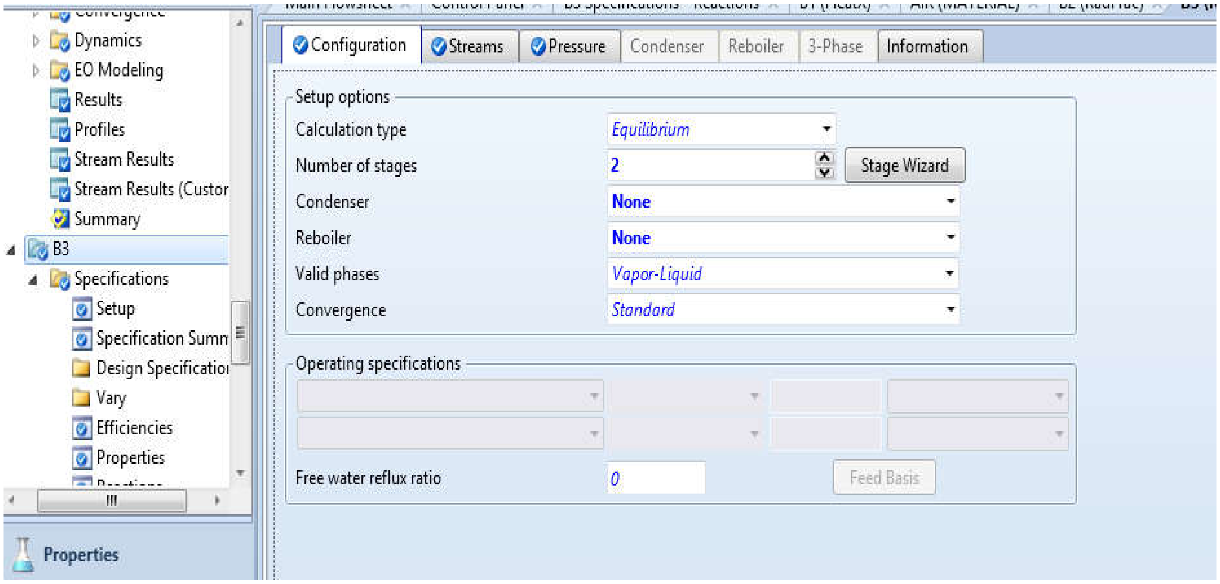
View: Pressure profile

Pressure profile

Stage	Pressure
	bar
1	1
5	1
8	1

Figure 31: Stripper

7.4 Absorber (B3):



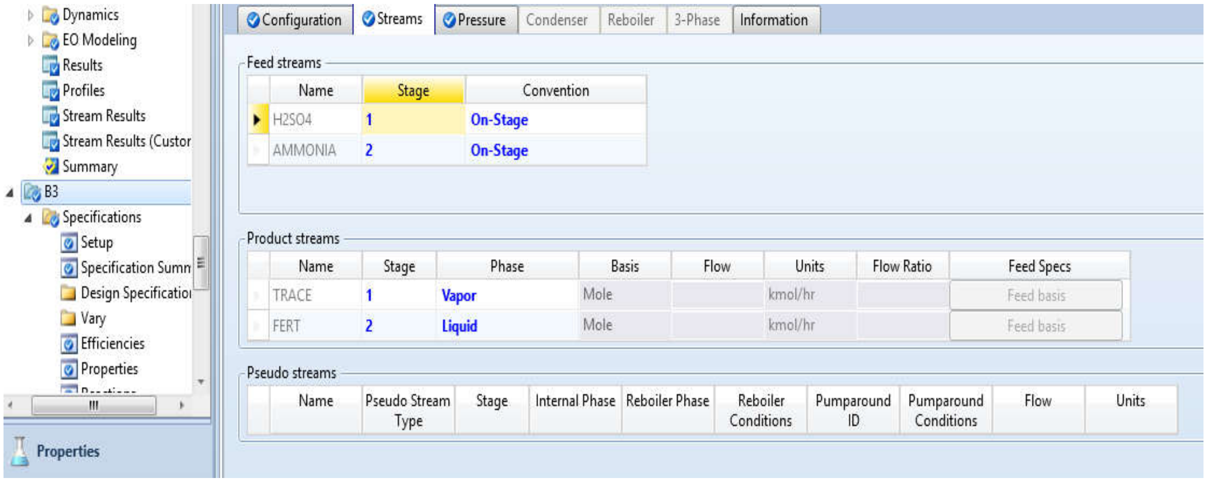
Setup options

Calculation type	Equilibrium
Number of stages	2
Condenser	None
Reboiler	None
Valid phases	Vapor-Liquid
Convergence	Standard

Operating specifications

Free water reflux ratio: 0

Figure 32: Absorber



Feed streams

Name	Stage	Convention
H2SO4	1	On-Stage
AMMONIA	2	On-Stage

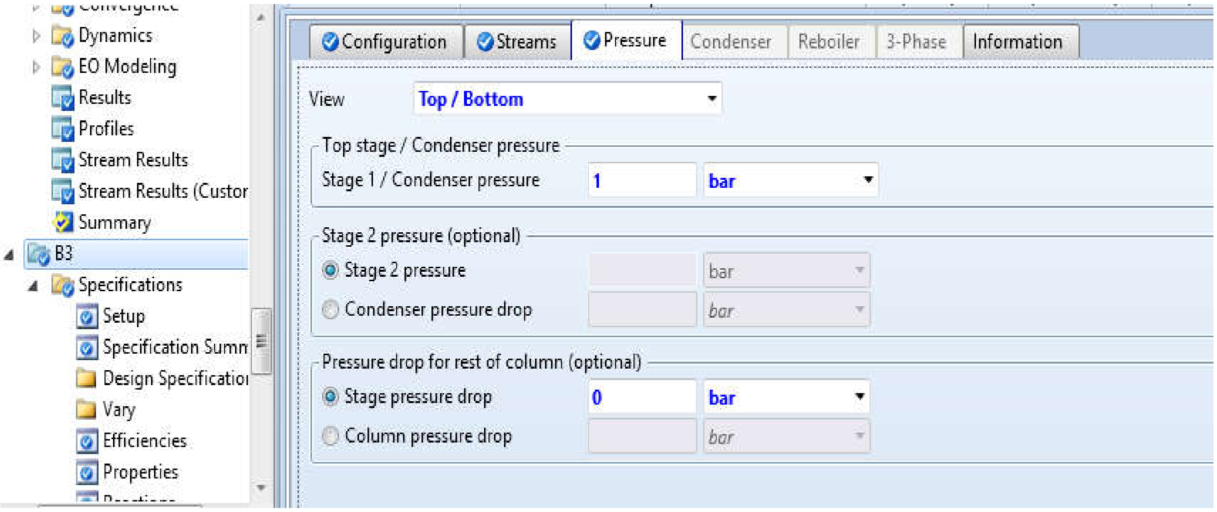
Product streams

Name	Stage	Phase	Basis	Flow	Units	Flow Ratio	Feed Specs
TRACE	1	Vapor	Mole		kmol/hr		Feed basis
FERT	2	Liquid	Mole		kmol/hr		Feed basis

Pseudo streams

Name	Pseudo Stream Type	Stage	Internal Phase	Reboiler Phase	Reboiler Conditions	Pumparound ID	Pumparound Conditions	Flow	Units

Figure 33: Absorber



View: Top / Bottom

Top stage / Condenser pressure

Stage 1 / Condenser pressure: 1 bar

Stage 2 pressure (optional)

● Stage 2 pressure — bar

○ Condenser pressure drop — bar

Pressure drop for rest of column (optional)

● Stage pressure drop: 0 bar

○ Column pressure drop — bar

Figure 34: Absorber

8 HAZOP Analysis

We conducted HAZOP analysis to see what are the hazards that certain deviations can cause and we have mentioned the recommendation for each one as well.

Guide Word	Parameter	Deviation	Possible Causes	Consequences	Recommendations
No	Flow	No Flow to stripper	Blockage of pipeline Valve Failure occurred Failure of water flow in previous units.	V/L is lower than required. High liquid holdup in the column	There should be a level controller in working condition alongside alarms in case excessive liquid holdup occurs.
More		More Flow to stripper	Valve stuck open	Flooding, which will lead to high Pressure Drop	Flow controller at inlet in accordance to gas velocity below flooding velocity
Less		Less Flow to stripper	Valve malfunctioning occurred	V/L ratio is lower than required	Flow controller in working condition
More	Pressure	High Pressure of air	Malfunction of Pressure regulating valve occurred	Breakage of the packing or bursting of column Effect on the reaction yield and kinetics occurring at the absorber	Pressure Relief valve and differential pressure control to monitor packing conditions
Less		Low	Malfunction of	It can affect the reaction yield and kinetics	

		Pressure of streams	Pressure regulating valve	occurring in the absorber	
More	Temp	High temperature of feed	Excess heat in absorber	Increased pressure in column	Pressure Relief Valve
Less		Low Temperature of streams	Low temperature in stripper.	Free ammonia is dissolved in water	Temperature control instrumentation.
More	Level	High level of liquid	Malfunction of level controller at bottom of stripper	Increased pressure drop and water carry over	Auto drainer in working condition

Table 46: HAZOP Analysis

8.1 Common Problems encountered in Columns

8.1.1 Flooding

It is the condition at which gas is at its maximum velocity so it offers very high resistance to the coming liquid from top of saturator and liquid fills the entire column and operation becomes difficult to carry out. High pressure can cause breakage of packings. We have to avoid this condition of flooding in the packed column.

To avoid this condition, we have a proper process control loop in which we measure the flow rate and the velocity of the gas entering from the bottom of column and we have given the set point to flow controller as 0.7 of flooding velocity. Which will compare current flow rate with the set point and to maintain flow rate as 0.7 of flooding velocity changes the percent opening of the flow control valve. The remaining gas will be bypassed from the saturator.

8.1.2 Weeping

This is the condition in which liquid trickles down from the holes in the trays of tray column. It occurs in situations in which gas pressure is used to hold the liquid on the tray and liquid does not come down through down comer but trickles down from the holes in the column. Weeping generally occurs in tray column and specifically in sieve tray columns. It is not the phenomenon occurring in the packed bed column.

8.1.3 Channeling

This is the phenomenon usually occurring in the packed column. Which occurs at low liquid rates. In this situation liquid follows some paths and flows down to the bottom of column leaving some packing surface as dry that is unequal wetting of the packings. This unequal wetting results in low efficiency of mass and heat transfer and reduced contact area of gas and liquid. To avoid this situation, we have liquid distributor to evenly distribute the liquid on the whole packing surface to increase the efficiency of the process.

9 Conclusion

By employing an ammonia removal and recovery process using a combination of an air stripper and absorber we have reduced the amount of Ammonia in our process stream from 800 ppm to 20 ppm. Ammonia in effluent air and water stream is 20 and 12 ppm, respectively which meets the Environmental Protection Agencies standard for Ammonia presence in the effluent streams. This amount of ammonia in air and water doesn't pose any threats to human health or environment. So, it is safe to discard these streams in both air and water.

We are producing a lean solution of Ammonium Sulphate, which is a useful product and if processed further it can be used as a fertilizer, as it has 21% Nitrogen. It finds its application in other processes as well, e.g., water treatment, tanning and food additive.

From the cost benefit analysis of our proposed technique we have seen that the payback period is of three years and seven months, as this time is less than five years so we can say that it is profitable to undertake this project.

10 References

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