

**TREATMENT OF BENFIELD PROCESS
CONDENSATE FOR REMOVAL OF
DISSOLVED CHEMICALS LIKE CO₂ AND
NH₃ FOR REUSE**



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TO OUR PARENTS

Without whom none of this would have been possible and for their loving support throughout our lives.

AND TEACHERS

For inspiring us and supporting us throughout the entirety of this project.

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Authors

ABSTRACT

CO₂ Removal from Syngas is an important step before ammonia production to enhance the life of iron catalysts in ammonia reactor. At FFBL, hot potassium carbonate (Benfield) solution is used to dissolve CO₂ from syngas. This solution is regenerated through steam and the overhead of carbonate regenerator is separated into CO₂ rich vapor stream and Benfield process condensate PC. About 84000 kg/h of condensate is produced, 1/4th of which is dumped into the Indus River.

According to environmental and economic aspects, it is beneficial to reuse this water as boiler feed instead of discharging it as an industrial effluent. Although free of major impurities like total dissolved solids TDS and total suspended solids TSS, Benfield PC contains two dissolved gases CO₂ and NH₃ which render it unfit for boilers.

Several biological, chemical, and physical processes were reviewed including nitrification-denitrification, ion-exchange resins, membrane separation, steam stripping, and flash distillation. Steam stripping was found to be a commercial and proven technology for water degassing even at very low concentrations.

In this project, a cost-effective stripping system with economized feed preheating has been optimally designed to bring Benfield PC within boiler feed specifications.

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LIST OF ABBREVIATIONS

ABMA	American Boilers Manufacturers Association
ASME	American Society of Mechanical Engineers
BFW	Boiler Feed Water
Dg	Surface Tension Number
LS	Low Pressure Steam
N	Number of Theoretical Stages
N_m	Minimum Number of Theoretical Stages
Nu	Nusselt number
PC	Process Condensate
PEC	Purchase Equipment Cost
PPC	Physical Plant Cost
PPE	Personal Protective Equipment
Pr	Prandtl Number
R	Reflux Ratio
R_m	Minimum Reflux Ratio
Re	Reynolds Number
Sc	Liquid Schmidt Number

INTRODUCTION

1.1 Background

Ammonia is the largest-volume synthetic chemical produced in the world. The annual ammonia production has multiplied manifold during the last century.

The current worldwide ammonia production capacity is 235.34 million metric tons and is expected to increase to 290 metric tons over the next decade. Currently ammonia is used to produce ammonia-based fertilizers and chemicals [1]. There is

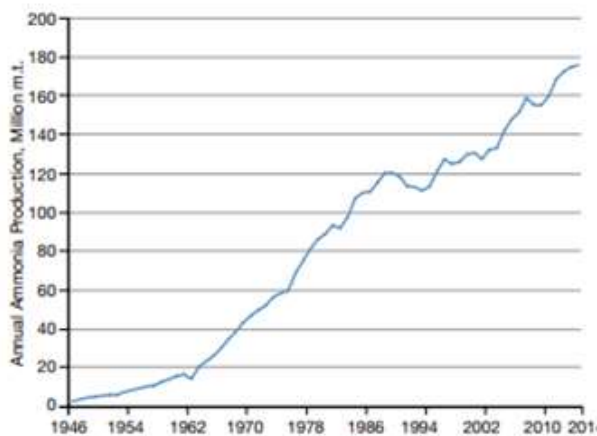


Figure 1 Ammonia Production Trend till 2014 [1]

an increasing trend in the use of ammonia as a sustainable fuel as well as a liquid energy carrier, justifying the increase in its production rate in the near future [2]. A major portion of Pakistan's economy is dependent upon agriculture. The agriculture sector, in turn, depends upon the continuous supply of fertilizer. Urea is one of the most used fertilizers in the country and it is produced by the reaction of ammonia and carbon dioxide. Therefore, the production of ammonia is a significant process, not only from the industrial perspective, but also from the perspective of Pakistan's economy. Continuous and abundant production of ammonia would ensure stability in the agriculture sector. Around 80% of the total ammonia produced universally is utilized for fertilizer production.

The present production technology of ammonia involves the Haber-Bosch process for the conversion of syngas containing H_2 and N_2 to NH_3 . The steps for ammonia production can be summarized as:

- Natural gas pretreatment
- Steam and Air Reforming
- Methanation
- High and Low Temperature Shift Conversion
- CO₂ Removal
- Ammonia Reactor

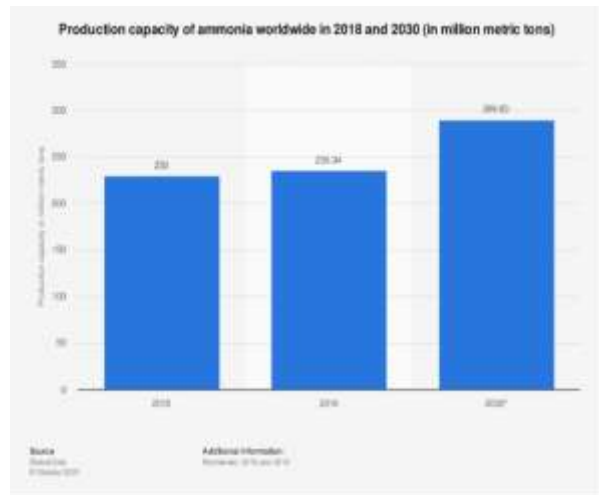
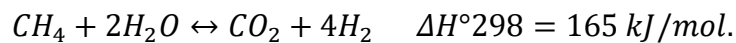
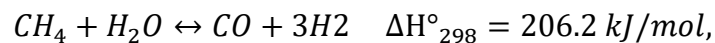


Figure 2 Ammonia Production till 2030^[2]

These steps have been carefully designed over the years and are being used globally, with slight

modifications. The first and a very important step in Ammonia production is the production of Hydrogen via 'Steam Reforming'. The reforming process begins with the pretreatment of natural gas to remove sulfur. Steam is prepared at desired reaction conditions, and it reacts with natural gas inside the primary reactor. The reforming reactions are endothermic in nature and occur as follows: ^[3]



Any remaining methane reacts with air in the secondary reformer. To achieve maximum conversion of methane, the temperature of secondary reformer is kept high. Nitrogen also enters the gas stream along with air and is sent to the ammonia reactor after undergoing pre-treatment.

After reforming, high and low-temperature shift conversion reactions convert the carbon monoxide to carbon dioxide. The catalysts used in both reactions are different and so are their conditions. The reaction equation for both high and low-temperature shift reactions is the same.^[3]



The carbon dioxide produced here is further separated by a suitable solvent in scrubbing and stripping columns. The separated carbon dioxide is sent to the urea unit as a reactant for urea formation reaction. The hydrogen left behind, after going

through several other steps to bring it to the required conditions for ammonia reaction, is sent to the ammonia reactor.

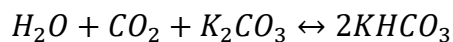
In ammonia reactor, the following reaction occurs in the presence of an iron-based catalyst to produce ammonia [4]



In a single pass, the conversion of the reactants to ammonia is low. To enhance conversion, the unconverted reactants are recycled back to the reactor. Due to recycling, the inerts such as argon and methane that are present along with the reactants start to accumulate inside the reactor and can decrease the required reaction conversion. To tackle this problem, a purge stream is provided. The purge gas composition is usually 60% hydrogen, 20-25% nitrogen, 10% methane, 5% argon and 4% ammonia. In the purge stream, the useful gases such as ammonia and hydrogen are also present and must be separated and sent back to reactor.

1.2 Problem Statement

FFBL uses Benfield solution for CO₂ removal from syngas. It contains 30% hot potassium carbonate solution with 1% V₂O₅ and 3 % DEA. Chemically absorption of CO₂ by Benfield solution can be represented by the overall reaction [5],



After CO₂ absorption this solution is regenerated using steam stripping. As a result, 150000 kg/hr of CO₂- loaded steam is produced in stripper overhead. It is condensed and separated into the CO₂ vapor which is sent to urea plant and the Benfield process condensate which is divided into four sections: reflux, wash water, boiler feed water, and waste. This process condensate PC is contaminated by dissolved gases high enough to have adverse effects on boilers. This is the reason that FFBL wastes majority of its Benfield process condensate (20530 kg/hr) instead of reusing it as BFW.

On comparing the sample analysis of Benfield PC provided by FFBL with American Boilers Manufacturers Association ABMA standard for Boiler Feed Water the only major impurities in the FFBL Benfield Condensate that requires removal are the

dissolved CO₂ and NH₃. The condensate is free of total dissolved solids TDS and total suspended solids TSS. Hence it is lucrative to use this PC as BFW after cost-effective degasification.

1.3 Objective of Project

FFBL uses the Benfield process for CO₂ removal from syngas. Around 66500 kg/h of steam condensate is produced from the overhead of the Carbonate Regenerator Column at the CO₂ removal section of FFBL's Ammonia-1 plant. This condensate is contaminated with dissolved CO₂ and NH₃ gases which make it unfit for reuse. Despite the impurities, 33500 kg/h of the condensate is refluxed back into the column, 6800 kg/h is used as wash water in the methanator, 5400 kg/h is used as BFW. The remaining 20500 kg/h of water has to be dumped into the Indus River. Our task is to design a cost-effective method for purification of this currently wasted process condensate to bring it within the boiler feed water specifications. The recovered gases CO₂ and NH₃ can also be reused to increase profitability of the process.

LITERATURE REVIEW

2.1 Why Reuse Condensate?

Process Condensate holds economic value in terms of boiler feed water. It is free of major impurities and can be prepared to be used in boilers at much smaller expense as compared to raw water. In addition, usually ammonia plants are located at remote locations where accessibility to water is difficult. In such areas it is favorable to enhance water economy by reusing the process condensate.

Steam is a utility that is used in almost every step of ammonia production. The process condensate of ammonia plant refers to the condensed steam that comes out of the steam reforming, shift conversion, CO₂ removal, and methanation sections after performing its desired function. [6]

Reusing process condensate provides us the following advantages:

- Decreases water charges
- Reduces Effluent charges
- Fuel costs are reduced
- More steam can be produced from the boiler
- Chemical treatment of raw make-up water is reduced.

2.2 Benfield Process Condensate

The process condensate that is produced by the CO₂ removal unit of ammonia plant using Benfield Solution for CO₂ removal is specifically referred to as Benfield Process Condensate.

2.2.1 Sample Analysis

Prior to the literature survey to determine a suitable treatment process of FFBL's Benfield condensate, it was important to analyze a sample of the condensate to identify the impurities present and determine their concentrations. A comparison of

the sample analysis results provided by FFBL with the required boiler feed specifications is represented in the table below,

Impurity	Sample Analysis	Boiler Feed Requirement [7,8]
TDS	<10 ppm	120 ppm
Total Hardness	<1ppm	<1ppm
Chloride	0.5 ppm	2 - 6 ppm
Potassium Carbonate	1 ppm	<1 ppm
Ammonia Content	1300 ppm	<10 ppm ^[2]
Carbon Dioxide Content	5200 ppm	0 ppm

Table 1 Comparison of Benfield Sample Analysis with BFW Standard

As already expected, the Benfield process condensate is low in TDS and hardness. The chloride and potassium carbonate content also lie within the specification limits. Carbon dioxide and ammonia are the main impurities in the form of dissolved gases that require removal.

2.2.2 Condensate Chemistry (The NH₃ – CO₂ – H₂O system)

The NH₃ – CO₂ – H₂O system is a multi-component aqueous solution of volatile weak electrolytes ^[9,10].

In an electrolyte solution, the solute is present in two forms: molecular and ionic due to the chemical dissociation of electrolytic solute in the solvent. The distribution of a volatile electrolytic solute in the vapor and liquid phases is determined by the mole fraction of the molecular form of solute in the liquid phase. The ionic form does not contribute to the partial pressure since it cannot vaporize.

The figure to the right shows the distribution of a single electrolyte in vapor and liquid phases. Two types of equilibria are present in the system.

- Chemical Equilibrium – The dissociation of molecular solute into ionic form is determined by chemical equilibrium constants.
- Vapor Liquid Equilibrium – The distribution of molecular solute in vapor and liquid phases is governed by Henry's law (for dilute solutions as in our case).

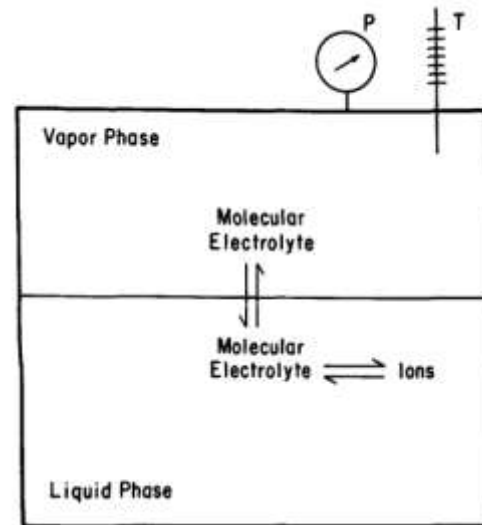
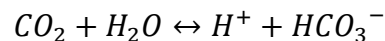
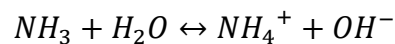


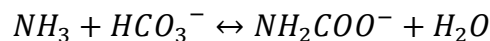
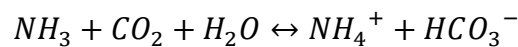
Figure 3 Vapor Liquid Equilibrium in an Electrolyte Solution^[9]

2.2.2.1 Chemical Equilibrium

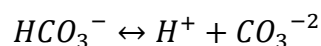
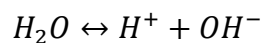
NH₃ and CO₂ are weak electrolytes and undergo following dissociation reactions in water,



Since NH₃ is a weak base and CO₂ is a weak acid, they react with each other as well.



The ionization of water and second dissociation of HCO₃⁻ also need to be considered in this system:



All these reactions result in a decrease in the molecular form of the electrolyte in the solution which determines the vapor pressure of the electrolyte solutes. The equilibrium constants for the above reactions can be written as,

$$K_1 = \frac{a_{NH_4^+} a_{OH^-}}{a_{NH_3} a_{H_2O}}$$

$$K_2 = \frac{a_{H^+} a_{HCO_3^-}}{a_{CO_2} a_{H_2O}}$$

$$K_3 = \frac{a_{NH_4^+} a_{HCO_3^-}}{a_{NH_3} a_{CO_2} a_{H_2O}}$$

$$K_4 = \frac{a_{NH_4^+} a_{CO_3^{2-}}}{a_{NH_3} a_{HCO_3^-}}$$

$$K_5 = \frac{a_{NH_2COO^-} a_{H_2O}}{a_{NH_3} a_{HCO_3^-}}$$

$$K_6 = \frac{a_{H^+} a_{OH^-}}{a_{H_2O}}$$

$$K_7 = \frac{a_{H^+} a_{CO_3^{2-}}}{a_{HCO_3^-}}$$

2.2.2.2 Vapor-Liquid Equilibrium

The distribution of molecular solute in vapor and liquid phases is given by the Henry's law as follows:

$$y_a \varphi_a P = m_a \gamma_a^\circ \underline{H}^{(P)}$$

Where,

y_a = vapor phase mole fraction of molecular solute

φ_a = vapour phase fugacity coefficient of molecular solute

P = Total pressure

m_a = molality of molecular solute

γ_a° = molal activity coefficient of molecular solute

$\underline{H}^{(P)}$ = Henry's constant for molecular solute at pressure P

2.3 Separation Processes

A literature review was carried out to find out different methods for the separation of dissolved gases from water at low concentration. Different physical, chemical, and biological processes were studied and compared in terms of efficiency, durability, and cost effectiveness.

2.3.1 Hollow Fiber Membrane Contactor

A hollow fiber membrane contactor is a device that brings about direct contact of two phases for mass transfer without intermixing of the two fluids [11]. The key concept is to use a hydrophobic membrane so that the fluids do not enter the pore. The solute particles travel across the membrane due to a concentration gradient across the pore.

2.3.1.1 Application of PTFE Membrane for Ammonia Removal in a Membrane Contactor [12]

In this article the feasibility of a membrane contactor for ammonia removal was studied. The author used a tubular membrane configuration in which water containing ammonia in the tubular side was contacted with a 10% w/w sulphuric acid extractant solution on the shell side. The effect of influent ammonia concentration, flow rates, suspended solids (SS), temperature gradient, and pH of water on the mass transfer was studied. The effect of the above-mentioned parameters is elaborated through the table below:

	NH ₃ in	SS	ΔT (°C)	Q _{feed} (mL/min)	Q _{strip} (mL/min)	Mass transfer coefficient. (× 10 ⁻³ m/h)
Influent	250	0	0	10	8	8.9
NH ₃ -N	1,000	0	0	10	8	7.0
SS	1,000	1,000	0	10	8	5.8
	1,000	3,000	0	10	8	5.4
Temperature	1,000	0	13	10	8	7.3
Flow condition	1,000	0	0	20	8	9.7
	1,000	0	0	10	16	11.0

Figure 4 Parameters Affecting Separation in Tubular Membrane Contactor[12]

The maximum mass transfer obtained for a 1000mg/L ammonia sample which is very close to our requirement of 1300 mg/L was more than 80% leaving less than 200 mg/L in the treated water. The suspended solids do not have much inhibiting effect on the membrane filtration process. This is because in a membrane contactor the driving force of separation is not pressure gradient but evaporation across the

membrane pores. Hence the suspended solids are not forced on the membrane walls due to which they do not cause membrane fouling. The author suggested that increasing contact time between water and extractant is the most effective way to enhance mass transfer. It was suggested that using a hollow fibre configuration can provide a large surface area and serve the purpose.

2.3.1.2 Simulation of Ammonia Removal from Industrial Wastewater Streams by HFMC

In this case study^[13] hollow fiber membrane contactors were used to provide an efficient way to remove dissolved ammonia from water by dispersion free contact with a sulphuric acid extractant liquid. The ammonia is volatilized at aqueous surface diffuses through the pores and instantaneously reacts with the extractant solution. In this study, aqueous ammonia feeds having 50 – 800 ppm ammonia were stripped through HFMC of pore size 0.03 microns. The equilibrium concentration of aqueous ammonia was examined. Passing the aqueous ammonia through lumen side provided a higher gas-liquid mass transfer interface at the shell side of the pore.

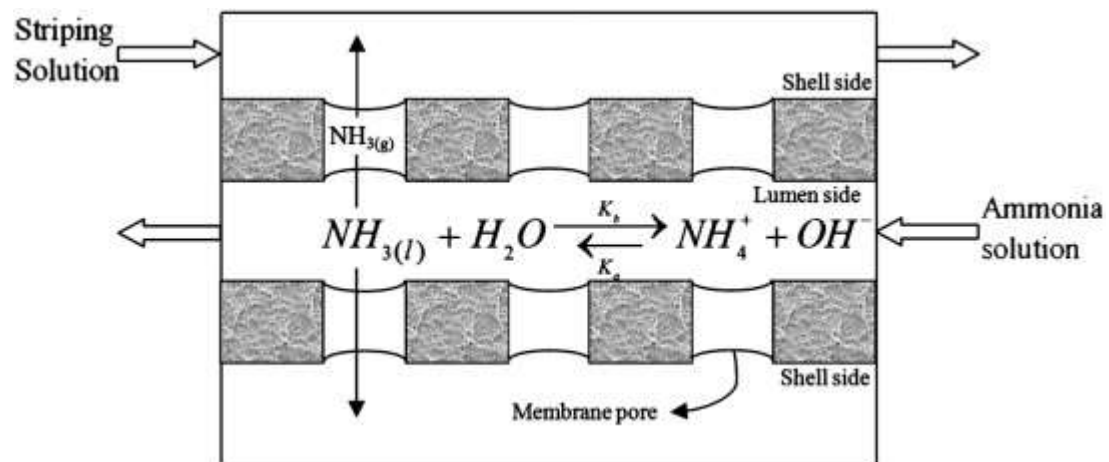


Figure 5 Ammonia Removal in HFMC^[13]

The stream flows through the tube and the gas molecules desorb from the feed to the extractant through the membrane pores via free molecular diffusion or Knudsen diffusion. Molecules are adsorbed and absorbed on the pore walls continuously. The feed containing ammonia in range 200-1500 ppm can be treated.

2.3.1.3 Drawbacks of Membrane Contactors

Membrane Contactors represent a promising solution to the degassing of water. However, this is a new phenomenon which needs to be further explored before being applied to industrial scale. Some of the limitations of membrane contactors include,

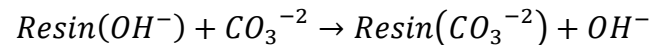
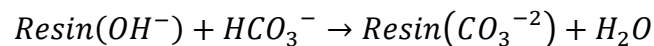
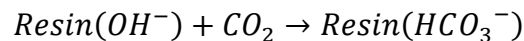
- Partial wetting of membrane pores reduces mass transfer^[14]
- High pressure drop is encountered for high flowrates
- Presence of impurities can cause membrane fouling
- Technology has been applied on lab or pilot scale, but industrial application is still under development

2.3.2 Ion Exchange Resin

Ion exchange resins are used for many water treatment applications. They consist of small plastic beads of half a millimeter size on which fixed ions have been permanently attached. To maintain electrical neutrality, counter ions are attached to the fixed ions which are mobile and can be replaced. Ion exchange resins can be anionic and cationic depending on the charge on the counter ion. Anionic resins remove anions and cationic resins remove cations from water.

2.3.2.1 Thermally Regenerable Resins

Since CO₂ and NH₃ are electrolytic in nature, their ionic forms can be removed from water using anion exchange resins for ammonium ions and cation exchange resins for carbonate and bicarbonate ions. Thermally regenerable resins are a new technology which can be used for water demineralization. Weak base ion exchange resins can remove carbon dioxide from water by replacing the hydroxide counter ions with carbonic acid ions ^[15]. The reactions will be as under,



Similarly, cation exchange resins with affinity towards ammonium ions such as clinoptilolite can be used to remove ammonium ions from water.

2.3.2.2 Drawbacks of ion Exchange Resins

Ion exchange resins have emerged as an effective water treatment technology over the years however there are certain limitations which still exist^[16].

- Strong base anion exchange resins are unstable at temperatures greater than 60°C.
- The available options for chelating anion exchange resins are much less as compared to chelating cation exchange resins
- Microbes can grow and reproduce in ion exchange resin beds
- As ion exchangers age fragments start sloughing off which reduces ion exchange capacity

In addition, this method requires regeneration of the resin after a certain time interval which makes it less suitable for continuous operation at high industrial flow rates.

2.3.3 Single-Stage Flash Distillation

It is a single stage process in which the feed stream is separated into liquid and vapor products at equilibrium. The composition of the two product streams depends on the vapor liquid equilibrium at the flash temperature and pressure which in turn depends on the volatility of the dissolved gases.

The process is carried out by partial evaporation of a saturated liquid by passing it through a throttling valve. The outlet of the throttling valve lies inside a vessel called a *knock-out drum*.

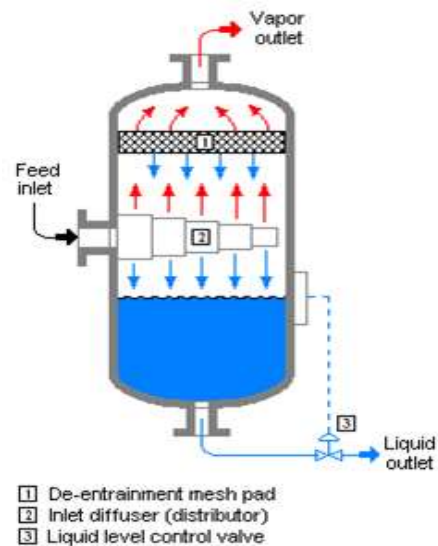
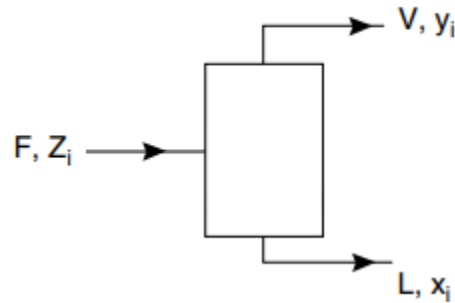


Figure 6 Flash Evaporation

2.3.3.1 Flash Equilibrium

The calculations of vapor and liquid compositions can be performed as follows [17].

Overall material balance on component i,



$$FZ_i = Lx_i + Vy_i$$

Let K be the vapor liquid equilibrium constant,

$$FZ_i = Lx_i + VKx_i \dots \dots \dots (i)$$

Rearranging eq (i) we get,

$$FZ_i = Lx_i \left(1 + \frac{V}{L} K_i \right)$$

$$Lx_i = \frac{FZ_i}{\left(1 + \frac{V}{L} K_i \right)}$$

Summing for all components,

$$\sum_{i=1}^n Lx_i = \sum_{i=1}^n \frac{FZ_i}{\left(1 + \frac{V}{L} K_i \right)}$$

$$L \sum_{i=1}^n x_i = \sum_{i=1}^n \frac{FZ_i}{\left(1 + \frac{V}{L} K_i \right)}$$

$$L = \sum_{i=1}^n \frac{FZ_i}{\left(1 + \frac{V}{L} K_i \right)}$$

Similarly,

$$V = \sum_{i=1}^n \frac{Fz_i}{\left(1 + \frac{L}{VK_i}\right)}$$

After calculating the liquid flow rate L and vapor flow rate V, vapor, and liquid compositions x_i and y_i can be calculated.

2.3.3.2 Feasibility Study on Aspen Plus

The feasibility of flash distillation for CO₂ and NH₃ removal from Benfield condensate was studied using Flash2 block in Aspen Plus at 1 bar pressure. The results obtained are tabulated below. Here the compositions and stream flows are given in mass basis.

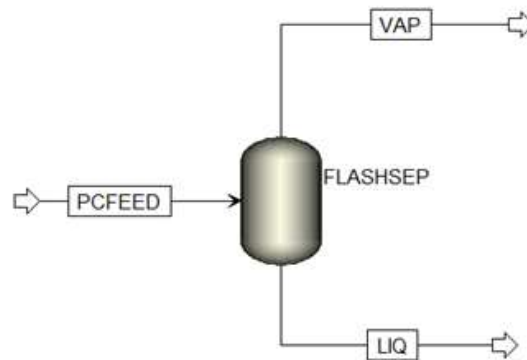


Figure 7 Flash2 Block in Aspen Plus

Pressure = 1bar						
Temperature	x_{NH3}	x_{CO2}	y_{NH3}	y_{CO2}	L	V
98	0.000732	0.0001	0.008432	0.06901	19016	1514
98.5	0.000501	5.817×10^{-5}	0.005960	0.03515	17522	3008
98.7	0.000365	3.637×10^{-5}	0.004396	0.00223	15768.5	4761.53
98.9	0.000194	1.630×10^{-5}	0.002366	0.01020	10079.2	10450.8
99	0	0	0.0013	0.0052	0	20530

Table 2 Flash Distillation at Different Temperatures

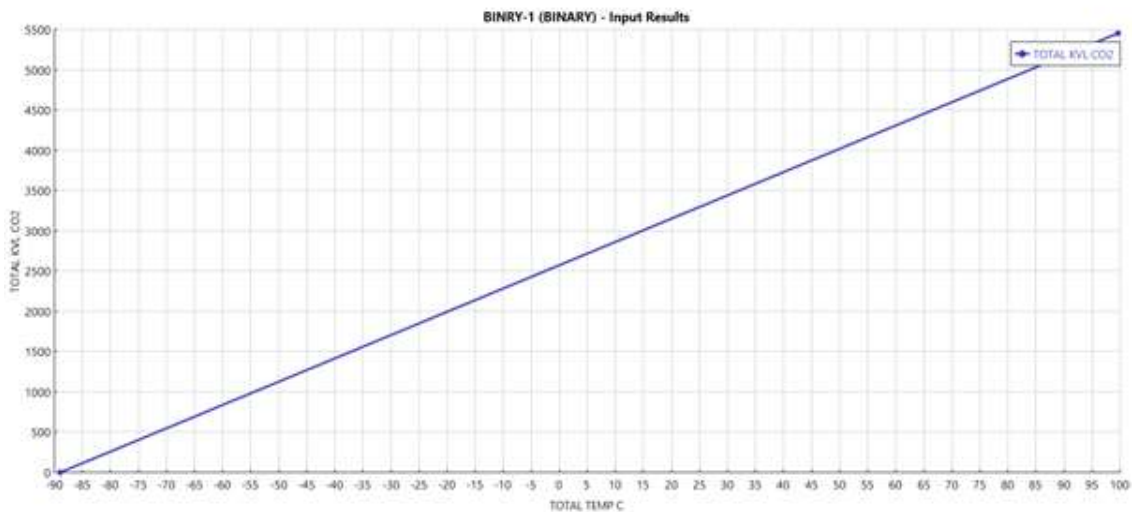
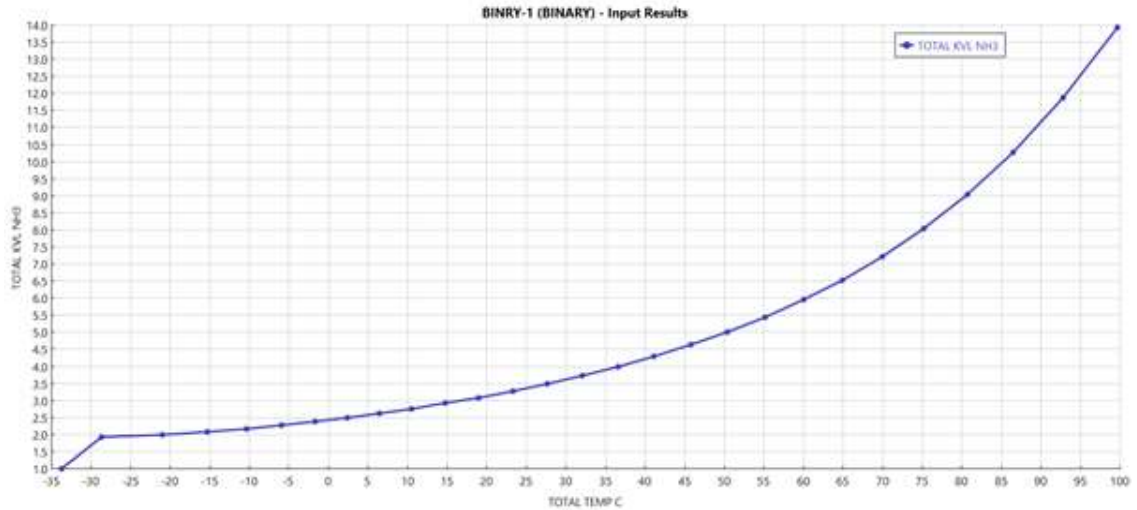


Figure 8 K-values of Ammonia and Carbon Dioxide

2.3.3.3 Conclusion from Feasibility Study

It can be concluded from the flash distillation results in the table and the vapor-liquid equilibrium constant graphs above that CO₂ is much more volatile than NH₃. At 40°C, the K-value of NH₃ is approximately 4.7 whereas that of CO₂ is 3700. Hence, the amount of CO₂ in Benfield condensate can be easily reduced to the desired standard for boiler feed water. However, the amount of NH₃ decreased to only 190 ppm at 98.9°C after which the liquid is completely vaporized. This is much greater than our

required ammonia concentration of 5ppm. In addition, the amount of water that is entrained into the vapor state is also very high (10450 kg/hr) at 98.9.

In conclusion, using single-stage flash distillation or knock-out drum is not sufficient for ammonia removal and involves a significant amount of water wasted in the vapor stream.

2.3.4 Stripping

Stripping is defined as a process involving physical separation in which using a vapor stream one or more constituents are removed from a liquid stream. Stripping works on the basis of the mass transfer.

A number of PC stripping system configurations that have been applied in ammonia industry for years have been studied and are discussed below [18].

2.3.4.1 Once Through PC Stripper

The process condensate is counter currently contacted with steam in a trayed column. Purge containing the removed gases is released into the atmosphere. Stripped process condensate in bottom stream is cooled and utilized as needed.

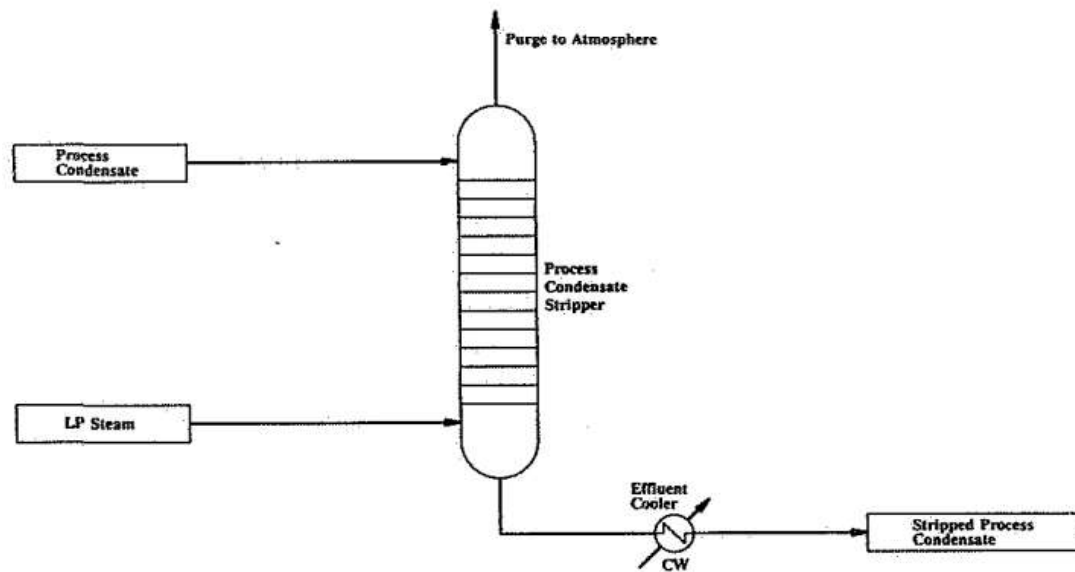


Figure 9 Once Through PC Stripper [18]

2.3.4.2 PC Stripper with Reflux System

To decrease the amount of moisture in the purge stream, an additional overhead condenser is added. A minimum but very concentrated purge now leaves the stripper and enters the atmosphere.

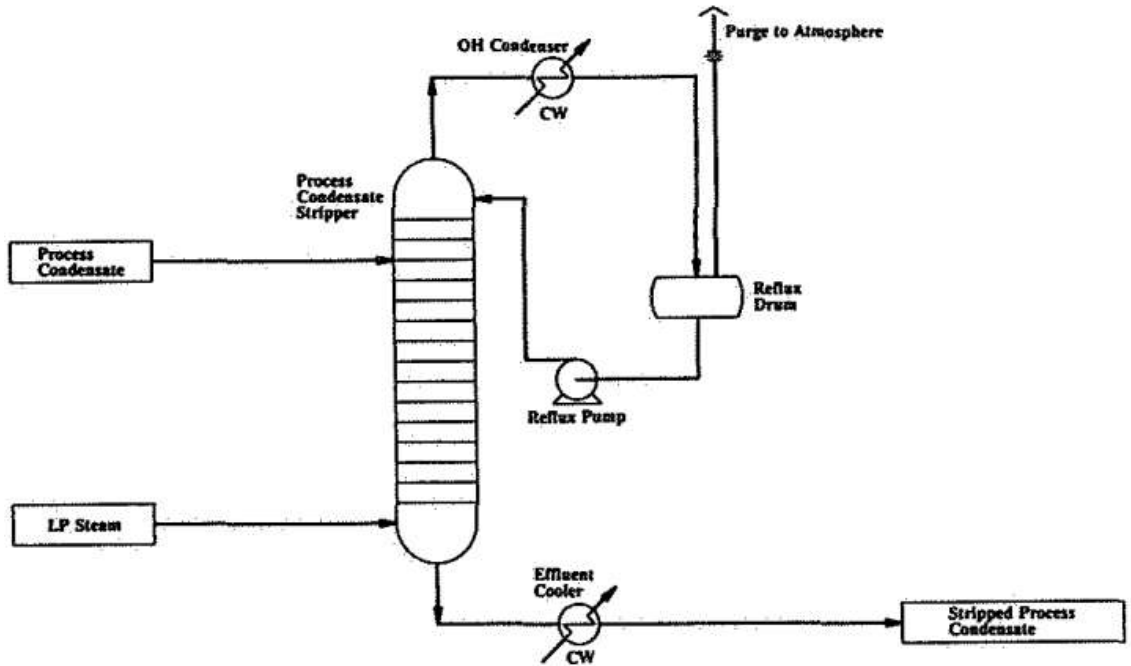


Figure 10 PC Stripper with Moisture reflux ^[18]

2.3.5 PC Stripper with Economizer Feed Preheater

In this PC system energy efficiency is enhanced and stripping is facilitated by preheating our process condensate feed by heat exchange with stripped process condensate in stripper bottoms. The overhead is also integrated back into the process instead of releasing it into the atmosphere.

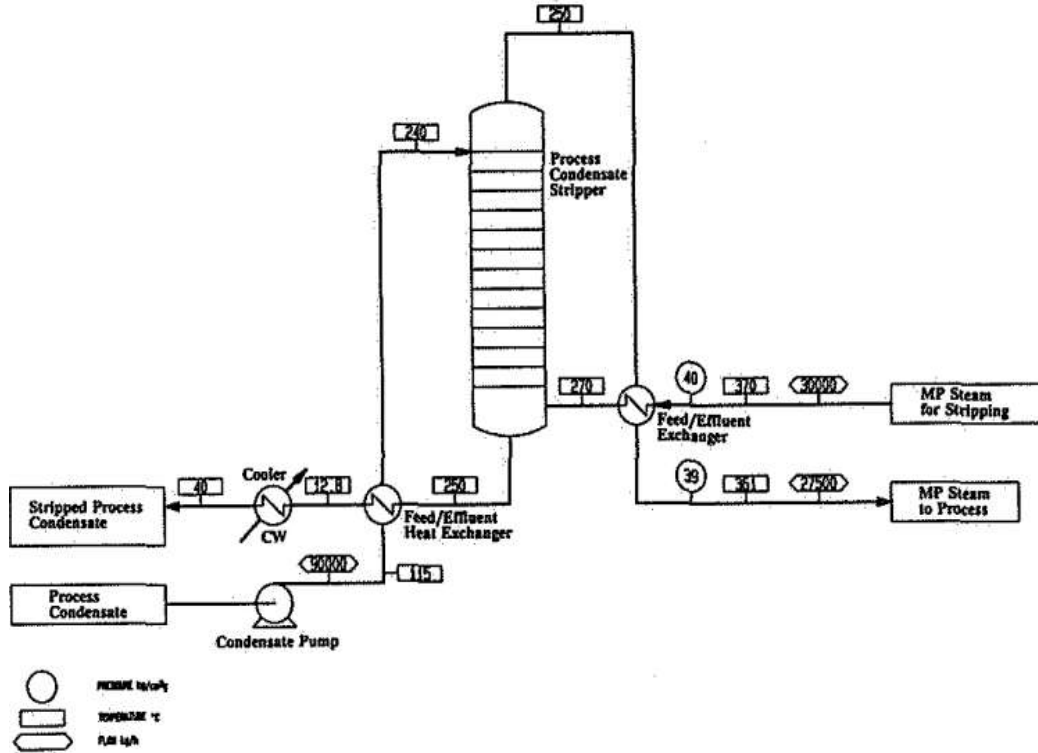


Figure 11 PC Stripper with Economizer Feed Preheater^[18]

2.3.5.1 Feasibility of PC Stripping System for Benfield Condensate

**Contaminants Measured in Process Condensate at Different Locations
in an Ammonia Plant Operating at S/C Ratio -3.3**

Location	Conductivity µs/cm	pH	NH ₃ ⁽¹⁾ ppm w	ΣCO ₃ ⁻⁽¹⁾ %	CH ₃ OH ppm w	HCOOH ppm w	CH ₂ COOH ppm w	Methylamines ppm w	Dimethylamines ppm w	Trimethylamines ppm w
Outlet LTS	5700	8.2	895	0.26	20	30	29	<0.5	<0.5	<0.5
Outlet HTS (17 t/h)	2420	8.8	455	0.10	300	20	<3	27	13	5
Inlet CO ₂ Absorber (35 t/h)	9500	7.6	1490	0.47	840	30	<3	21	9	23
OH CO ₂ Regenerator (37 t/h)	3500	6.8	500	0.21	1100	15	<3	4	<0.5	<2.5
Outlet CO ₂ Flash Vessel Bottom (88 t/h)	90	9.1	765	0.14	825	10	<3	<0.5	<0.5	<0.5
Outlet PC Stripper Bottom (110 t/h)	3400	9.4	10	0.025	45	5	<3	10	5	<0.5
PC Stripper Reflux (30 t/h)	24000	9.9	16700	1.21	17200	10	8	120	30	30
Vapour from Reflux Drum	-	-	-	-	46000	-	-	250	190	67

(1) Total carbon and nitrogen calculated as CO₃²⁻ and NH₃ respectively.

Figure 12 Results of PC Stripping at an Ammonia Plant [18]

- The table above shows the contaminant compositions in process condensate at various locations of an ammonia plant that uses stripping for treatment of its process condensate. The PC stripper bottoms has 10 ppm NH₃ and negligible CO₂ (mentioned in the form of CO₃²⁻ ions).
- The flow rate of process condensate entering the stripper is also 110t/h. the flow rate of waste Benfield process condensate is 22.53 t/h. Hence this process can be easily applied for our flow rate with the opportunity for scale-up to treat more PC in future if needed.
- The steam that is used for stripping process condensate can be easily taken from the process and the purge stream can be integrated back. This makes steam stripping an easy to implement process for PC stripping with no waste by-product streams produced.

2.4 Equipment

2.4.1 Heat Exchangers

They are used to transfer heat between fluids through direct or indirect contact and which may flow parallel or counter to each other. Depending upon the required duty

and application, there are several types of heat exchangers. A proper choice must be made of the heat exchanger during designing so that required heat transfer is achieved. Some prominent types of heat exchangers are discussed below.

2.4.2 Shell and Tube Heat Exchangers

They are the most popular heat exchangers in industries. The reason for this is that they can be operated on a wide range of temperatures and pressures. There are several tubes mounted in a cylindrical shell in a shell and tube exchanger. The typical unit found in a petrochemical plant is illustrated in the figure below. The heat can be exchanged for two fluids, and one fluid flows across the pipes while the second fluid passes through them. The fluids can be in single phase or two phases. The operation can be performed either by co-current flow or counter current flow.

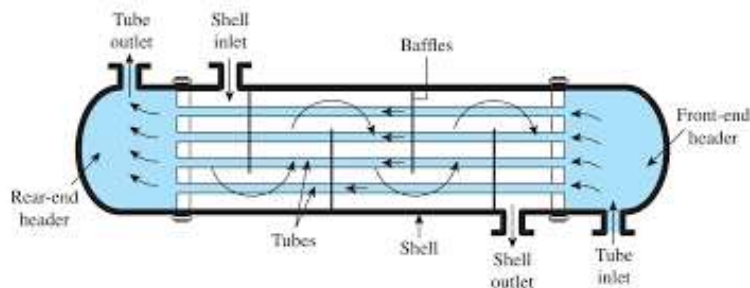


Figure 13 Shell and Tube Heat Exchanger Layout

- **Front Header:** This is the point from which the fluid is entered into the tube side of the heat exchanger. It is also called a stationary header.
- **Rear Header:** This is the point from where the fluid either leaves the heat exchanger or can be returned to the tube side for another pass.
- **Tube Bundle:** The tube bundle is the set of tubes, baffles and tube sheets and rods to keep the tubes fixed in one place.
- **Shell:** It is the outer body of the exchanger enclosing all the internal components. The fluids used in the exchanger can be both liquid and gas. The tubes inside the shell can be arranged in different geometries to broaden the use under various circumstances. The following three combinations are mostly used.

- **Fixed Tube Sheet Exchangers:** in this type of the tubes bundle is welded to the shell. These are not very recommended to use because of difficulty in the cleaning of tubes.
- **U-tube Exchangers:** Any front header types and the rear header is normally M-Type can be used in the U-tube exchange. The U tubes allow unlimited thermal expansion and can be removed for cleaning and small bundles can be achieved with shell clearance.
- **Floating Head Exchanger:** The tube sheet on the back of the header is not sold to the shell in this type of exchanger but can be moved or floated. The pipe board at the front end is larger in diameter than the shell and is sealed similarly as the one used in the design of the fixed tube board. The pipe sheet at the end of the shell's back header is slightly smaller in diameter than the shell, so that the shell can be pulled. Using a floating head allows for thermal expansion and can remove the tube bundle for purification.

2.4.2.1 Factors Influencing the Performance of Shell and Tube Heat Exchanger

- Tube length
- Shell diameter
- Fouling
- Baffles type and spacing
- Log mean temperature difference
- Arrangement of tubes

2.4.2.2 Advantages

- Less pressure drops on both shell and tube side
- Tube leaks are easy to find
- Lower cost comparing to plate-type cooler

2.4.2.3 Disadvantages

- Demands more space as compared to plate and frame heat exchanger
- Maintaining and cleaning is tough
- As compared to a plate-type cooler, the heat exchange is less efficient
- Expansion of tube cooler capacity is not possible

2.4.3 Double Pipe Heat Exchanger

In a double pipe heat exchanger, a smaller pipe is held concentrically inside a comparatively larger pipe. Hence one fluid flows through the inner tube while the second one flows through the annulus pipe. A threaded connection is placed properly outside the exchanger section to support the inner pipe within the outer pipe. The tees have nozzles or screwed connection attached to them to allow the entrance and exit of fluid present in annulus. The two lengths of annular pipe are connected through a return head which is sometimes exposed and does not provide surface

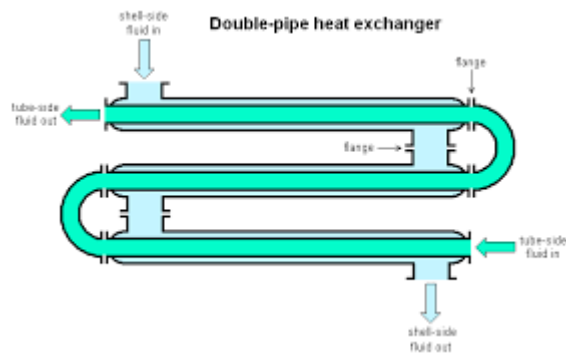


Figure 14 Double Pipe Heat Exchanger Layout

with effective heat transfer. Double pipe heat exchangers are usually assembled in 12, 15, or 20ft effective lengths.

2.4.3.1 Advantages

- The double pipe heat exchanger is exceptionally useful because it provides inexpensive heat transfer surface and can be assembled easily in a pipe fitting shop from standard parts.
- Suited to high pressure applications
- Provides flexibility since units can be added or removed
- Double pipe heat exchangers are compact

2.4.3.2 Disadvantages

- Limited to lower heat duties requiring surface area 100-200 ft
- The principal disadvantage of using double pipe heat exchanger is its small heat transfer surface in a single hairpin.

- When hairpins are employed more than 20 ft, the inner tube tends to sag and touch the outer pipe.

2.4.4 Plate Contactors ^[17]

In stripping columns, the interfacial area of contact between the vapor and liquid stream is enhanced by using packings or trays. The columns that use trays or plates for facilitating the contact between the two streams are called plate contactors. The commonly used plates allow the cross flow contact of liquid and vapor. They can be divided into three main types.

2.4.4.1 Sieve Plate Contactor

The sieve type trays consist of perforated trays which retain liquid on the tray by vapor flow through the holes. The holes vary in diameter depending on the liquid and vapor flows. There is no liquid seal due to which the liquid might weep at lower flow rates.

2.4.4.2 Bubble-Cap Plate Contactor

In bubble cap trays a certain level of liquid is maintained on the plate through risers. The risers allow the vapor to flow across the plates and are covered with serrated edge caps. The problem of liquid weeping is resolved in these plates, but the vapor pressure drop is high.

2.4.4.3 Valve Plate Contactor

Valve plates are essentially sieve plates with larger diameter holes covered with flaps. The hole opening varies according to the vapor flow rate. In this way valve plates are suitable for low vapor flow rates as well.

2.4.4.4 Selection of Plate Type ^[17]

Parameter	Sieve Plates	Valve Plates	Bubble-Cap Plates
Cost	Least Expensive	Moderate	Most expensive
	Sieve: Valve: Bubble-Cap = 1.0 : 1.5 : 3.0		
Capacity	Sieve > Valve > Bubble-Cap		
Operating Range	Not effective at very low vapor flow rates	Can operate at low vapor flow rates	Can operate at low vapor flow rates
Efficiency	Essentially same when operating over design flow range		
Pressure Drop	Sieve < Valve < Bubble-Cap		

Table 3 Comparison for Selection of Plate Type

Since our vapor and liquid flow rates lie in moderate range, we will use sieve plate contactor for cost effectiveness and low pressure drop.

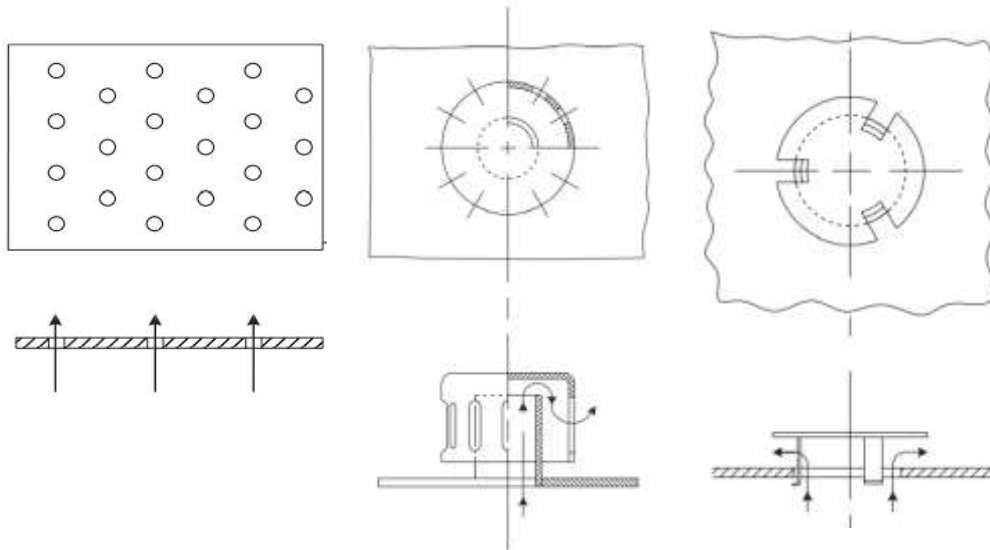


Figure 15 Sieve, Bubble-cap, and Valve Plates ^[17]

PROCESS DESCRIPTION

3.1 General Layout of Proposed PC Stripping System

After choosing steam stripping as our technique for degasification of Benfield process condensate, we needed to decide the equipment and design the layout of the system.

- As studied in the literature review, increasing temperature of condensate decreases the ammonia and carbon dioxide solubility. We decided to use an economizer to preheat our process condensate feed. The bottoms of the stripper is used as the heating fluid in this preheater.
- The preheated feed enters the top stage of stripper where it counter currently contacts Low pressure saturated steam. The stripped process condensate exits in the bottoms stream and is taken to the economizer for heat recovery.
- Vapor overhead from the PC stripper contains stripped ammonia and carbon dioxide gases as well as some entrained moisture. Due to the presence of these harmful gases, it cannot be directly released to the atmosphere. We integrated this vapor overhead stream with the CO₂ stream stripped from the syngas.
- After mixing, the vapor stream enters a knock-out drum for the removal of excess moisture. The vapor stream leaving the knock-out drum is taken to the urea plant. The condensed moisture is refluxed back to the PC stripping column through a reflux pump.

3.2 Components of Benfield PC Stripping System

The Benfield process condensate stripping system consists of the following components:

- An Economizer Feed Preheater
- A Process Condensate Steam Stripper

- A Knock-Out Drum
- A Mixing Tee
- A Reflux Pump
- A Bottoms Pump

3.2.1 Economizer

The economizer is a shell and tube type heat exchanger which provides heat transfer from the stripped condensate in stripper bottoms to the process condensate feed. Since the mass flow rates of PC feed and stripped PC are similar, the temperature of the PC feed is efficiently increased from 40°C to 65°C. The Feed temperature of 65°C was set after optimization discussed in the chapter of simulation

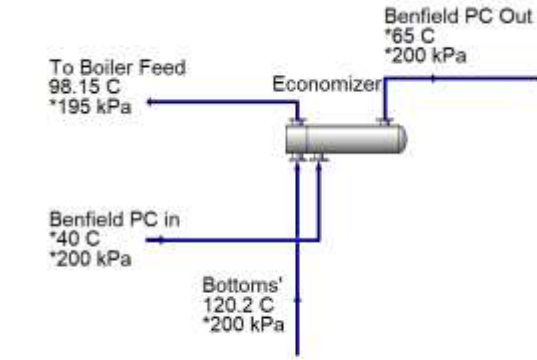


Figure 16 Economizer

and optimization. The dependent variables considered in optimization are level of moisture entrainment in stripper overhead and ammonia content in stripped PC.

3.2.2 PC Stripper

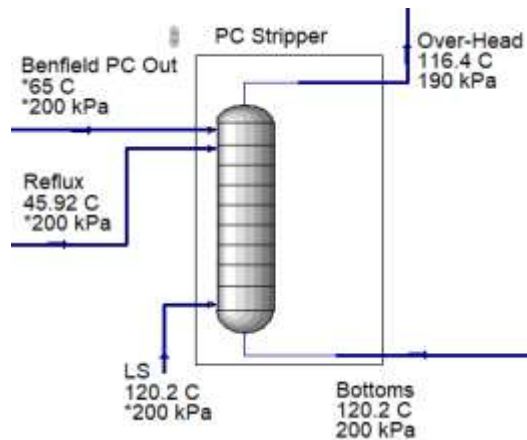
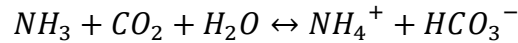


Figure 17 PC Stripper

A sieve tray PC stripper consisting of 13 real stages is used to strip carbon dioxide and ammonia from Benfield process condensate using low pressure steam. The temperature and pressure of stripper is kept at 2 bar and 120°C respectively. A relatively low pressure enhances ammonia removal as it allows for more vapor formation. The

temperature of steam which also determines the temperature of stripper is kept 120°C.

This temperature is required to decompose ammonium bicarbonate formed in the stripper because of reaction between weakly basic ammonia and weakly acidic carbon dioxide.



At 2 bar or 0.2 MPa, the decomposition temperature of ammonium bicarbonate is 120°C.

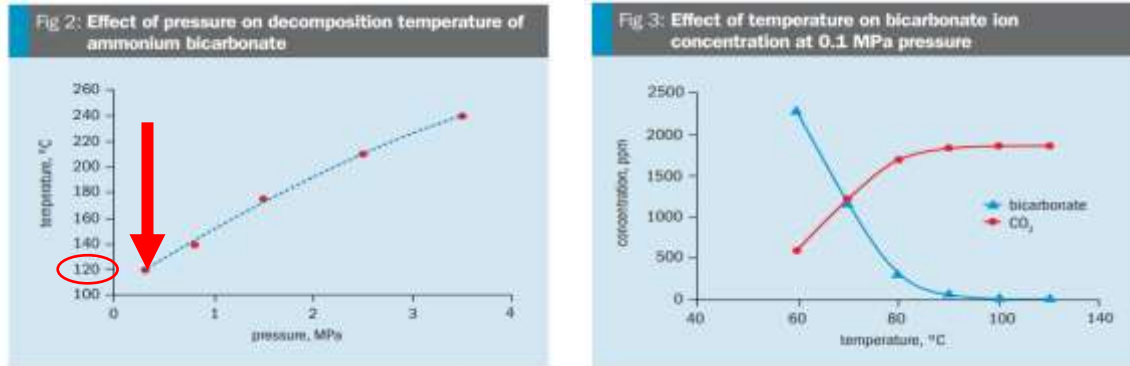


Figure 18 Decomposition Temperature of Ammonium Bicarbonate

3.2.3 Mixing

The vapor overhead of the PC stripper is usually partially condensed with liquid part refluxed and the purge stream released into the atmosphere or integrated back into the process.

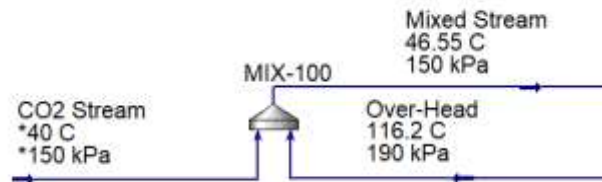


Figure 19 Mixing Tee

During simulation, we ran two case study of steam flow rate vs. moisture entrainment and steam flow rate vs. ammonia mass fraction in stripper bottoms. This investigation helped us to optimize our steam flow rate to achieve a minimum vapor overhead flow of 930 kg/h. After minimizing the vapor overhead flowrate, it is combined with the CO₂ stream that is separated from syngas in CO₂ removal unit and is sent to a knock-out drum.

3.2.4 The Knock-Out Drum

The knock-out drum is a two-phase separator which removes the excess moisture from CO₂ to Urea stream and refluxes it back to the PC stripper column. Since the flow rate of stripper vapor overhead is much less as compared to the CO₂ stream it is mixed with, there is a very small increase in the moisture content in the original CO₂ stream.

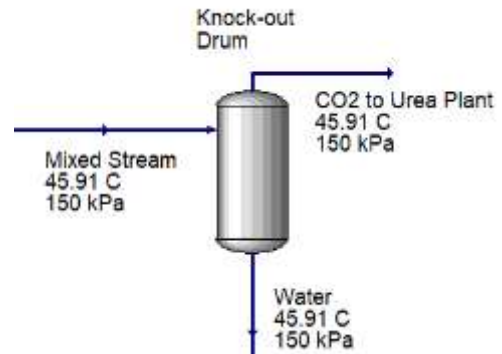


Figure 20 Knock-Out Drum

Components	Before Mixing	After Mixing
H ₂ O	0.0208	0.029
NH ₃	0	0.0003
CO ₂	0.9753	0.9668

Table 4 CO₂ Stream Before and After Mixing

3.2.5 Reflux and Bottoms Pumps

To provide the desired head to the reflux stream back to stripper and the bottoms stream through the economizer two centrifugal pumps need to be installed.

3.3 Process Flow Sheet

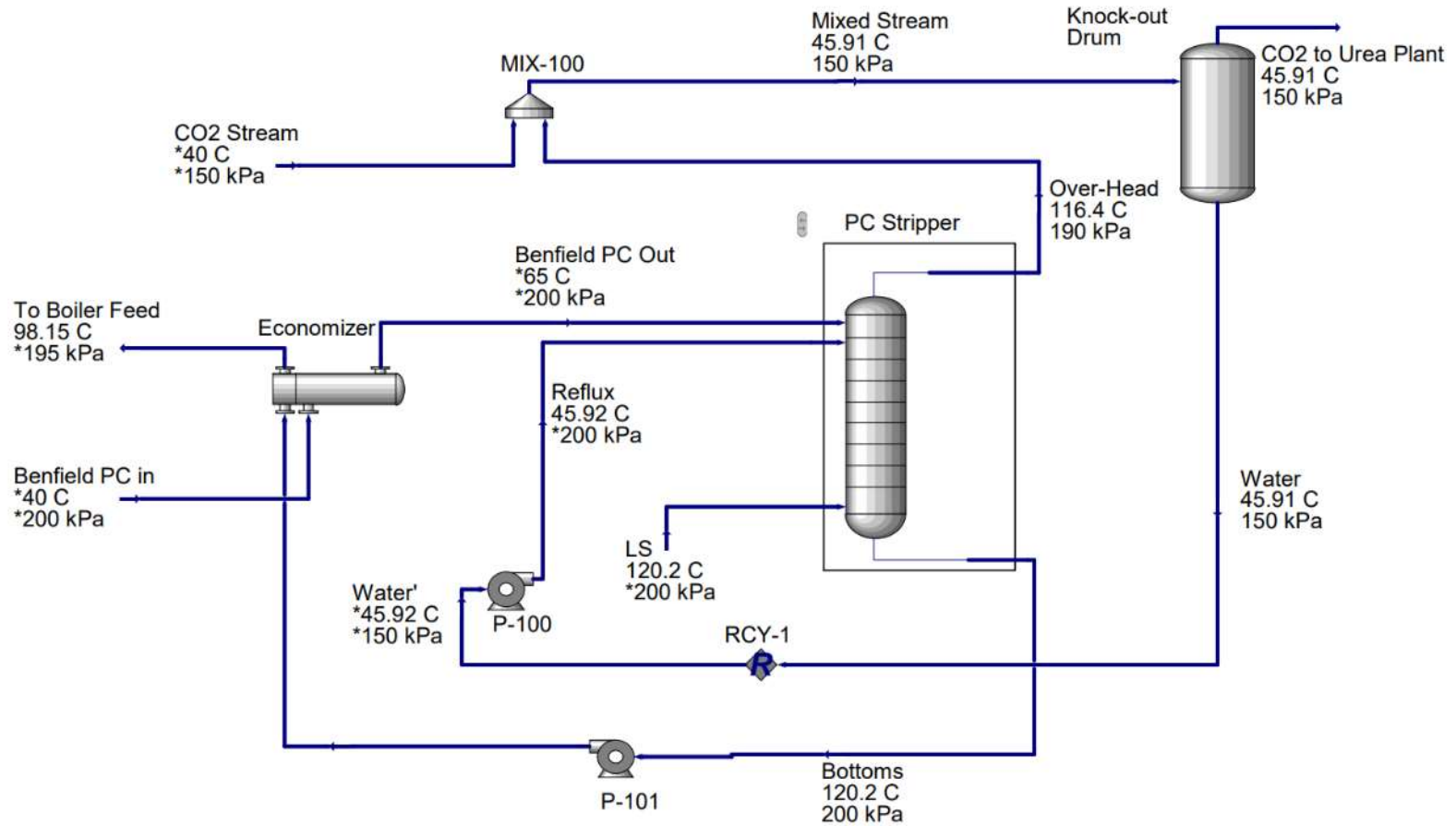


Figure 21 Process Flow Diagram of Proposed Benfield PC Stripping System

MATERIAL BALANCE

4.1 Assumptions

Material Balance involves the application of the law of conservation of mass to account for all the material that is entering and leaving a system. With the help of mass balance, the, flow rates and compositions of unknown streams can be calculated.

We performed a material balance on the CO₂ absorption unit of the ammonia plant at FFBL to calculate the amount and composition of Benfield Condensate. The assumptions that were made during analysis were:

- Basis = 1hr
- All equipment operates at steady state

This implies that in the equation,

$$\textit{Accumulation} = \textit{In} - \textit{Out} + \textit{Generation} - \textit{Consumption}$$

The Accumulation term amounts to zero.

- Ideal gas law applies
- Knockout drums operate under isobaric conditions
- There are no material losses
- The steam is purely water

To calculate the compositions of lean, semi-lean, and rich carbonate solutions we assumed the following conversions in terms of K₂CO₃,

- Lean Solution: 14-18%
- Semi-lean Solution: 25-32%
- Rich Solution: 75-80%

The process flow sheet below shows the CO₂ removal unit or the Benfield system of the ammonia plant. The material balance has been applied on each equipment step by step beginning from the Boiler E-2002. All the streams have been labelled as given

in the flow sheet. The compositions of the gaseous streams were provided by FFBL in Nm³/hr and were converted to mass flows using ideal gas law and molecular weights of the gases.

4.2 Mass Balance on Carbon Dioxide Removal Unit

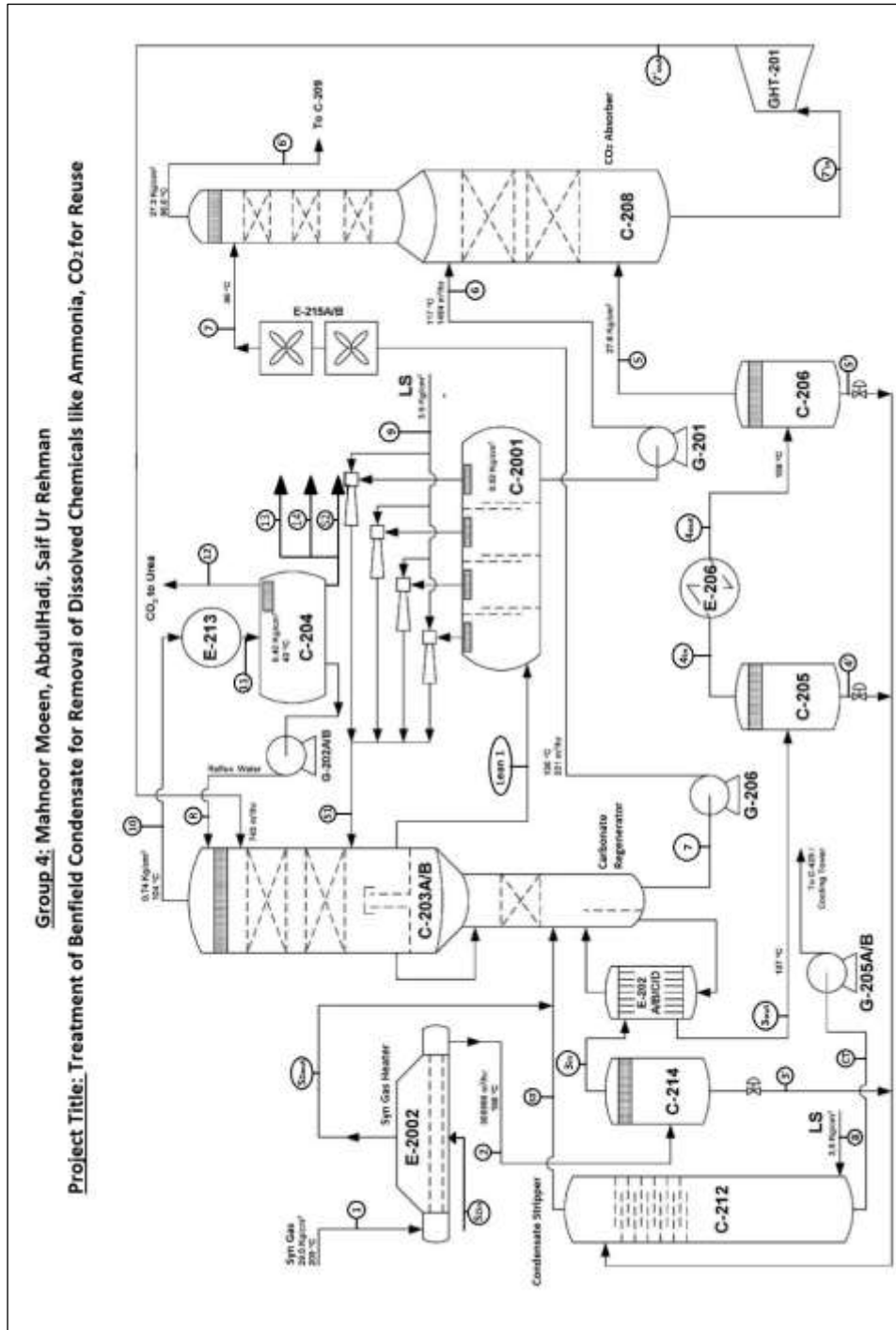
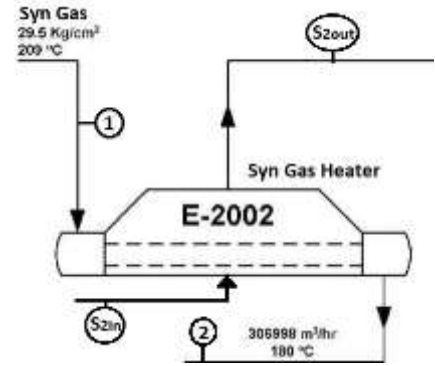


Figure 22 PFD of Carbon Dioxide Removal Unit

4.2.1 E-2002 (Boiler and Syn-Gas Heater)

This is a heat recovery boiler which uses the enthalpy of process syngas from secondary shift convertor to produce steam by heating a fraction of Benfield process condensate. The mass flow rates are conserved as shown in the table below.

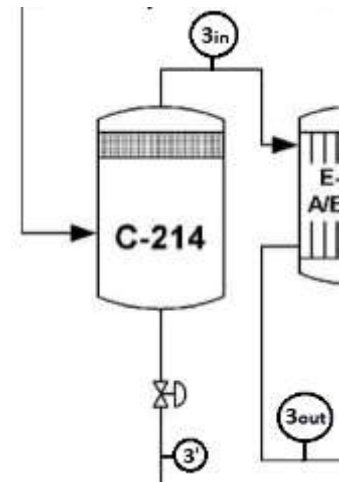


Components	Input (1) Mass flow rate (kg/h)	Output (2) Mass flow rate (kg/h)
NH ₃	-----	36.23
K ₂ CO ₃	-----	-----
H ₂ O	64201.4	64201.4
CO ₂	81143.0	81143.0
H ₂	12225.2	12225.2
CO	565.9	565.9
Ar	1004.7	1004.7
N ₂	59410	59410
CH ₄	873.7	873.7
Total	139092.3 kg/h	139092.3 kg/h

Table 5 Syngas Heater Mass Balance

4.2.2 Flash Vessel C-214

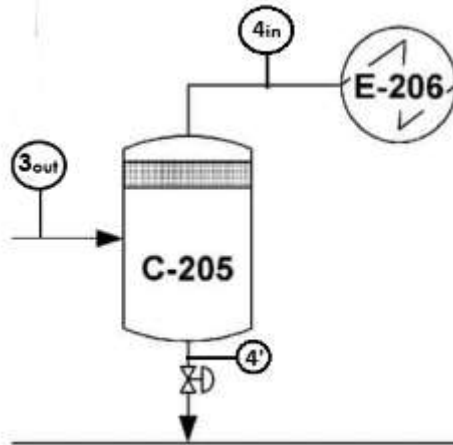
Three flash vessels are installed in series with interstage heaters. The purpose of these flash vessels is to remove excess moisture from the process syngas. This condensed moisture is then used for steam generation in condensate stripper C-212 by direct contact with low pressure steam. The input and output flows are shown.



Components	Input(2) Mass flow rate (kg/h)	Output(3) Mass flow rate (kg/h) <i>In = out</i>	Output (3') Mass flow rate (kg/h)
NH ₃	36.23	36.23	-----
K ₂ CO ₃	-----	-----	-----
H ₂ O	64201.4	64201.4	-----
CO ₂	81143.0	81143.0	-----
H ₂	12225.2	12225.2	-----
CO	565.9	565.9	-----
Ar	1004.7	1004.7	-----
N ₂	59410	59410	-----
CH ₄	873.7	873.7	-----
Total	139092.3 kg/h	139092.3 kg/h	0

Table 6 Flash Vessel C-214 Mass Balance

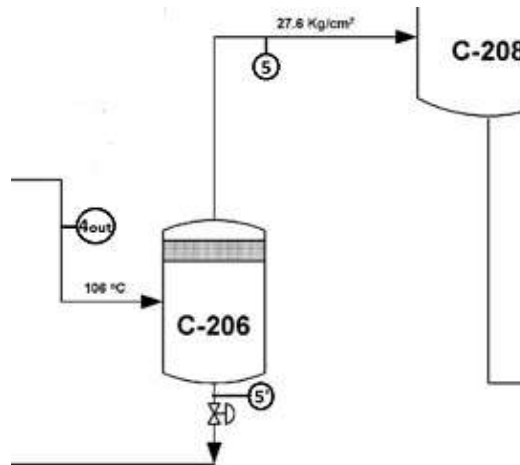
4.2.3 Flash vessel (C-205)



Components	Input (3) Mass flow rate (kg/h)	Output(4) Mass flow rate (kg/h) <i>In = out</i>	Output (4') Mass flow rate (kg/h)
NH ₃	36.23	36.23	-----
K ₂ CO ₃	-----	-----	-----
H ₂ O	64201.4	24303.4	39898
CO ₂	81143.0	81143.0	-----
H ₂	12225.2	12225.2	-----
CO	565.9	565.9	-----
Ar	1004.7	1004.7	-----
N ₂	59410	59410	-----
CH ₄	873.7	873.7	-----
Total	219,459.73 kg/h	179561.73 kg/h	39898 kg/h

Table 7 Flash Vessel C-205 Mass Balance

4.2.4 Flash Vessels (C-206)



Components	Input (4) Mass flow rate (kg/h)	Output(5) Mass flow rate (kg/h) In = out	Output (5') Mass flow rate (kg/h)
NH ₃	36.23	36.23	-----
K ₂ CO ₃	-----	-----	-----
H ₂ O	24303.4	8489	15814
CO ₂	81143.0	81143.0	-----
H ₂	12225.2	12225.2	-----
CO	565.9	565.9	-----
Ar	1004.7	1004.7	-----
N ₂	59410	59410	-----
CH ₄	873.7	873.7	-----
Total	139092.3 kg/h	83380.3 kg/h	15814 kg/h

Table 8 Flash Vessel C-206 Mass Balance

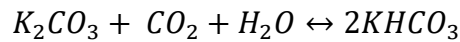
4.2.5 CO₂ Absorber

4.2.5.1 Absorber Inputs

The composition of carbonate solution used in Benfield system is

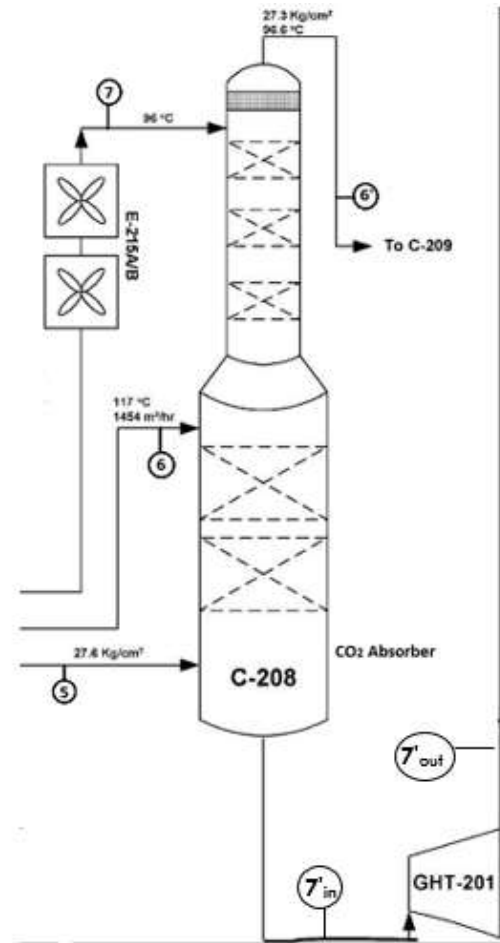
- 30% K₂CO₃
- 1% V₂O₅
- 3% DEA

During the Carbon dioxide absorption and desorption, K₂CO₃ converts to KHCO₃ as follows,



To calculate the compositions of lean, semi-lean, and rich solutions we assumed the following conversions in terms of K₂CO₃,

- Lean Solution: 14-18%
- Semi-lean Solution: 25-32%
- Rich Solution: 75-80%



Components	Stream Input(5) Mass flow rate (kg/h)
NH ₃	36.23
K ₂ CO ₃	-----
H ₂ O	8489
CO ₂	81143.0
H ₂	12225.2
CO	565.9
Ar	1004.7
N ₂	59410
CH ₄	873.7
Total	83380.3 kg/h

Table 9 CO₂ Absorber Mass Balance – Syngas Input

Input 7 (Lean Carbonate Solution)

$$\text{Temperature} = T_7 = 96^\circ\text{C}$$

$$\text{Density of 30\% carbonate solution at } 96^\circ\text{C} = 1320 \frac{\text{kg}}{\text{m}^3}$$

$$\text{Volumetric flow rate of Lean Carbonate Solution} = \frac{221\text{m}^3}{\text{hr}}$$

$$\begin{aligned}\text{Mass flow rate of Lean Carbonate Solution} &= 221 \frac{\text{m}^3}{\text{h}} \times 1320 \frac{\text{kg}}{\text{m}^3} \\ &= 291720 \frac{\text{kg}}{\text{h}}\end{aligned}$$

Conversion of K_2CO_3 in lean solution = 16%

$$\begin{aligned}\text{Amount of } \text{K}_2\text{CO}_3 \text{ in lean solution} &= 0.3 \frac{\text{kg}}{\text{kg}} \times 291720 \frac{\text{kg}}{\text{h}} \times \frac{1}{138 \frac{\text{kg}}{\text{kmol}}} \times (1 - 0.16) \\ &= 532.7 \frac{\text{kmol}}{\text{h}}\end{aligned}$$

$$\begin{aligned}\text{Amount of } \text{H}_2\text{O} \text{ in lean solution} &= 0.66 \frac{\text{kg}}{\text{kg}} \times 291720 \frac{\text{kg}}{\text{h}} \times \frac{1}{18 \frac{\text{kg}}{\text{kmol}}} \times (1 - 0.16) \\ &= 8984.9 \frac{\text{kmol}}{\text{h}}\end{aligned}$$

$$\begin{aligned}\text{Amount of } \text{KHCO}_3 \text{ in lean solution} &= 0.3 \frac{\text{kg}}{\text{kg}} \times 291720 \frac{\text{kg}}{\text{h}} \times \frac{1}{138 \frac{\text{kg}}{\text{kmol}}} \times (0.16) \times 2 \\ &= 202.64 \frac{\text{kmol}}{\text{h}}\end{aligned}$$

$$\begin{aligned}\text{Amount of } \text{V}_2\text{O}_5 \text{ in lean solution} &= 0.01 \frac{\text{kg}}{\text{kg}} \times 291720 \frac{\text{kg}}{\text{h}} \times \frac{1}{181.9 \frac{\text{kg}}{\text{kmol}}} \\ &= 16.04 \frac{\text{kmol}}{\text{h}}\end{aligned}$$

$$\begin{aligned}\text{Amount of DEA in lean solution} &= 0.03 \frac{\text{kg}}{\text{kg}} \times 291720 \frac{\text{kg}}{\text{h}} \times \frac{1}{105.14 \frac{\text{kg}}{\text{kmol}}} \\ &= 83.24 \frac{\text{kmol}}{\text{h}}\end{aligned}$$

Components	Lean solution Input (7) Molar Flow Rate kmol/h
K_2CO_3	532.7
KHCO_3	202.64
V_2O_5	16.04
DEA	83.24
H_2O	8984.9
Total	9819.52

Table 10 CO_2 Absorber Mass Balance - Lean Solution Input

Input 6 (Semi-Lean Carbonate Solution)

$$\text{Temperature} = T_6 = 117^\circ\text{C}$$

$$\text{Density of 30\% carbonate solution at } 117^\circ\text{C} = 1317 \frac{\text{kg}}{\text{m}^3}$$

$$\text{Volumetric flow rate of Semi - Lean Carbonate Solution} = 1454 \frac{\text{m}^3}{\text{hr}}$$

$$\begin{aligned} \text{Mass flow rate of Semi - Lean Carbonate Solution} &= 1454 \frac{\text{m}^3}{\text{h}} \times 1317 \frac{\text{kg}}{\text{m}^3} \\ &= 1914918 \frac{\text{kg}}{\text{h}} \end{aligned}$$

Conversion of K_2CO_3 in semi-lean solution = 28.5 %

$$\begin{aligned} \text{Amount of } \text{K}_2\text{CO}_3 \text{ in semi-lean solution} &= 0.3 \frac{\text{kg}}{\text{kg}} \times 1914918 \frac{\text{kg}}{\text{h}} \times \frac{1}{138 \frac{\text{kg}}{\text{kmol}}} \times (1 - 0.285) \\ &= 2972.14 \frac{\text{kmol}}{\text{h}} \end{aligned}$$

$$\begin{aligned} \text{Amount of } \text{H}_2\text{O} \text{ in semi-lean solution} &= 0.66 \frac{\text{kg}}{\text{kg}} \times 1914918 \frac{\text{kg}}{\text{h}} \times \frac{1}{18 \frac{\text{kg}}{\text{kmol}}} \times (1 - 0.285) \\ &= 50202.76 \frac{\text{kmol}}{\text{h}} \end{aligned}$$

$$\begin{aligned} \text{Amount of } \text{KHCO}_3 \text{ in semi-lean solution} &= 0.3 \frac{\text{kg}}{\text{kg}} \times 1914918 \frac{\text{kg}}{\text{h}} \times \frac{1}{138 \frac{\text{kg}}{\text{kmol}}} \times (0.285) \times 2 \\ &= 2369.39 \frac{\text{kmol}}{\text{h}} \end{aligned}$$

$$\begin{aligned} \text{Amount of } \text{V}_2\text{O}_5 \text{ in semi-lean solution} &= 0.01 \frac{\text{kg}}{\text{kg}} \times 1914918 \frac{\text{kg}}{\text{h}} \times \frac{1}{181.9 \frac{\text{kg}}{\text{kmol}}} \\ &= 105.3 \frac{\text{kmol}}{\text{h}} \end{aligned}$$

$$\begin{aligned} \text{Amount of DEA in semi-lean solution} &= 0.03 \frac{\text{kg}}{\text{kg}} \times 1914918 \frac{\text{kg}}{\text{h}} \times \frac{1}{105.14 \frac{\text{kg}}{\text{kmol}}} \\ &= 546.4 \frac{\text{kmol}}{\text{h}} \end{aligned}$$

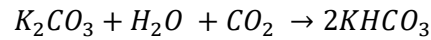
Components	Semi – Lean solution Input (6) Molar Flow Rate kmol/h
K_2CO_3	2972.14
KHCO_3	2369.39
V_2O_5	105.3
DEA	546.4
H_2O	50202.76
Total	56196

Table 11 CO_2 Absorber Mass Balance - Semi Lean Solution Input

4.2.5.2 Absorber Outputs

There are two outputs of the absorber named as 6' and 7'.

As we know the reaction in absorber is,

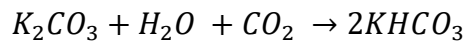


We assume the reaction is only in forward direction.

Components	Stream Output (6') Mass flow rate (kg/h) $= \frac{\text{Vol. Flow Rate}}{22.414 \text{ Nm}^3/\text{kmol}} \times \text{Mol. wt.} (\text{kg}/\text{kmol})$
NH_3	-----
K_2CO_3	-----
H_2O	3441.8
CO_2	327.83
H_2	12208.43
CO	565.9
Ar	856.6
N_2	59245.47
CH_4	873.74
Total	77519.77 kg/h

Table 12 CO_2 Absorber Mass Balance - Syngas Output

Now we define 7' by applying mass balance with chemical reaction:



$$\text{Input} = \text{output} + \text{production} - \text{consumption}$$

Moles of CO_2 absorbed = $1844.15 - 7.45 = 1836.7$ kgmole

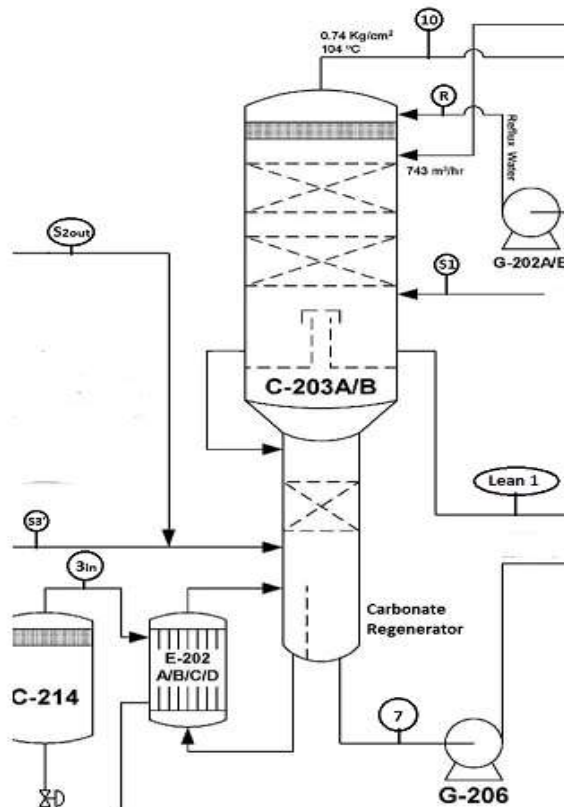
Moles of K_2CO_3 consumed in reaction = 1836.7 kgmole

Moles of H₂O consumed in reaction = 1836.7 kgmole

Moles of KHCO₃ produced in reaction = 2 × 1836.7 = 3673.4 kgmole

Components	Stream 7' Molar flow kmol/h
NH ₃	2.13
K ₂ CO ₃	1668.14
H ₂ O	57631.4
V ₂ O ₅	121.34
DEA	629.64
CO ₂	0
KHCO ₃	6245.4
H ₂	8.33
CO	0
Ar	3.705
N ₂	5.89
CH ₄	0

4.2.6 Carbonate Regenerator C-203 A/B



The carbonate regenerator is a packed stripping column in which steam is desorbs CO₂ from the potassium carbonate solution. The stripping column operates at a higher

temperature and lower pressure as compared to the CO₂ absorption column to facilitate desorption.

4.2.6.1 Inlets of Regenerator

Components	Mass flow (7') Input (kg/h)	Steam 1'	Reflux	Steam 2	Steam 3'
NH ₃	36.26	-----	43.35	7.01	-----
K ₂ CO ₃	230536.9	-----	-----	-----	-----
H ₂ O	1038287.3	37914 kg/hr	33210	5158.7	6896
CO ₂	0 kg	-----	138	80974.6	-----
KHCO ₃	624590	-----	-----	-----	-----
H ₂	16.8	-----	1.76	18.81	-----
CO	0 kg	-----	0	0	-----
Ar	148.2 kg	-----	0	0	-----
N ₂	164.92 kg	-----	4.02	0.63	-----
CH ₄	0 kg	-----	-----	-----	-----
Total		37914	33353	5397	6896

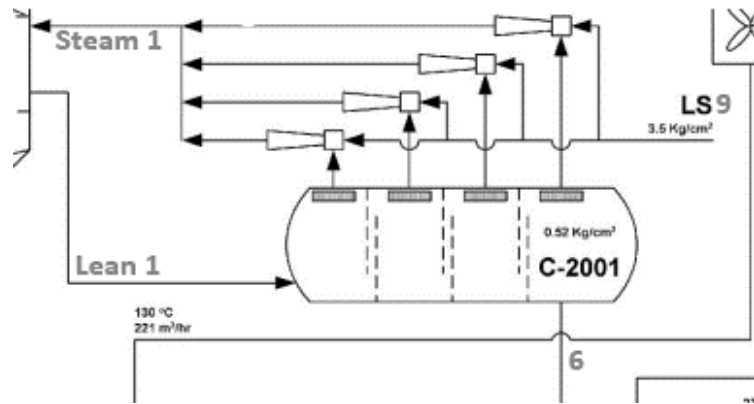
Table 13 Carbonate Regenerator Mass Balance - Inputs

4.2.6.2 Outlets of Regenerator

Components	Lean 1	Lean solution Input (7) Mass flow rate (kg/h)	V
NH ₃	-----	-----	86.65
K ₂ CO ₃	2972.14	73619.14	-----
V ₂ O ₅	105.3	2917.7	-----
DEA	546.4	8751.9	-----
KHCO ₃	2369.39	20264	-----
H ₂ O	993425	161871.96	67955
CO ₂	-----	-----	80974.6
H ₂	-----	-----	18.81
CO	-----	-----	0
Ar	-----	-----	148.12
N ₂	-----	-----	169.18
CH ₄	-----	-----	0
Total	999418	9819.52	149265

Table 14 Carbonate regenerator Mass Balance - Outputs

4.2.7 Flash Drum C-2001



Lean solution from the carbonate regenerator is first stored in flash drum C-2001. The ejectors attached to the drum are used to create vacuum with the help of steam. This vacuum suck and recovers the vaporized gases from the lean solution.

4.2.7.1 Inlets

Components	Lean 1	LS (9) Mass flow rate (kg/h)
NH ₃	-----	-----
K ₂ CO ₃	2972.14	-----
V ₂ O ₅	105.3	-----
DEA	546.4	-----
KHCO ₃	2369.39	-----
H ₂ O	993425	17641 kg/h
CO ₂	-----	-----
H ₂	-----	-----
CO	-----	-----
Ar	-----	-----
N ₂	-----	-----
CH ₄	-----	-----
Total	999418	17641 kg/h

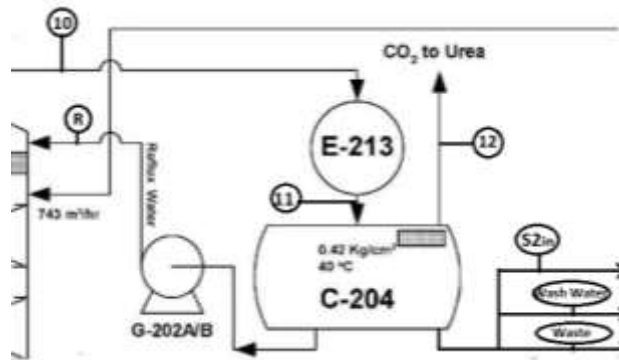
Table 15 Flash Drum C-2001 Mass Balance - Inputs

4.2.7.2 Outlets

Components	Steam 1'	Lean 1' (same as stream 6)
NH ₃	-----	nil
K ₂ CO ₃	-----	442432.88kg/hr
V ₂ O ₅	-----	19149
DEA	-----	57447.5
KHCO ₃	-----	251185kg/hr
H ₂ O	37914kg/hr	973152kg/hr
Total	37914kg/hr	1743365.9

Table 16 Flash Drum C-2001 Mass Balance - Outputs

4.2.8 C-204



C-204 is the condensate accumulator. All the overhead vapour from carbonate regenerator is sent to this vessel after cooling. The moisture content condenses whereas most of the CO₂ remains in the vapor form. This CO₂ vapor is sent to the Urea plant. The condensate is divided into four sections. Reflux, waste, and wash water, and boiler feed water.

4.2.8.1 Inlets

Component	10 (V)
NH ₃	86.65
H ₂ O	67955
CO ₂	80974.6
H ₂	18.81
CO	0
Ar	148.12
N ₂	169.18
CH ₄	0
Total	149265

Table 17 PC Accumulator Mass Balance - Input

4.2.8.2 Outlets

Component	Volumetric Flow rate of 12 (CO ₂ Vapor to Urea Stream) Nm ³ /h	Mass Flow rate of 12 (CO ₂ Vapor to Urea Stream) kg/h = $\frac{\text{Vol. Flow Rate}}{22.414 \text{ m}^3/\text{kmol}} \times \text{Mol. wt. (kg/kmol)}$
H ₂	170	15.29
N ₂	129	161.15
CO	-----	-----
CO ₂	41109	80699.38

Ar	83	148.12
CH ₄	-----	-----
H ₂ O	2141	1719.37
Total	43632	82743.31

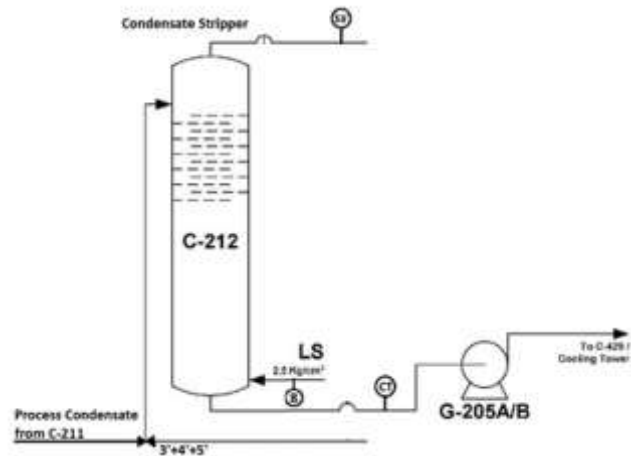
Table 18 CO₂ PC Accumulator Mass Balance - Output

Components	Reflux Mass Flow ($\frac{kg}{h}$)	13 +14 (Waste + Wash water) Mass Flow ($\frac{kg}{h}$)	Steam 2 Mass Flow ($\frac{kg}{h}$)	12 (CO ₂ vapor to Urea Stream) (V CO ₂) Mass Flow ($\frac{kg}{h}$)
NH ₃	43.4	36.23	7.02	-----
H ₂ O	33210	27867.23	5158.7	1719.37 kg/h
CO ₂	138	115.79	21.43	80699.38 kg/h
H ₂	1.76	1.48	0.274	15.29 kg/h
CO	0	0	0	0
Ar	0	0	0	148.12 kg/hr
N ₂	4.02	3.38	0.63	161.15 kg/h
CH ₄	0	0	0	-----
Total	33353	27987.65	5181	82743.31

Table 19 Distribution of Benfield PC

4.2.9 C-212

C-212 is the condensate stripper in which direct heat exchange takes place between low pressure steam and the process condensate removed from the process syngas. The steam leaving C-212 enters the carbonate regenerator.



4.2.9.1 Inlets

Components	3'+4'+5' Mass Flow ($\frac{kg}{h}$)	LS 8 Mass Flow ($\frac{kg}{h}$)
H ₂ O	0+39898+15814	7000
Total	55712	7000

4.2.9.2 Outlets

Components	Steam 3'	Cooling tower
------------	----------	---------------

	<i>Mass Flow</i> $\left(\frac{kg}{h}\right)$	<i>Mass Flow</i> $\left(\frac{kg}{h}\right)$
H2O	6896	55814

4.2.10 Stream Specifications Summary Sheet

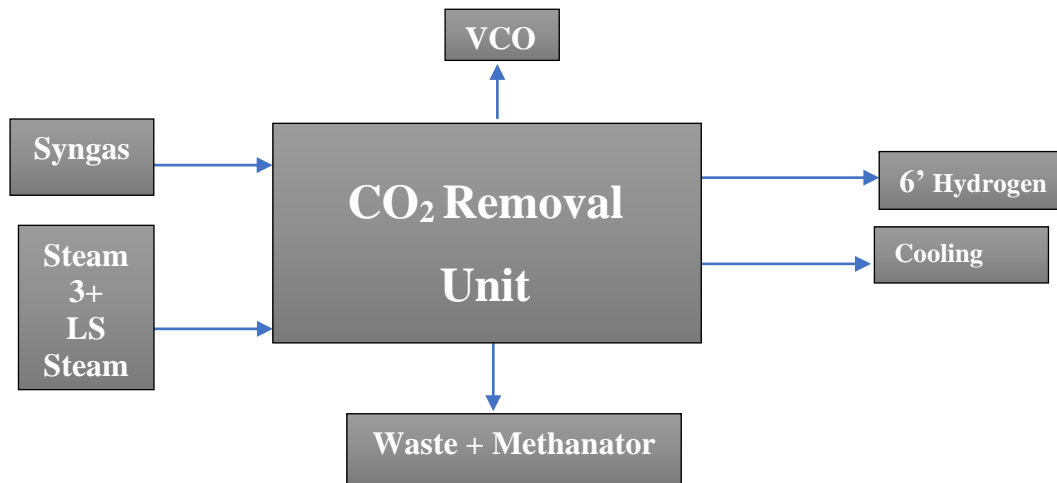
Mass flows in kg/h

Line No.	1	2	3	3'	4	4'	5	5'	6
Stream	E-2002 Inlet	E-2002 Outlet	Flash Vessel C-214 Vapour	Flash Vessel C-214 Liquid	Flash Vessel C-205 Vapour	Flash Vessel C-205 Liquid	Flash Vessel C-206 Vapour	Flash Vessel C-206 Liquid	Semi-lean Solution to Absorber
NH ₃	----	----	----	----	----	----	----	----	----
K ₂ CO ₃	----	----	----	----	----	----	----	----	57447.4
KHCO ₃	----	----	----	----	----	----	----	----	----
V2O5	----	----	----	----	----	----	----	----	19149.18
DEA	----	----	----	----	----	----	----	----	57447.54
H ₂ O	64201.4	64201.4	64201.4	----	24302.4	29898	5489	15814	128299.5
CO ₂	81143.0	81143.0	81143.0	----	81143.0	----	81143.0	----	----
H ₂	12226.2	12226.2	12226.2	----	12226.2	----	12226.2	----	----
CO	565.9	565.9	565.9	----	565.9	----	565.9	----	----
Ar	1004.7	1004.7	1004.7	----	1004.7	----	1004.7	----	----
N ₂	59410	59410	59410	----	59410	----	59410	----	----
CH ₄	873.7	873.7	873.7	----	873.7	----	873.7	----	----
Total	219423.6	219423.6	138092.3	0	99194.3	29898	83380.3	15814	191491.8
Temperature °C	208	180	180	180	137	137	108	108	117
Pressure kg/cm ²	29.5	28.8	28.8	28.8	28.2	28.2	27.6	27.6	----

Line No.	7	7'	8	9	10	11	12	13+14
Stream	Lean Solution to Absorber	Rich Solution from Absorber	LS to C-212	LS Steam to C-201	Vapors to E-213	To C-204	CO ₂ Stream to Urea	Waste + Washwater
NH ₃	----	----	----	----	----	----	----	----
K ₂ CO ₃	8751.6	652595.8	----	----	----	----	----	----
KHCO ₃	----	68148.3	----	----	----	----	----	----
V2O5	8751.6	8751.6	----	----	----	----	----	----
DEA	2917.2	2917.2	----	----	----	----	----	----
H ₂ O	195452.4	1450335.7	7000	17641	67955	67955	1719.13	27867
CO ₂	----	----	----	----	80974.6	80974.6	80699.2	115.79
H ₂	----	17.34	----	----	18.81	18.81	15.29	1.48
CO	----	----	----	----	----	----	----	----
Ar	----	148.2	----	----	148.12	148.12	148.12	----
N ₂	----	164.92	----	----	169.18	169.18	161.15	3.38
CH ₄	----	----	----	----	----	----	----	----
Total	291720	2183079	7000	17641	149265	149265	43632	28000
Temperature °C	96	125	----	152	104	40	40	----
Pressure kg/cm ²	----	----	2.5	3	0.74	0.42	0.50	----

Line No.	S1	S2	S3'	R	CT
Stream	From Flash Drum C-2001 to Absorber	From Flash Drum C-204 to E-2002	From C-212 to C-203 A/B	From Flash Drum C-204 to C-203 A/B	C-212 Bottom Product
NH ₃	----	----	----	----	----
K ₂ CO ₃	+	----	----	----	----
KHCO ₃	----	----	----	----	----
V2O5	----	----	----	----	----
DEA	----	----	----	----	----
H ₂ O	37914	5397	6896	33210	55814
CO ₂	----	----	----	138	----
H ₂	----	----	----	1.76	----
CO	----	----	----	----	----
Ar	----	----	----	----	----
N ₂	----	----	----	4.02	----
CH ₄	----	----	----	----	----
Total	37914	5397	6896	33353	55814
Temperature °C	----	40	126	40	126
Pressure kg/cm ²	----	----	----	----	----

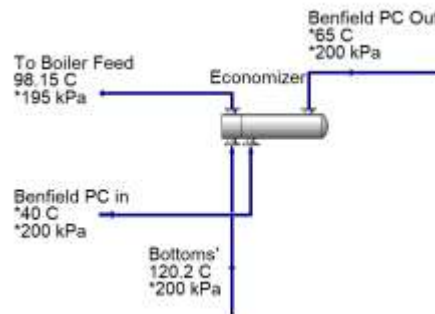
4.2.11 Overall Mass Balance on Carbon Dioxide Removal Unit



Component	Syn Gas $\frac{kg}{h}$	Steam 3 $\frac{kg}{h}$	LP $\frac{kg}{h}$	6', $\frac{kg}{h}$	VCO ₂ $\frac{kg}{h}$	Cooling tower $\frac{kg}{h}$	Waste + Wash Water $\frac{kg}{h}$
NH ₃	36.23	-----	-----	-----	-----	-----	36.23
H ₂ O	64201.4	7000	17641	3441.8	1719.37	55814	27867
CO ₂	81143.0	-----	-----	327.83	80699.38	-----	115.79
H ₂	12225.2	-----	-----	12208.43	15.29	-----	1.48
CO	565.9	-----	-----	565.9	-----	-----	0
Ar	1004.7	-----	-----	856.6	148.12	-----	0
N ₂	59410	-----	-----	59245.47	161.15	-----	3.38
CH ₄	873.7	-----	-----	873.73	-----	-----	0

Table 20 CO₂ Removal Unit Overall Material Balance

4.3 Material Balance on Proposed Benfield PC Stripping System



4.3.1 Economizer

Economizer - Overall				
Quantity	Shell-Side Hot Fluid		Tube-Side Cold Fluid	
	Bottoms-in	Bottoms-out	Benfield PC-out	Benfield PC-in
Temp (°C)	120.2	98.16	65	40
P (bar)	2	1.95	2	2
Mass Flow (kg/h)	22850	22850	20530	20530

Table 21 Mass balance on Economizer

4.3.1.1 Shell-Side Hot Fluid

Hot fluid is stripped process condensate that leaves the stripper bottoms at 120°C.

Shell-Side Hot Fluid	
Component	Mass Flow (kg/h)
H ₂ O	22849.67
CO ₂	3.743750x 10 ⁻⁶
NH ₃	0.3000152
Total	22850

Table 22 Shell-Side Fluid in Economizer

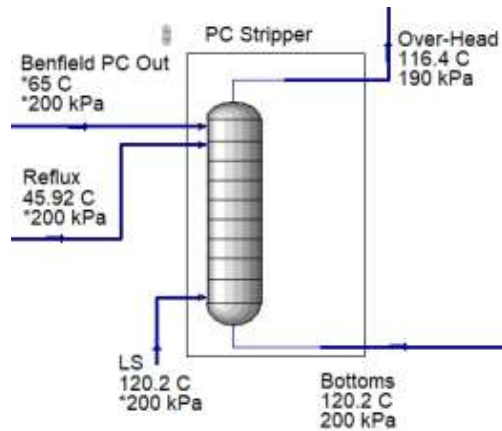
4.3.1.2 Tube -Side Cold Fluid

The process condensate feed at 40°C is the tube side fluid which gains heat from the stripped process condensate.

Tube-Side Cold Fluid	
Component	Mass Flow (kg/h)
H ₂ O	20421.00
CO ₂	80.00000
NH ₃	29.00000
Total	20530

Table 23 Tube Side Fluid in Economizer

4.3.2 Benfield Process Condensate Stripper



4.3.2.1 Inlet Streams

Benfield PC Stripper Inlets flow (kg/h)			
Component	Benfield PC - Out (from Economizer)	LS (Low Pressure Steam)	Reflux
H ₂ O	20421.00	3100.00	153.95
CO ₂	80.00	0.00	6.85
NH ₃	29.00	0.00	2.76
Total	20530	3100	163.57

Table 24 PC Stripper Inlets

4.3.2.2 Outlet Streams

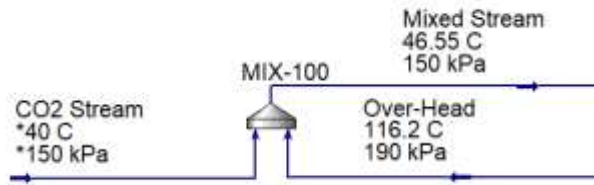
Benfield PC Stripper Outlets flow (kg/h)		
Component	Over-Head	Bottoms-in (to Economizer)
H ₂ O	825.27	22849.67
CO ₂	86.85	3.74 x 10 ⁻⁶
NH ₃	31.46	0.30
Total	943.59	22849.97

Table 25 PC Stripper Outlets

Benfield PC Stripper - Overall					
Quantity	Benfield PC - out	LP Steam	Reflux	Bottoms-in	Vapour OVHD
T (°C)	65	111.3	45.82	120.2	116.3
P (bar)	2	2	2	2	190
Mass Flow (kg/h)	20530.00	3100.00	163.57	22849.97	943.59

Table 26 PC Stripper Overall Mass Balance

4.3.3 Mixing Tee



Mixing Tee - Overall			
Quantity	CO ₂ Stream	Over-Head	Mixed Stream
T (°C)	40	116.2	46.55
P (bar)	1.5	1.9	1.5
Mass Flow (kg/h)	82742.56	943.60	83686.15

Table 27 Mixing Tee Material Balance

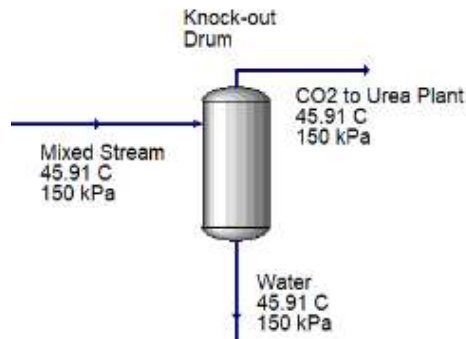
4.3.3.1 Inlet Streams

Mixing Tee Inlet flow (kg/h)		
Component	CO ₂ Stream	Over-Head
H ₂ O	1719.00	825.27
CO ₂	80699.00	86.85
NH ₃	0.00	31.46
Argon	148.12	0.00
CO	0.00	0.00
Hydrogen	15.29	0.00
Nitrogen	161.15	0.00
Total	82742.56	943.60

4.3.3.2 Outlet Streams

Mixing Tee Outlet flow (kg/h)	
Component	Mixed Stream
H ₂ O	2544.28
CO ₂	80785.85
NH ₃	31.46
Argon	148.12
CO	0.00
Hydrogen	15.29
Nitrogen	161.15
Total	83686.15

4.3.4 Knock-Out Drum



Knockout Drum - Overall			
Quantity	Mixed Stream	CO ₂ to Urea	Water
T (°C)	46.55	46.55	46.55
P (bar)	1.5	1.5	1.5
Mass Flow (kg/h)	83686.15	83522.53	163.63

Table 28 Knock-out Drum Material Balance

4.3.4.1 Inlet Streams

Knock-Out Drum Inlet flow (kg/h)	
Component	Mixed Stream
H ₂ O	2544.28
CO ₂	80785.85
NH ₃	31.46
Argon	148.12
CO	0.00
Hydrogen	15.29
Nitrogen	161.15
Total	83686.15

Table 29 Knockout Drum Inlets

4.3.4.2 Outlet Streams

Knock-Out Drum Outlet flow (kg/h)		
Component	CO₂ to Urea	Water
H ₂ O	2390.30	153.97
CO ₂	80778.97	6.88
NH ₃	28.69	2.78
Argon	148.12	0.00
CO	0.00	0.00
Hydrogen	15.29	0.00
Nitrogen	161.15	0.00
Total	83522.53	163.63

Table 30 Knockout Drum Outlets

ENERGY BALANCE

5.1 Assumptions

- A C_p vs. temperature graph at the partial pressures of the components was generated through Aspen HYSYS and the average of C_p at $T_{ref} = 25^\circ\text{C}$ and T_{stream} was calculated. The graphs are compiled at the end of this chapter.
- Since water changes phase between the reference temperature and the stream temperatures, its latent heat of vaporization was also accounted for while calculating the enthalpy. The C_p of water also changes on the change of phase. The calculations for calculating the enthalpy of water are shown separately.
- Aspen HYSYS was used to find the latent heat of vaporization of water at given conditions.
- The enthalpy of a stream includes enthalpy of formation of components and sensible heat of stream. To make simplify calculations for enthalpy balance without chemical reaction, standard enthalpies of formation (H_f°) are not added since they simply cancel out in the inlet and outlet streams. Hence, the enthalpies of the streams were found as follows:

$$Total\ enthalpy\ of\ stream = \sum_i^{i=n} m_i \left(\int_{25}^T C_{p_i} dT \right) + H_{water}$$

Where,

m_i = mas flow rate of component 'i' in kg/h

C_{p_i} = Specific heat capacity of component 'i'

dT = differential change in temperature

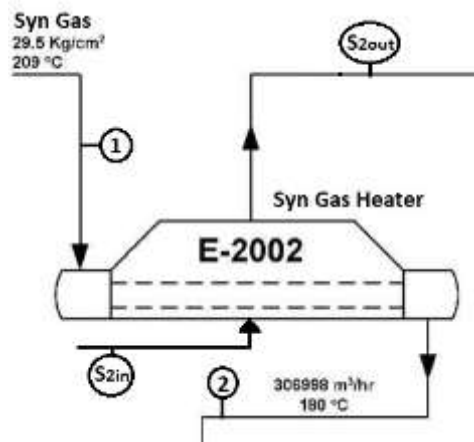
H_{water} = total enthalpy of water in the stream

- The enthalpy calculations are done taking Basis = 1 hour.

Before designing our proposed Benfield PC stripping system, an energy balance was applied on the CO₂ Removal Unit of Ammonia Plant where the Benfield PC comes from. It helped us understand the working of the Benfield process and estimate the energy flows and compositions of the Benfield Process condensate. The balance of CO₂ removal unit is discussed first followed by the energy balance of our proposed Benfield PC stripping (refer to Figure 22 for PFD).

5.2 Energy Balance on CO₂ Removal Unit

5.2.1 Syn Gas Heater (E-2002)



Enthalpy Balance Equation

Using Syngas enthalpies to calculate heat duty of syngas heater.

$$\text{Heat Duty in Given Data} = Q = -3.21 \text{ Gcal/h}$$

$$\Delta H = H_2 - H_1 = 224613744.5 - 237897255.9 = -13283511 \text{ kJ/h}$$

$$\text{Calculated Heat Duty} = Q = -\frac{13280968.4 \text{ kJ}}{\text{hr}} = -3.17 \text{ Gcal/h}$$

5.2.1.1 Inlet (1)

$$\text{Temperature} = T_1 = 209^\circ\text{C}$$

$$\text{Pressure} = 30.5 \text{ kg/cm}^2$$

Components	Stream (1) Mass flow kg/h	Partial Pressure kg/cm²	$C_{p,avg}$ kJ/kg.K	Enthalpy H1 (excluding H_f°) kJ
NH ₃	36.23	0.005	2.26	15065.88
H ₂ O	64201.4	7.93	-----	179455753.3
CO ₂	81143.0	4.118	0.95	14183796.4
H ₂	12225.2	13.51	14.225	31998238.48
CO	565.9	0.046	1.0515	109488.0684
Ar	1004.7	0.055	0.5211	96333.04728
N ₂	59410	4.73	1.064	11631052.16
CH ₄	873.7	0.122	2.535	407528.628
Total	139092.3	30.5	-----	237897255.9

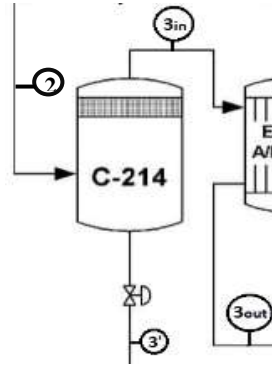
5.2.1.2 Outlet (2)

$$\text{Temperature} = T_1 = 180^\circ\text{C}$$

$$\text{Pressure} = 29.8 \text{ kg/cm}^2$$

Components	Stream (2) Mass flow kg/h	Partial Pressure kg/cm²	$C_{p,avg}$ kJ/kg.K	Enthalpy H2 (excluding H_f°) kJ
NH₃	36.23	0.005	2.23	12522.9
H₂O	64201.4	7.93	-----	175590829
CO₂	81143.0	4.118	0.9375	11791092.19
H₂	12225.2	13.51	14.215	26936088.79
CO	565.9	0.046	1.049	92012.5105
Ar	1004.7	0.055	0.5211	81150.12135
N₂	59410	4.73	1.0615	9774875.825
CH₄	873.7	0.122	2.475	335173.1625
Total	139092.3	30.5	-----	224613744.5

5.2.2 Flash Vessel C-214



Enthalpy Balance Equation

Considering Adiabatic Flash Vessel,

$$H_2 = H_{3in} + H_{3'}$$

$$224613744.5 = 224613744.5 + 0$$

5.2.2.1 Inlet (2)

Calculations already described in syngas heater.

Components	Stream (2) Mass flow kg/h	Pressure kg/cm ²	Temperature °C	Enthalpy H ₂ (excluding H _f [°]) kJ
Total	139092.3	30.5	180	224613744.5

5.2.2.2 Outlet (3_{in})

$$\text{Temperature} = T_3 = 180^\circ\text{C}$$

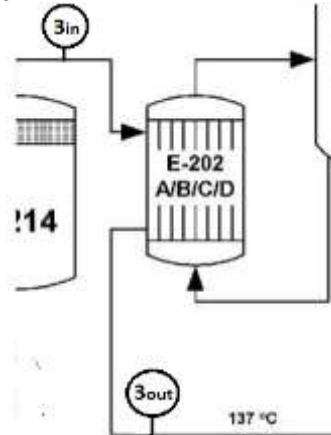
$$\text{Pressure} = 29.8 \text{ kg/cm}^2$$

Components	Stream (3 _{in}) Mass flow kg/h	Partial Pressure kg/cm ²	C _{p,avg} kJ/kg.K	Enthalpy H _{3in} (excluding H _f [°]) kJ
NH ₃	36.23	0.005	2.23	12522.9
H ₂ O	64201.4	7.75	-----	175590829
CO ₂	81143.0	4.118	0.9375	11791092.19
H ₂	12225.2	13.51	14.215	26936088.79
CO	565.9	0.046	1.049	92012.5105
Ar	1004.7	0.055	0.5211	81150.12135
N ₂	59410	4.73	1.0615	9774875.825
CH ₄	873.7	0.122	2.475	335173.1625
Total	139092.3	29.8	-----	224613744.5

5.2.2.3 Outlet (3')

Components	Stream (3')
	Mass flow rate kg/h
Total	0

5.2.3 Heaters (E-202 A/B/C/D)



Enthalpy Balance Equation

Using syngas enthalpies to calculate heater duty.

$$\text{Given Heat Duty } Q = -24.74 \text{ Gcal}$$

$$H_{3_{out}} - H_{3_{in}} = Q$$

$$119021915.7 - 224613744.5 = -105591828.8 \text{ kJ}$$

$$\text{Calculated Heat Duty } Q = -25.23 \text{ Gcal}$$

5.2.3.1 Inlet 3_{in}

Calculations already described in Flash Vessel C-214.

Components	Stream (3 _{in}) Mass flow rate kg/h	Pressure kg/cm^2	Temperature $^{\circ}C$	Enthalpy H _{3_{in}} (excluding H_f°) kJ
Total	139092.3	29.8	180	224613744.5

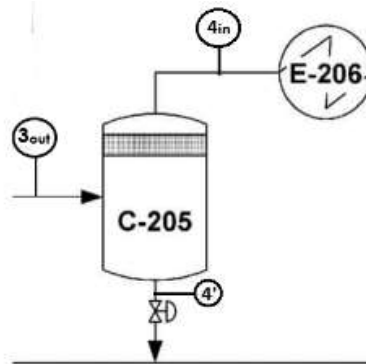
5.2.3.2 Outlet 3_{out}

$$\text{Temperature} = T_3 = 137^{\circ}C$$

$$\text{Pressure} = 29.2 \text{ } kg/cm^2$$

Components	Stream (3 _{out}) Mass flow rate (kg/h)	Partial Pressure kg/cm ²	C _{p,avg} kJ/kg.K	Enthalpy H _{3out} (excluding H _f ^o) kJ
NH ₃	36.23	0.005	2.20	8927.1
H ₂ O	64201.4	7.59	-----	83734481.2
CO ₂	81143.0	3.94	0.9275	8429134.84
H ₂	12225.2	12.93	14.205	19449804.19
CO	565.9	0.044	1.045	66232.936
Ar	1004.7	0.053	0.5211	58637.50704
N ₂	59410	4.53	1.058	7039847.36
CH ₄	873.7	0.112	2.4	234850.56
Total	139092.3	29.2	-----	119021915.7

5.2.4 Flash Vessel (C-205)



Enthalpy Balance Equation

Considering Adiabatic Flash Vessel,

$$H_{3out} = H_{4in} + H_{4'}$$

$$119021915.7 = 102011747.4 + 19214876.8$$

$$119021915.7 \text{ kJ} \approx 121226624.2 \text{ kJ}$$

5.2.4.1 Inlet (3_{out})

Calculations already described in Heaters (E-202 A/B/C/D).

Components	Stream (3 _{out}) Mass flow rate kg/h	Pressure kg/cm ²	Temperature °C	Enthalpy H _{3out} (excluding H _f ^o) kJ
Total	139092.3	29.2	137	119021915.7

5.2.4.2 Outlet 4'

$$\text{Temperature} = T_4 = 137^\circ\text{C}$$

$$\text{Pressure} = 29.2 \text{ kg/cm}^2$$

Components	Stream(4') Mass flow rate kg/h	Partial Pressure kg/cm^2	$C_{p,avg}$ kJ/kg.K	Enthalpy H4' (excluding H_f°) kJ
H ₂ O	39898	29.2	4.3	19214876.8
Total	39898	29.2	4.3	19214876.8

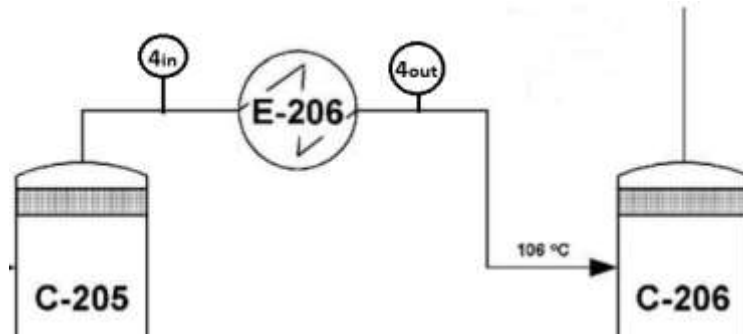
5.2.4.3 Outlet 4_{in}

$$\text{Temperature} = T_{4,in} = 137^\circ\text{C}$$

$$\text{Pressure} = 29.2 \text{ kg/cm}^2$$

Components	Stream(4 _{in}) Mass flow rate kg/h	Partial Pressure kg/cm^2	$C_{p,avg}$ kJ/kg.K	Enthalpy H4 _{in} (excluding H_f°) kJ
NH ₃	36.23	0.005	2.20	8927.1
H ₂ O	24303.4	7.59	4.5	12248913.6
CO ₂	81143.0	3.94	0.9275	8429134.84
H ₂	12225.2	12.93	14.205	19449804.19
CO	565.9	0.044	1.045	66232.936
Ar	1004.7	0.053	0.5211	58637.50704
N ₂	59410	4.53	1.058	7039847.36
CH ₄	873.7	0.112	2.4	234850.56
Total	179561.93	29.2	-----	102011747.4

5.2.5 Heat exchanger (E-206)



Enthalpy Balance Equation

Using Syngas enthalpies to calculate heat duty.

$$\text{Given Heat Duty } Q = -11.04 \text{ Gcal}$$

$$H_{4_{out}} - H_{4_{in}} = Q$$

$$53286076.3 - 102011747.4 = -48725671.4 \text{ kJ}$$

$$\text{Calculated Heat Duty} = -11.64 \text{ Gcal}$$

5.2.5.1 Inlet 4_{in}

Calculations already described in Flash Vessel (C-205)

Components	Stream (4 _{in}) Mass flow rate kg/h	Pressure kg/cm^2	Temperature $^{\circ}\text{C}$	Enthalpy H _{4_{in}} (excluding H_f°) kJ
Total	179561.93	29.2	137	102011747.4

5.2.5.2 Output 4_{out}

$$\text{Temperature} = T_{4_{out}} = 106^{\circ}\text{C}$$

$$\text{Pressure} = 28.6 \text{ kg/cm}^2$$

Components	Stream (4 _{out}) Mass flow rate kg/h	Partial Pressure kg/cm^2	$C_{p,avg}$ kJ/kg.K	Enthalpy H _{4_{out}} (excluding H_f°) kJ
NH ₃	36.23	0.0052	2.16	6338.8
H ₂ O	24303.4	3.25	4.5	27873554.4
CO ₂	81143.0	4.44	0.919	6040203.777

H₂	12225.2	14.57	14.175	14036669.01
CO	565.9	0.05	1.043	47808.9297
Ar	1004.7	0.061	0.5212	42415.62084
N₂	59410	5.11	1.054	5072069.34
CH₄	873.7	0.13	2.36	167016.492
Total	179561.93	27.6	-----	53286076.3

5.2.5.3 Flash Vessel (C-206)

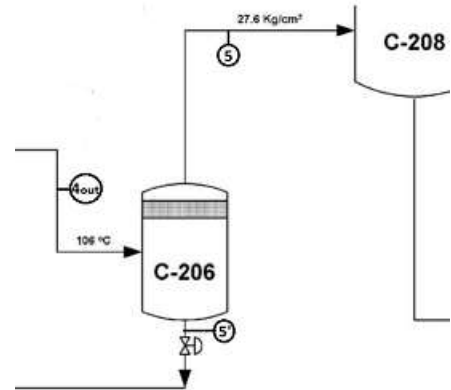
Enthalpy Balance Equation

Considering Adiabatic Flash Vessel,

$$H_{4_{out}} = H_5 + H_{5'}$$

$$53286076.3 = 47649967 + 5636109.6$$

$$53286076.3 \text{ kJ} = 53286076.3 \text{ kJ}$$



5.2.5.4 Inlet 4_{out}

Calculations already described in heat exchanger (E-206).

Components	Stream (4 _{out}) Mass flow kg/h	Pressure kg/cm ²	Temperature °C	Enthalpy H _{4_{out}} (excluding H _f [°]) kJ
Total	179561.93	27.6	106	53286076.3

5.2.5.5 Outlet 5'

$$\text{Temperature} = T_5 = 106^\circ\text{C}$$

$$\text{Pressure} = 28.6 \text{ kg/cm}^2$$

Components	Stream (5') Mass flow rate kg/h	Pressure kg/cm ²	C _{p,avg} kJ/kg.K	Enthalpy H _{5'} (excluding H _f [°]) kJ
H₂O	15814	0.95	4.4	5636109.6

5.2.5.6 Outlet 5

$$\text{Temperature} = T_5 = 106^\circ\text{C}$$

$$\text{Pressure} = 28.6 \text{ kg/cm}^2$$

Components	Stream (5) Mass flow kg/h	Partial Pressure kg/cm ²	$C_{p,avg}$ kJ/kg.K	Enthalpy H5 (excluding H_f°) kJ
NH ₃	36.23	0.005544	2.16	6338.8
H ₂ O	8489	1.227518	4.5	22237444.8
CO ₂	81143.0	4.800127	0.919	6040203.777
H ₂	12225.2	15.78412	14.175	14036669.01
CO	565.9	0.052604	1.043	47808.9297
Ar	1004.7	0.065462	0.5212	42415.62084
N ₂	59410	5.522764	1.054	5072069.34
CH ₄	873.7	0.141857	2.36	167016.492
Total	163747.53	27.6	-----	47649967

5.2.6 CO₂ Absorber (C-208)

Enthalpy Balance Equation

Enthalpy of streams in = Enthalpy of streams out

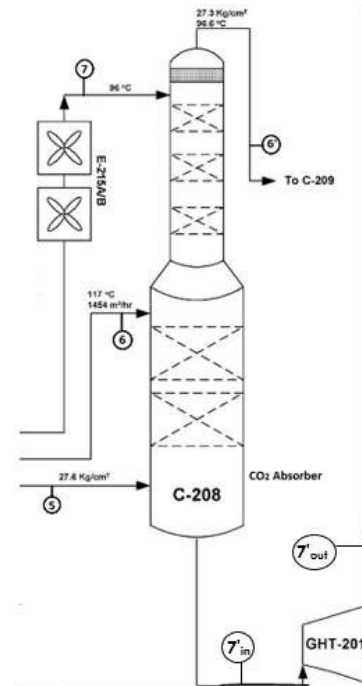
$$H5 + H6 + H7 = H6' + H7_{in}'$$

$$47649967 + (-1.816 \times 10^{10}) + (-3060450289)$$

$$= 26069217 + (-2.21 \times 10^{10})$$

$$-2.11 \times 10^{10} \approx -2.19 \times 10^{10}$$

Here the difference can be accounted for heat losses in the absorber column.



5.2.6.1 Inlet 5

Detailed calculations have been described in flash vessel (C-206).

Components	Stream (5) Mass flow kg/h	Partial Pressure kg/cm ²	Temperature °C	Enthalpy H5 (excluding H_f°) kJ
Total	163747.53	27.6	106	47649967

5.2.6.2 Inlet 6 (Semi-lean Potassium Carbonate Solution)

For the potassium carbonate solutions, the value of molar enthalpy was taken from aspen hysys at the relevant compositions and conditions.

Components	Semi – Lean solution Inlet (6) Molar Flow Rate kmol/h
K₂CO₃	2972.14
KHCO₃	2369.39
V₂O₅	105.3
DEA	546.4
H₂O	50202.76
Total	56196

Molar Enthalpy of stream 6 = -323153.58 kJ/kmol

$$H6 = -323153.58 \frac{\text{kJ}}{\text{kmol}} \times 56196 \frac{\text{kmol}}{\text{h}} = -1.816 \times 10^{10} \frac{\text{kJ}}{\text{h}}$$

5.2.6.3 Inlet 7 (Lean Potassium Carbonate Solution)

Components	Lean solution Inlet (7) Molar Flow Rate kmol/h
K₂CO₃	532.7
KHCO₃	202.64
V₂O₅	16.04
DEA	83.24
H₂O	8984.9
Total	9819.52

Molar Enthalpy of stream 7 = -311670.05 kJ/kmol

$$H7 = -311670.05 \frac{\text{kJ}}{\text{kmol}} \times 9819.52 \frac{\text{kmol}}{\text{h}} = -3060450289 \frac{\text{kJ}}{\text{h}}$$

5.2.6.4 Outlet 6' (Sweet Syngas)

$$\text{Temperature} = T_{6'} = 96.6 \text{ }^{\circ}\text{C}$$

$$\text{Pressure} = 28.3 \text{ kg/cm}^2$$

Components	Stream (6') Mass flow kg/h	Partial Pressure kg/cm ²	$C_{p,avg}$ kJ/kg.K	Enthalpy H6' (excluding H_f°) kJ
NH ₃	-----	-----	-----	-----
H ₂ O	3441.8	0.616546	-----	8941934.1
CO ₂	327.83	0.024022	0.903	21184
H ₂	12208.43	19.52648	14.23	12434408.1
CO	565.9	0.065166	1.043	42244.5
Ar	856.6	0.069132	0.521	31972.7
N ₂	59245.47	6.822598	1.049	4449832.5
CH ₄	873.74	0.176055	2.36	147641.1
Total	77519.77	27.3	-----	26069217

5.2.6.5 Output 7' (CO₂ Rich Potassium Carbonate Solution)

$$\text{Temperature} = T_{7'} = 125 \text{ }^{\circ}\text{C}$$

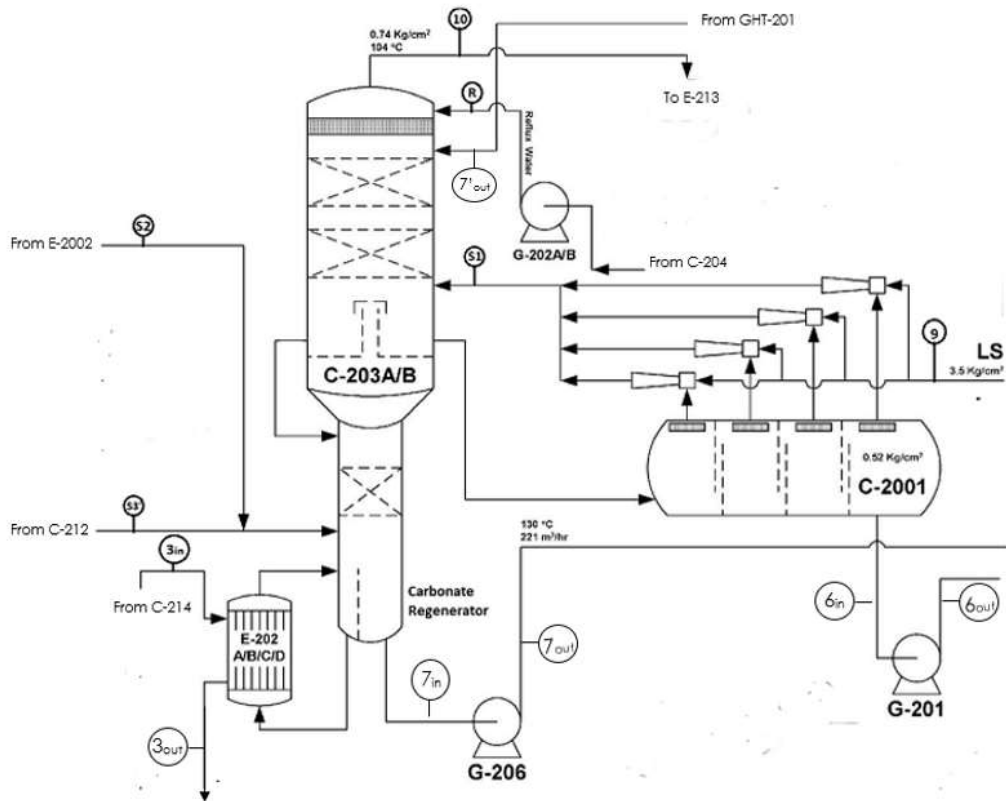
Components	Rich Solution (7') Molar Flow Rate kmol/h
NH ₃	2.13
K ₂ CO ₃	1668.14
H ₂ O	57631.4
V ₂ O ₅	121.34
DEA	629.64
CO ₂	0
KHCO ₃	6245.4
H ₂	8.33
CO	0
Ar	3.705
N ₂	5.89
CH ₄	0
Total	66316

$$\text{Molar Enthalpy of stream 7} = -333306.06 \text{ kJ/kmol}$$

$$H7 = -333306.06 \frac{\text{kJ}}{\text{kmol}} \times 66316 \frac{\text{kmol}}{\text{h}} = -2.21 \times 10^{10} \frac{\text{kJ}}{\text{h}}$$

5.2.7 Sub-System 1

The Carbonate Regenerator (C-203 A/B), Flash drum (C-2001) and heat exchangers E-202 (A/B/C/D) were clumped together as one system to apply energy balance on this part of the plant. This was done since the temperature and pressure conditions of some intermediate streams such as steam S1 entering the regenerator from flash drum C-2001 were unknown and difficult to determine. Hence, an overall balance was applied on this subsystem.



Enthalpy Balance Equation

Enthalpy of streams in = Enthalpy of streams out

$$\begin{aligned}
 H_{3_{in}} + H_{S2} + H_9 + H_{S3'} + H_{7'_{out}} + H_R &= H_{3_{out}} + H_{6_{out}} + H_{7_{out}} + H_{10} \\
 224613744.5 + 1123410.4 \text{ kJ} + 47198495.5 + 17579553 + (-2.2097 \times 10^{10}) \\
 + 2.20 \times 10^6 \\
 &= 119021915.7 + (-1.816 \times 10^{10}) + (-3060450289) + 17043354 \\
 &\quad - 2.18 \times 10^{10} \approx -2.09 \times 10^{10}
 \end{aligned}$$

Here the difference can be accounted for heat losses in the stripper column.

5.2.7.1 Inlet S2_{out} (from E-2002)

$$\begin{aligned}
 S_{2,out} &= S_{2,in} + Q = \dot{m}_2 Cp(T_{2,in} - T_{ref}) + Q \\
 &= 5397 \frac{kg}{h} \times 4.4 \frac{kJ}{kg^\circ C} \times (40 - 25)^\circ C + 3.21 \frac{Gcal}{h} \times \frac{10^6 kJ}{4.184 Gcal} \\
 S_{2,out} &= 356202 + 767208.4 = 1123410.4 kJ
 \end{aligned}$$

5.2.7.2 Inlet S3' (from C-212)

$$Temperature = 126^\circ C$$

$$Pressure = 1 kg/cm^2$$

$$H3' @ 126^\circ C, 1 \frac{kg}{cm^2}$$

$$= m_{S3'} \left(\int_{25}^{100} Cp_{avg} dT + \int_{130}^{126} Cp_{avg} dT + H_{latent} \right)$$

$$= 6896 \left(\int_{25}^{100} 4.5 dT + \int_{100}^{126} 2.0 dT + 2159 \right)$$

$$HS3' = 17579553 \frac{kJ}{h}$$

5.2.7.3 Inlet 7'_{out} (CO₂ Rich Potassium Carbonate Solution)

$$H7_{out} = H7_{in} + Q_{GHT-201}$$

$$H7_{out} = -2.21 \times 10^{10} \frac{kJ}{h} + 743 \frac{kJ}{s} \times 3600 \frac{s}{h} = -2.2097 \times 10^{10} \frac{kJ}{h}$$

5.2.7.4 Inlet 9 (Low Pressure Steam)

$$Temperature = 152^\circ C$$

$$Pressure = 4 kg/cm^2$$

$$H9 @ 152^\circ C, 4 \frac{kg}{cm^2}$$

$$H9 = m_9 \left(\int_{25}^{130} Cp_{avg} dT + \int_{130}^{152} Cp_{avg} dT + H_{latent} \right)$$

$$= 17641 \left(\int_{25}^{130} 4.5 dT + \int_{130}^{152} 2.0 dT + 2159 \right)$$

$$H_9 = 47198495.5 \frac{kJ}{h}$$

5.2.7.5 Outlet 10 (Potassium Carbonate Solution Regenerator Overhead)

$$Temperature = 104^\circ C$$

$$Pressure = 1.74 \text{ kg/cm}^2$$

Component	Stream 10 Mass Flow Rate Kg/h	Partial Pressure kg/ cm ²	C _{p,avg} kJ/kg.K	Enthalpy H10 (excluding H _f ^o) kJ
NH ₃	86.65	0.00087	2.16	14785.956
H ₂ O	67955	1.1658	-----	164498668
CO ₂	80974.6	0.56724	0.919	5878836.93
H ₂	18.81	0.002871	14.175	21063.9083
CO	0	0	1.043	0
Ar	148.12	0.001218	0.5212	6098.81138
N ₂	169.18	0.0018618	1.054	14086.9419
CH ₄	0	0	2.36	0
Total	149265	1.74	-----	170433541

5.2.7.6 Inlet R (Reflux)

$$Temperature = T_1 = 40^\circ C$$

$$P = 1.42 \frac{kg}{cm^2}, \text{Subcooled Liquid}$$

Component	Stream (R) Mass flow Kg/h
NH ₃	43.4
H ₂ O	33210
CO ₂	138
H ₂	1.76
CO	0
Ar	0
N ₂	4.02
CH ₄	0
Total	33353

Since the composition of reflux is mainly water, we calculated the enthalpy of reflux using the Cp of water at the given pressure and temperature.

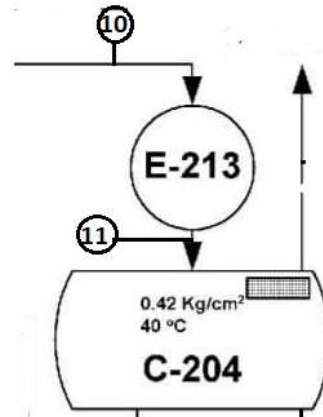
$$H_R = m_R \left(\int_{25}^{40} C_{p,avg} dT \right)$$

$$= 33353 \left(\int_{25}^{40} 4.5 dT \right) = 2.20 \times 10^6 \text{ kJ}$$

5.2.8 Regenerator OVHD Condenser (E-213, E-213 A/B/C)

Enthalpy Balance Equation

$$\begin{aligned}
 H_{10} - H_{11} &= Q_{E-213, E-213} \\
 170433541 - 10075387.5 \\
 &= 40.82 \text{ Gcal} \times 4.18 \\
 &\times 10^6 \frac{\text{kJ}}{\text{Gcal}} \\
 160358153.5 &\approx 170709240
 \end{aligned}$$



5.2.8.1 Inlet 10

Component	Stream 10 Mass Flow kg/h	Partial Pressure kg/cm ²	Temperature °C	Enthalpy H10 (excluding H_f°) kJ
Total	149265	1.74	104	170433541

5.2.8.2 Outlet H11

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

$$Pressure = 1.5 \text{ kg/cm}^2$$

$$H_{11@ 40^\circ\text{C}, 1.5 \frac{\text{kg}}{\text{cm}^2}}$$

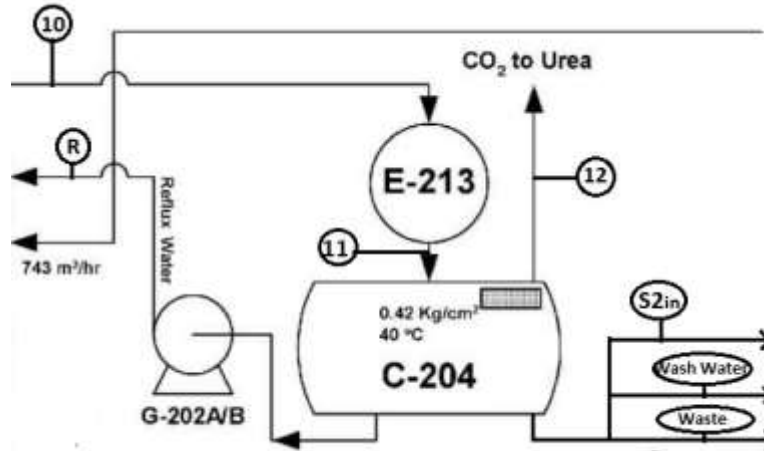
$$= m_{11} \left(\int_{25}^{40} C_{p_{avg}} dT \right)$$

$$= 149265 \left(\int_{25}^{40} 4.5 dT \right)$$

$$H_{11} = 10075387.5 \text{ kJ}$$

5.2.9 Condensate Accumulator C-204

Enthalpy Balance Equation



$$H_{11} = H_R + H_{S2_{in}} + H_{washwater} + H_{waste} + H_{12}$$

$$10075387.5 = 2.20 \times 10^6 + 356202 + 4.49 \times 10^5 + 1.35 \times 10^6 + 5350019.06$$

$$10075387.5 \approx 9705221$$

5.2.9.1 Output 12

$$Temperature = T_1 = 40^\circ C$$

Components	12 (CO ₂ vapor to Urea Stream) (V CO ₂) Mass Flow kg/h	Partial Pressure kg/cm^2	$C_{p,avg}$ $kJ/kg.K$	Enthalpy H_{3in} (excluding H_f°) kJ
NH ₃	-----	-----	-----	-----
H ₂ O	1719.37 kg/h	0.078	4.5	4289828
CO ₂	80699.38 kg/h	1.41	0.9375	1053126.909
H ₂	15.29 kg/h	0.006	14.215	3231.5415
CO	-----	-----	-----	-----
Ar	148.12 kg/hr	0.003	0.5211	1155.336
N ₂	161.15 kg/h	0.0045	1.0615	2677.2735
CH ₄	-----	-----	-----	-----
Total	82743.31	1.5	-----	5350019.06

5.2.9.2 Output Wash Water

$$\text{Temperature} = T_1 = 40^\circ\text{C}$$

$$P = 1.42 \text{ kg/cm}^2, \text{ sub cooled liquid}$$

$$\begin{aligned} H_{\text{washwater}} &= m_W \left(\int_{25}^{40} C_{p_{\text{avg}}} dT \right) \\ &= 6805 \left(\int_{25}^{40} 4.4 dT \right) = 4.49 \times 10^5 \text{ kJ} \end{aligned}$$

5.2.9.3 Output Waste Water

$$\text{Temperature} = T_1 = 40^\circ\text{C}$$

$$P = 1.42 \text{ kg/cm}^2, \text{ sub cooled liquid}$$

$$\begin{aligned} &= m_{\text{waste}} \left(\int_{25}^{40} C_{p_{\text{avg}}} dT \right) \\ &= 20530 \left(\int_{25}^{40} 4.4 dT \right) = 1.35 \times 10^6 \text{ kJ} \end{aligned}$$

5.2.10 Summarized Energy Balance on Carbon Dioxide Removal Unit

Equipment	P kg/cm ²	Temp IN °C	Temp OUT °C	H _{IN} kJ/h	Syn Gas mass flow OUT	H _{OUT} kJ/h	Energy Balance Equation	ΔH = Q kJ/h
Syn Gas Heater (E-2002)	30.5	209	180	2.38 × 10 ⁸	219,459.7	2.25 × 10 ⁸	Q = H ₂ – H ₁	–1.32 × 10 ⁷
Flash Vessel C-214	29.8	180	180	2.25 × 10 ⁸	219,459.7	2.25 × 10 ⁸	H ₂ = H _{3in} + H _{3'}	0
Heaters (E-202 A/B/C/D)	29.8	180	137	2.24 × 10 ⁸	219,459.7	1.19 × 10 ⁸	H _{3out} – H _{3in} = Q	–1.06 × 10 ⁸
Flash Vessel (C-205)	29.2	137	137	1.19 × 10 ⁸	179561.93	1.02 × 10 ⁸	H _{3out} = H _{4in} + H _{4'}	1.92 × 10 ⁷ *
Heat exchanger (E-206)	28.6	137	106	1.02 × 10 ⁸	179561.93	5.32 × 10 ⁷	H _{4out} – H _{4in} = Q	–4.87 × 10 ⁷
Flash Vessel (C-206)	27.6	106	106	5.32 × 10 ⁷	163747.53	4.76 × 10 ⁷	H _{4out} = H ₅ + H _{5'}	–5.63 × 10 ⁶ **
CO ₂ Absorber (C-208)	28	106	96	–2.117 × 10 ¹⁰	77519.77* **	–2.21 × 10 ¹⁰	H ₅ + H ₆ + H ₇ = H _{6'} + H _{7in'}	0

Table 31 Summarized Energy Balance on CO₂ Removal Unit

*This is the energy content as 39898 Kg/hr water is removed from flash vessel C-205.

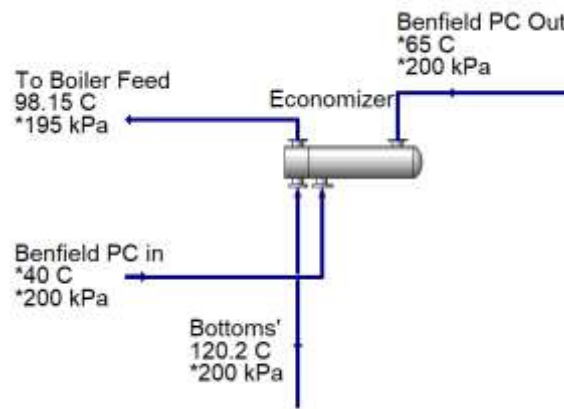
** This is the energy content as 15814 Kg/hr water is removed from flash vessel C-206.

***This the flow rate syngas after the removal of CO₂ gas in absorber column.

5.3 Energy Balance on Proposed PC Stripping System

5.3.1 Economizer

The stripped Benfield process condensate leaving as stripper bottoms is used to preheat the process condensate feed which helps decrease NH₃ and CO₂ solubility in water. In this way waste heat from the process is utilized to enhance the separation carried out by stripper. The temperature of stripped process condensate decreases from 120.2 to 98.16 °C while the PD feed is heated from 40 to 65°C.



Enthalpy Balance Equation

$$\text{Heat Duty } Q = -\Delta H(\text{Bottoms}) = \Delta H(\text{Benfield PC})$$

$$Q = 355768654.73 - 355768654.73 = 2,217,906.49 \frac{\text{kJ}}{\text{h}}$$

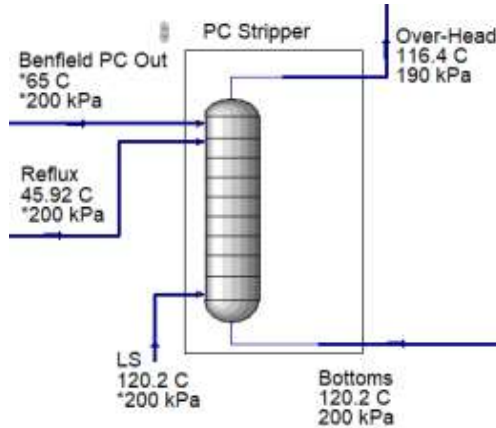
$$\text{Heat Duty } Q = 6161 \text{ kW}$$

Economizer - Overall				
Quantity	Shell-Side Hot Fluid		Tube-Side Cold Fluid	
	Bottoms-in	Bottoms-out	Benfield PC-out	Benfield PC-in
Temp (°C)	120.2	98.16	65	40
P (bar)	2	1.95	2	2
Heat Flow (kJ/h)	-353550749.24	-355768654.73	-321757283.55	-323975189.05
Heat Duty Q = 6161 kW				

Table 32 Energy Balance on Economizer

5.3.2 Benfield Process Condensate Stripper

The heated process condensate feed at 65°C counter currently contacts low pressure steam from the plant utilities at 120°C and 2 bar pressure. The stripped process condensate leaves the stripper at 120.2 °C whereas the vapor carrying removed gases exits at 116.4 °C.



Enthalpy Balance Equation

$$\Delta H_{in} = \Delta H_{out}$$

$$H_{Benfield\ PC - out} + H_{LS\ Steam} + H_{Reflux} = H_{Bottoms-in} + H_{overhead}$$

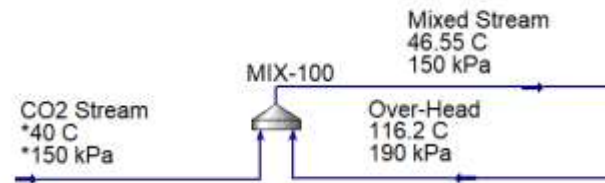
There is no heat accumulation in the stripper and energy remains conserved.

Benfield PC Stripper - Overall					
Quantity	Benfield PC - out	LS Steam	Reflux	Bottoms-in	Overhead
Temp (°C)	65	111.3	45.82	120.2	116.3
P (bar)	2	2	2	2	190
Heat Flow (kJ/h)	-3.218 x 10 ⁸	-4.108 x 10 ⁷	-2.506 x 10 ⁶	-3.536 x 10 ⁸	-1.179 x 10 ⁷

Table 33 Energy Balance on PC Stripper

5.3.3 Mixing Tee

Since the vapor overhead can not be discarded into the atmosphere due to environmental considerations, it



is combined with the CO₂ stream that is separated from syngas. Since the flow rate

of vapor is much less as compared to flow rate of CO₂ stream, a very small change in CO₂ stream conditions takes place.

Enthalpy Balance Equation

$$\Delta H_{in} = \Delta H_{out}$$

$$H_{CO_2 \text{ Stream}} + H_{Over-Head} = H_{Mixed \text{ Stream}}$$

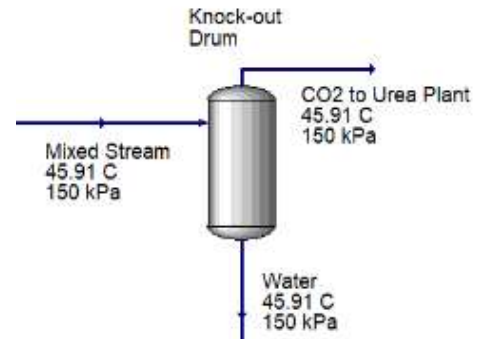
There is no heat accumulation in the mixing tee and energy remains conserved.

Mixing Tee - Overall			
Quantity	CO ₂ Stream	Over-Head	Mixed Stream
T (°C)	40	116.2	46.55
P (bar)	1.5	1.9	1.5
Heat Flow (kJ/h)	-744156518.28	-11790171.78	-755946690.06

Table 34 Energy Balance on Mixing Tee

5.3.4 Knock-Out Drum

The knock-out drum separates the excess moisture from mixed stream and refluxes it back to PC stripper. The vapor stream leaving the knock-out drum now carries inherent moisture only.



Enthalpy Balance Equation

Considering Adiabatic flash vessel,

$$\Delta H_{in} = \Delta H_{out}$$

$$H_{Mixed \text{ Stream}} = H_{CO_2 \text{ to Urea}} + H_{\text{water reflux}}$$

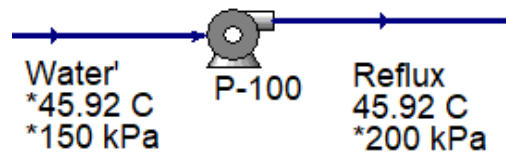
There is no heat accumulation in the mixing tee and energy remains conserved.

Knock Out Drum - Overall			
Quantity	Mixed Stream	CO ₂ to Urea	Water
T (°C)	46.55	46.55	46.55
P (bar)	1.5	1.5	1.5
Heat Flow (kJ/h)	-755946690.06	-753440350.25	-2506339.81

Table 35 Energy Balance on Knock-out Drum

5.3.5 Reflux Pump

The power-driven reflux pump provides a head for the PC reflux to flow. Isentropic compression increases the



pressure of the incoming stream of process condensate from 1.5 bar to 2 bar. The efficiency of the pump is taken to be 75%.

Isentropic Work, $\Delta H_s = -2103000 + 2103008.9 = 8.9 \text{ kJ/h}$

Actual Work, Assuming an efficiency of 75%,

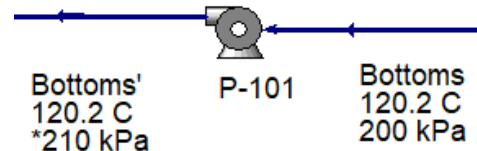
$$W = \Delta H = \frac{\Delta H_s}{\eta} = \frac{8.9}{0.75} = 11.87 \text{ kJ/h}$$

Quantity	Condensed Liquid	Reflux
Temperature (°C)	45.81	45.82
Pressure (bar)	1.5	2
Mass Flow (kg/h)	133.2	133.2
Heat Flow (kJ/h)	-2103008.9	-2103000
Isentropic Work	8.9 kJ/h	
Actual Work	11.87 kJ/h	
Power	0.0033 kW	

Table 36 Energy Balance on Reflux Pump

5.3.6 Bottoms Pump

The stripped PC bottoms are provided the required head to counter the pressure drop in economizer by the power-driven bottoms.



Isentropic compression increases the pressure of the incoming stream of process condensate from 2 bar to 2.1 bar. The efficiency of the pump is taken to be 75%.

Isentropic Work, $\Delta H_s = -353100327 + 353100327 = 327 \text{ kJ/h}$

Actual Work, Assuming an efficiency of 75%,

$$W = \Delta H = \frac{\Delta H_s}{\eta} = \frac{327}{0.75} = 436 \text{ kJ/h}$$

Quantity	Bottoms in	Bottoms out
Temperature (°C)	120.2	120.2
Pressure (bar)	200	210
Mass Flow (kg/h)	22850	22850
Heat Flow (kJ/h)	-353100000	-353100327
Isentropic Work	327 kJ/h	
Actual Work	436 kJ/h	
Power	0.121 kW	

Table 37 Energy Balance on Bottoms Pump

EQUIPMENT DESIGN

6.1 Benfield Process Condensate Stripper

The design of process condensate stripper consisted of the following step

- Estimating the required theoretical stages using empirical correlations
- Finding theoretical stages using Ammonia-water vapor liquid equilibrium (VLE) data obtained from Aspen Plus after generating the electrolyte chemistry of components
- Finding plate efficiency using Van Winkle correlation
- Calculating real number of stages
- Choosing plate type and plate spacing
- Calculating column height
- Finding column diameter

6.1.1 Minimum Number of Stages – Fenske Equation (1932)

Fenske equation can be written as,

$$\left(\frac{x_{LK}}{x_{HK}}\right)_d = \alpha_{LK}^{N_m} \left(\frac{x_{LK}}{x_{HK}}\right)_b \dots \dots \dots (i)$$

Where,

$\frac{x_{LK}}{x_{HK}}$ = ratio of concentration of light key to heavy key component

N_m = minimum stages at total reflux

α_{LK} = Average relative volatility of light key component

Equation (i) can be re-arranged as,

$$N_m = \frac{\log \left(\frac{x_{LK}}{x_{HK}}\right)_d \left(\frac{x_{LK}}{x_{HK}}\right)_b}{\log \alpha_{LK}} \dots \dots \dots (ii)$$

In our case, the light key component is ammonia and the heavy key component is water. Since the volatility of carbon dioxide is many times greater than ammonia, it does not require extensive separation. Hence ammonia is taken as the light key component.

Required Mole Fractions of Ammonia and Carbon dioxide			
Component	xNH ₃	xCO ₂	xH ₂ O
Vapor Distillate	0.038906	0.042914	0.91818
Bottoms	0.000001	0	0.999999

$$\left(\frac{x_{LK}}{x_{HK}}\right)_d = \frac{0.038906}{0.91818} = 0.0424$$

$$\left(\frac{x_{LK}}{x_{HK}}\right)_b = \frac{0.000001}{0.999999} = 0.0000001$$

$$\alpha_{LK} = 9.5$$

$$N_m = \frac{\log\left(\frac{x_{LK}}{x_{HK}} \times \frac{x_{LK}}{x_{HK}}\right)}{\log\alpha_{LK}} = 6.43$$

6.1.2 Minimum Reflux Ratio (R_m) - Underwood (1948)

Underwood used to find minimum reflux ratio is as follows,

$$\sum \frac{\alpha_i x_{i,d}}{\alpha_i - \theta} = R_m + 1 \dots \dots (iii)$$

α_i = the relative volatility of light key with respect to heavy key

R_m = the minimum reflux ratio,

$x_{i,d}$ = concentration of light key in the vapor distillate at minimum reflux

Here θ is the root of equation,

$$\sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 1 - q \dots \dots (iv)$$

Above equation can be re-arranged as,

$$q + \sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 1$$

$x_{i,f}$ = concentration of component 'i' in the feed and q depends on feed condition

$$q = 1 + \frac{c_{pL}(T_{bubble} - T_f)}{\Delta H_{vap}} \dots \dots (v)$$

$$q = 1 + \frac{77.73 \frac{kJ}{kg} (95.37 - 65)}{43190}$$

$$q = 1.051$$

By trial and error we find that at $\theta = 9.17$

$$q + \sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = q + \frac{\alpha_{LK} x_{LK,f}}{\alpha_{LK} - \theta} + \frac{\alpha_{HK} x_{HK,f}}{\alpha_{HK} - \theta}$$

$$q + \sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 0.969$$

Hence the value of $\theta = 9.16$ can be taken as a satisfactory root. From equation (iii)

$$R_m = \sum \frac{\alpha_i x_{i,d}}{\alpha_i - \theta} - 1$$

$$R_m = \frac{\alpha_{LK} x_{LK,f}}{\alpha_{LK} - \theta} + \frac{\alpha_{HK} x_{HK,f}}{\alpha_{HK} - \theta} - 1$$

$$R_m = 0.042$$

Our reflux ratio $R = 0.178$

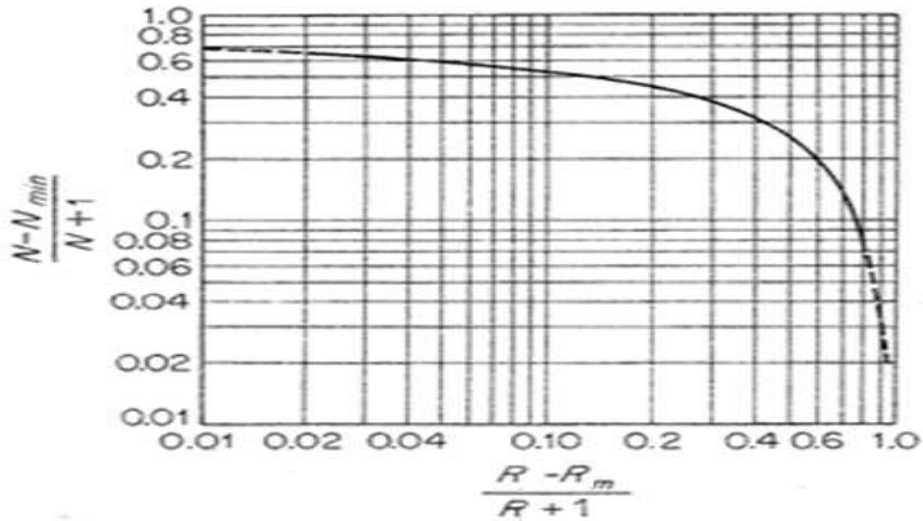


Figure 23 Gilliland's Chart for Number of Stages

Using Gilliland's equation-based chart above,

At,

$$\frac{R - R_m}{R + 1} = 0.114$$

$$\frac{N - N_m}{N + 1} = 0.5$$

Which gives us,

$$N = 13.85$$

6.1.3 Number of stages by McCabe Thiele Method

6.1.3.1 VLE from Aspen Plus

Since our system included volatile weak electrolyte solutes, McCabe Thiele Method was also used to find the number of stages for process condensate stripper to verify the calculations through Fenske and Underwood method. To consider the ionization reactions of NH₃ and CO₂ in our vapour-liquid equilibrium curve, we generated the chemistry of molecular solutes using Electrolyte wizard.

Component ID	Type	Component name	Alias
H2O	Conventional	WATER	H2O
NH3	Conventional	AMMONIA	H3N
CO2	Conventional	CARBON-DIOXIDE	CO2
H3O+	Conventional	H3O+	H3O+
NH4+	Conventional	NH4+	NH4+
AMMON(S)	Solid	AMMONIUM-CARBAMATE	NH2COONH4
SALT1	Solid	AMMONIUM-HYDROGEN-CA...	NH4HCO3
NH2COO-	Conventional	CARBAMATE	NH2COO-
HCO3-	Conventional	HCO3-	HCO3-
OH-	Conventional	OH-	OH-
CO3--	Conventional	CO3--	CO3-2

Figure 24 Ammonia Water VLE - Components

The equilibrium constants were retrieved from Aspen Plus database. A binary Analysis was run to generate the T-x-y and y-x curves of NH₃. To adjust the scales, these curves were re-plotted in Excel by taking values from the results section. This VLE curve was used to find number of stages by McCabe Thiele Method.

Method of specifying chemistry

Specify reactions Specify reactive components
 Specify inert components All components are reactive

Reaction stoichiometry

	Reaction	Type	Stoichiometry
▶	1	Equilibrium	NH ₃ + HCO ₃ ⁻ <--> H ₂ O + NH ₂ COO ⁻
▶	2	Equilibrium	NH ₃ + H ₂ O <--> OH ⁻ + NH ₄ ⁺
▶	3	Equilibrium	H ₂ O + HCO ₃ ⁻ <--> CO ₃ ²⁻ + H ₃ O ⁺
▶	4	Equilibrium	2 H ₂ O + CO ₂ <--> HCO ₃ ⁻ + H ₃ O ⁺
▶	5	Equilibrium	2 H ₂ O <--> OH ⁻ + H ₃ O ⁺
▶	AMMON(S)	Salt	AMMON(S) <--> NH ₂ COO ⁻ + NH ₄ ⁺
▶	SALT1	Salt	SALT1 <--> HCO ₃ ⁻ + NH ₄ ⁺

New Edit

Figure 25 Ammonia Water VLE - Chemistry

Binary Analysis Tabulate Calculation Options Diagnostics Results Comments Status

Analysis type: **Txy**

Components

Component 1: **H2O**

Component 2: **NH3**

Pseudo-Binary

Pseudo-binary system

Entrainer:

Entrainer fraction:

Compositions

Basis: **Mole fraction**

Vary: **NH3**

Equidistant Logarithmic List of values

Start point:

End point:

Number of intervals:

Increment:

Pressure

Units: **bar**

Equidistant List of values

Enter Values:

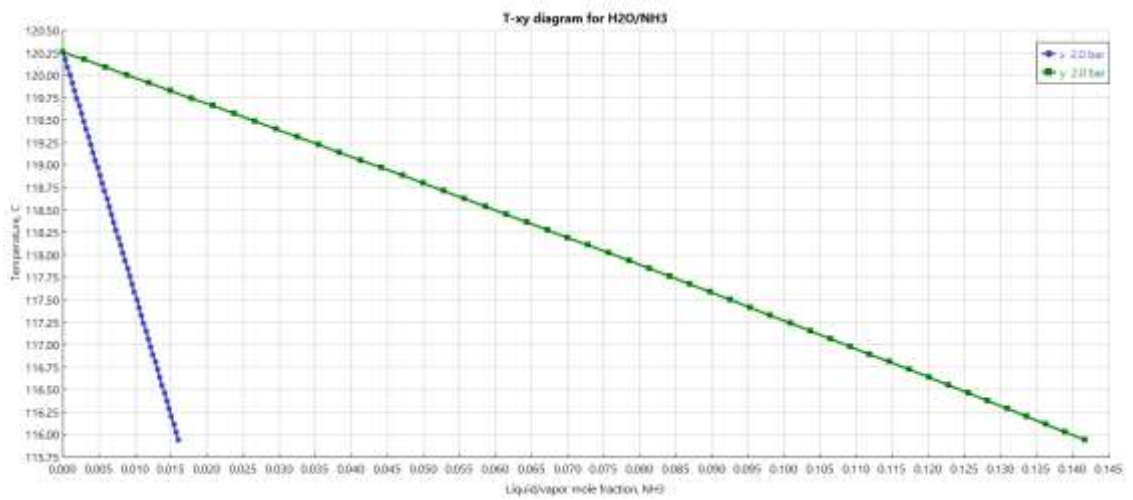


Figure 26 Ammonia Water T-x-y Diagram

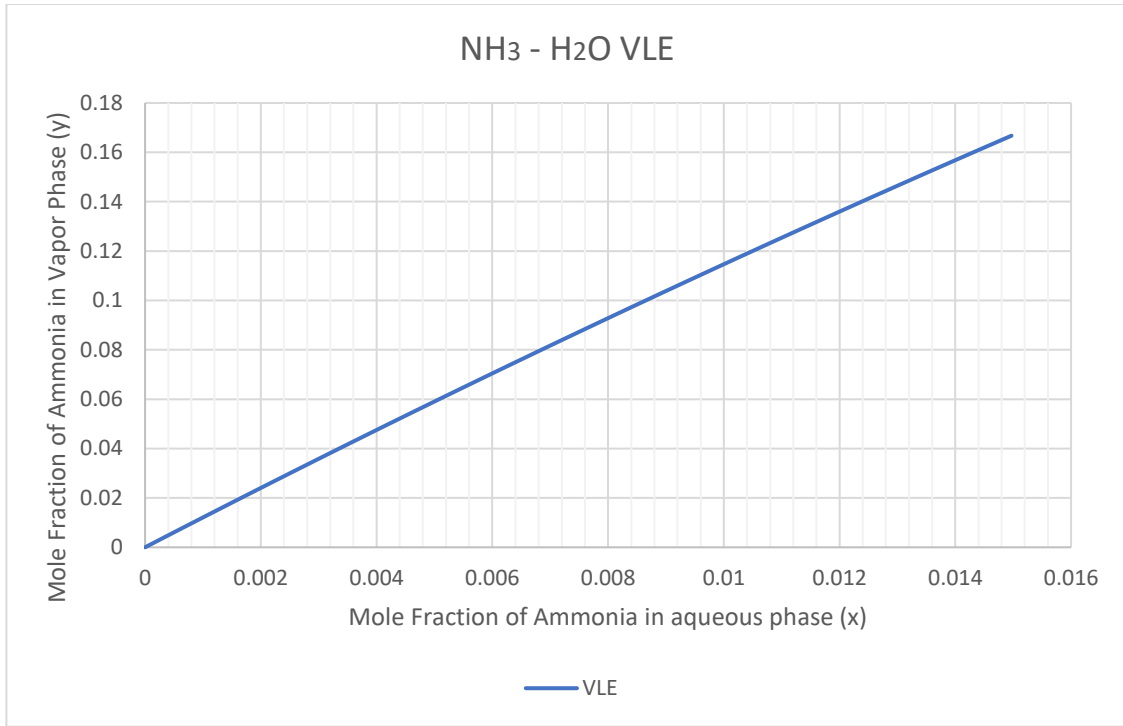


Figure 27 Ammonia Water VLE

6.1.3.2 Material Balance for Operating Line

The material balance diagram for a plate column is shown below.

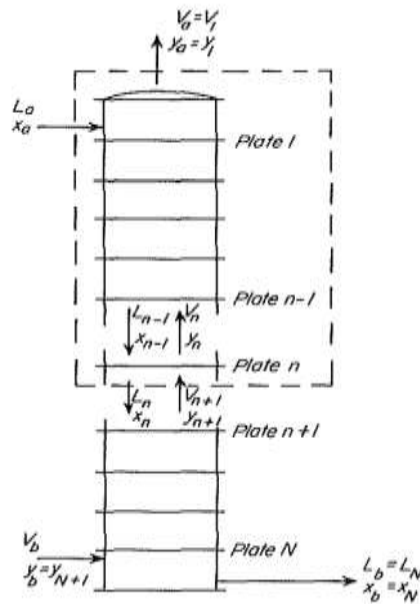


Figure 28 Material Balance Diagram for Plate Column

Our column is has another side stream named as the reflux stream that brings condensed moisture from flash vessel back to the column.

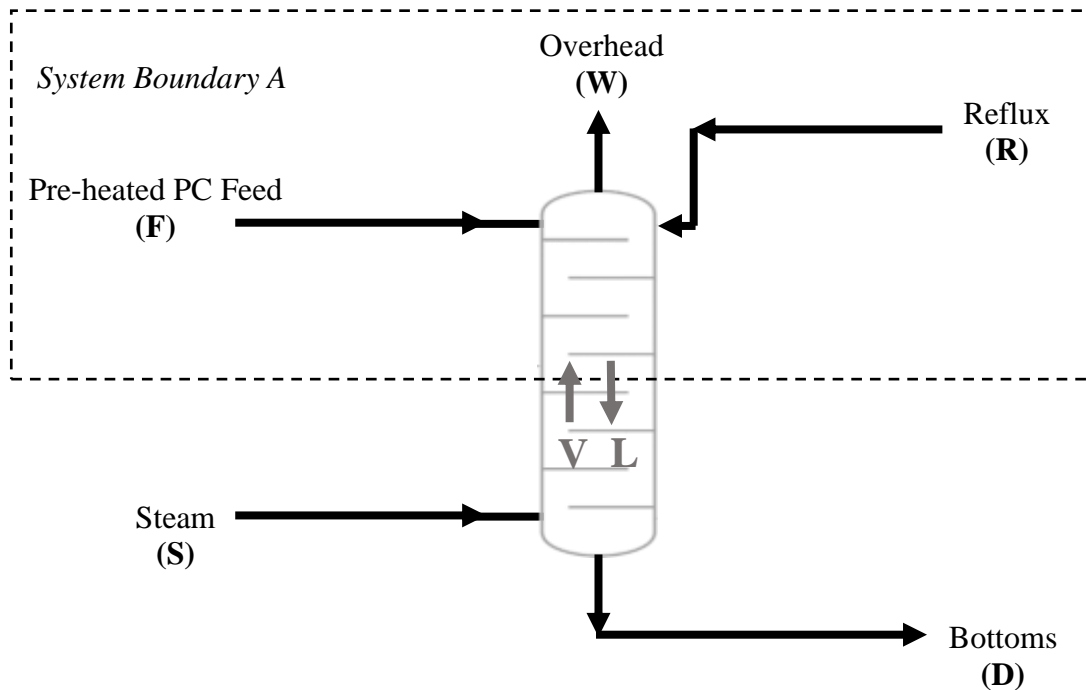


Figure 29 Material Balance on PC Stripper for Operating Line

Overall Balance on System Boundary A:

$$V + F + R = L + W$$

Ammonia Balance System Boundary A:

$$Ly + Fx_f + Rx_r = Lx + Wy_w$$

$$y = \frac{L}{V}x + \frac{Wy_w - Fx_f - Ry_r}{V} \dots \dots \dots (i)$$

Data estimated using Aspen Plus Simulation (Basis = 1hr)			
L	1234.985	F	1137
V	188.88	x_f	0.0015
D	94.19	R	3.235
x_d	0.018	x_r	0.0004

Inserting values of stream flows and compositions from Aspen data, in equation (i),

$$y = \left(\frac{1234.985}{188.88} \right) x + \frac{(94.19 \times 0.018) - (1137 \times 0.0015) - (3.235 \times 0.0004)}{188.88}$$

$$y = 6.54x - 6 \times 10^{-5} \dots \dots (ii)$$

Equation (ii) is the operating line of our stripper.

6.1.3.3 McCabe-Thiele Diagram

Since our product concentrations are very low, the steps on the McCabe Thiele diagram become very small and difficult to plot at stages near the bottom, Hence the VLE curve was replotted near the bottom on a larger scale to draw the stages. The stages one to six are drawn on the first graph and stages seven to ten are drawn on the replotted graph of bottom section.

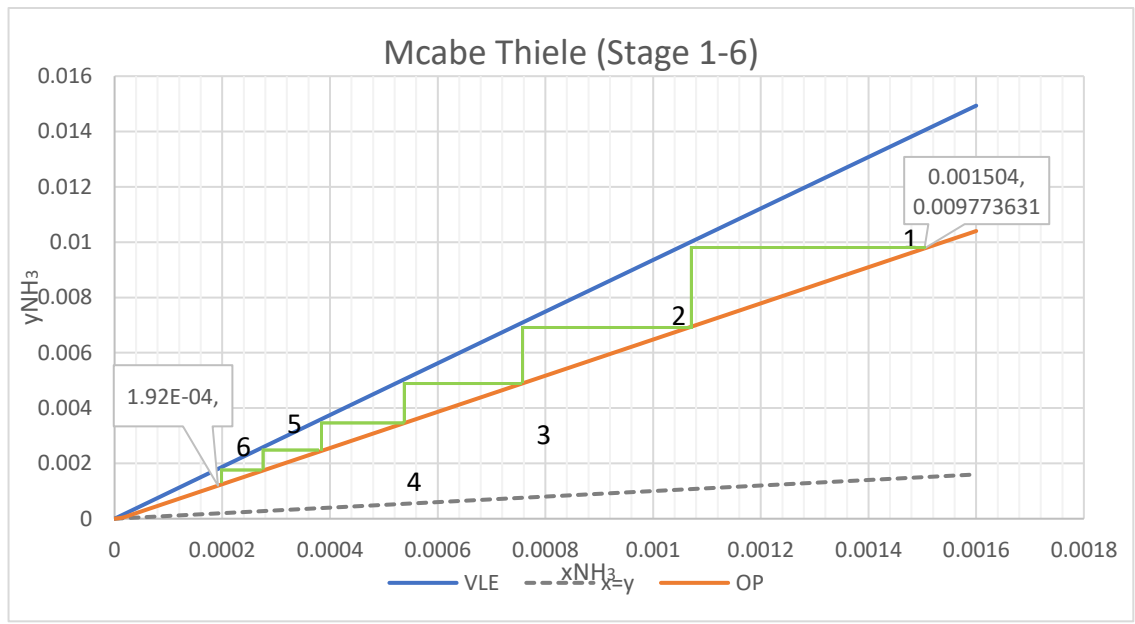


Figure 30 McCabe Thiele Diagram Stages (1-6)

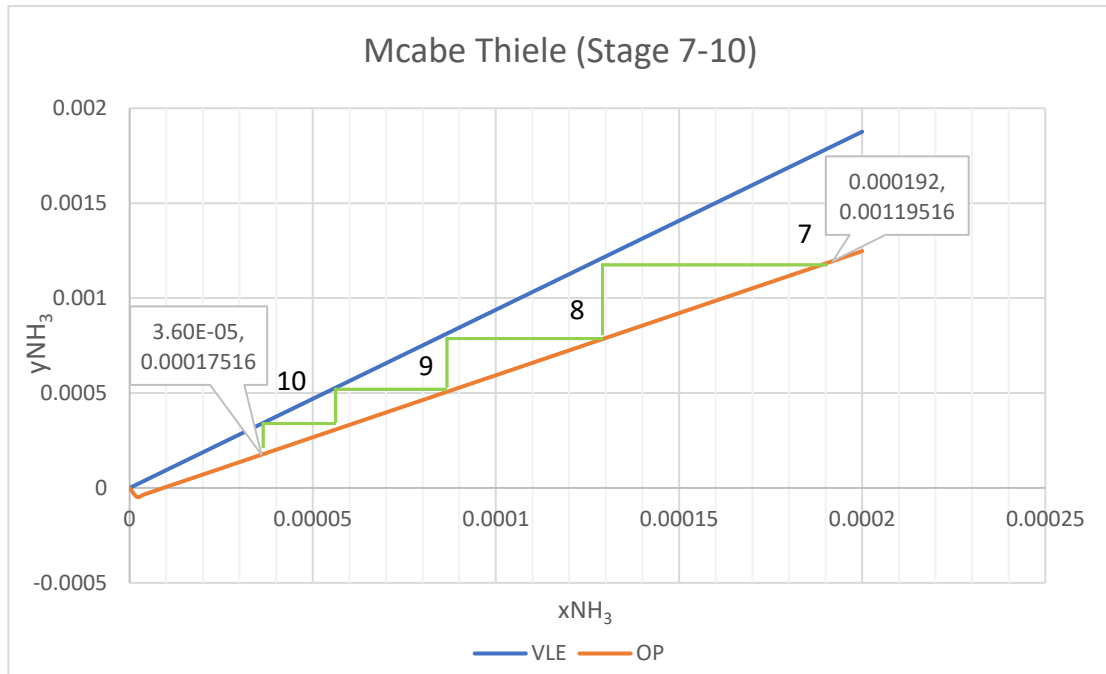


Figure 31 McCabe Thiele Diagram Stages (7-10)

The diagrams show that ten theoretical stages are required to bring ammonia concentration down from 0.0015 to 3.6×10^{-5} (1300ppm - 10 ppm).

6.1.4 Plate Efficiency

Using Van Winkle Correlation for column efficiency,

$$E_{mv} = 0.07Dg^{0.14}Sc^{0.25}Re^{0.08}$$

Where,

$$Dg = \text{surface tension number} = \left(\frac{\sigma_L}{\mu_L u_v} \right)$$

u_v = superficial vapor velocity

σ_L = liquid surface tension

μ_L = liquid viscosity

$$Sc = \text{liquid Schmidt number} = \left(\frac{\mu_L}{\rho_L D_{LK}} \right)$$

ρ_L = liquid density

D_{LK} = light key component liquid diffusivity

$$Re = \text{Reynolds number} = \left(\frac{h_w \rho_v u_v}{\mu_L} \right)$$

h_w = weir height

ρ_v = vapor density

Data		
Surface Tension Number	Dg	195.0542
liquid Surface Tension	N/m	0.05817
Liquid Viscosity	Ns/m ²	0.000272
Superficial Vapour Velocity	m/s	1.094612
Calculating Reynolds Number		
Reynold's Number	Re	146.5554
Weir height	m	5.00 × 10 ⁻²
Vapor density	kg/m ³	0.72955
Calculating Schmidt Number		
Liquid Schmidt Number	Sc	1.87E+02
Liquid density	kg/m ³	969.5
Liquid Diffusivity (LK)	m ² /s	1.50 × 10 ⁻⁹
Plate Efficiency		
Plate Efficiency	Emv	0.808

Hence the % plate efficiency is,

$$\% \text{ Plate Efficiency} = 0.808 \times 100\% = 80.8\%$$

6.1.5 No. of Real Plates

$$\text{Real no. of plates} = \frac{\text{Theoretical Stages}}{\text{Plate Efficiency}}$$

$$\text{Real no. of plates} = \frac{10}{0.808} = 12.37 = 13$$

$$\text{No. of Real Plates} = 13$$

6.2 Knockout Drum

Data			
Liquid Volumetric flowrate	0.13339m ³ /hr	Vapor Volumetric flowrate	34880 m ³ /hr
Liquid mass flowrate	292.7 lb/hr	Vapor mass flowrate	184200 lb/hr
Liquid density	61.89 lb/ft ³	Vapor density	0.1495 lb/ft ³

6.2.1 Maximum Design Vapour Velocity (U_a)

The maximum design vapor velocity formula was obtained from Perry 26-36,

$$U_a = k \sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$$

Where,

$$U_a = \text{allowable vapor velocity, } \frac{ft}{s}$$

$$\rho_L = \text{liquid density, } \frac{\text{lb}}{\text{ft}^3}$$

$$\rho_v = \text{vapor density, } \frac{\text{lb}}{\text{ft}^3}$$

$k = \text{capacity coefficient}$

The value of 'k' depends on the ratio,

$$F_{lv} = \frac{W_l}{W_v} \sqrt{\frac{\rho_l}{\rho_v}} = \frac{292.7 \frac{\text{lb}}{\text{hr}}}{184200 \frac{\text{lb}}{\text{hr}}} \sqrt{\frac{61.89 \frac{\text{lb}}{\text{ft}^3}}{0.1495 \frac{\text{lb}}{\text{ft}^3}}} = 0.0323$$

$$F_{lv} = 0.0323$$

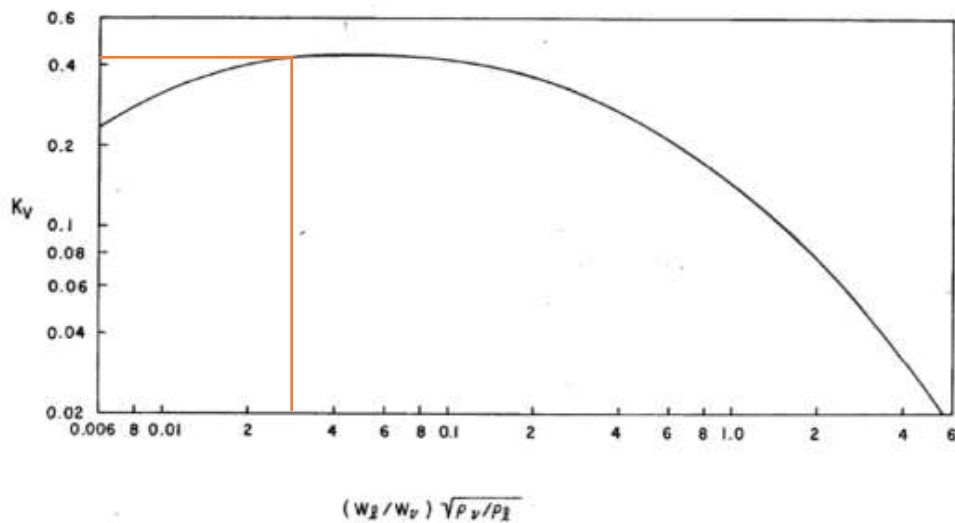


Figure 32 Capacity Coefficient for Knockout Drum

$$K=0.42$$

$$U_a = 0.42 \sqrt{\frac{61.89 - 0.1495}{0.1495}}$$

$$U_a = 8.53 \frac{\text{ft}}{\text{s}}$$

6.2.2 Drum Diameter

$$A_v = \frac{Q_v}{U_a}$$

Where,

$Q_v = \text{vapor flow rate, ft}^3/\text{s}$

A_v = vapor flow area, ft²

$$A_v = \frac{342.17}{8.53} = 40.1 \text{ ft}^2$$

$$D = \sqrt{\frac{2A_v}{0.785}}$$

$$\text{Drum Diameter} = 10.1 \text{ ft} = 3.98 \text{ m}$$

6.2.3 Drum Height

For vertical flash drums L/D should be less than 3. Using an L/D value of 2.

$$\text{Drum Height} = 2 \times 3.98 \text{ m} = 7.96 \text{ m}$$

6.3 Economizer

The economizer is a heat exchanger that we used to transfer the heat from Stripper bottoms to pre-heat our process condensate feed. A shell and tube heat exchanger is selected as it is commonly used in industries due to higher surface area to volume ratio. A pull – through floating head type tube bundle arrangement is chosen. It is easy to clean and reduces the problem of differential expansion. Some design parameters had to be selected prior to design calculations and their values were assumed after reviewing a distilled-water-raw-water shell and tube heat exchanger design (Example 7.4) in *Process Heat Transfer* by D.Q. Kern (1983).

6.3.1 Initial Design Specifications

Heat exchanger type – Shell and tube heat exchanger

Tube bundle and shell arrangement pattern – Pull through floating head type

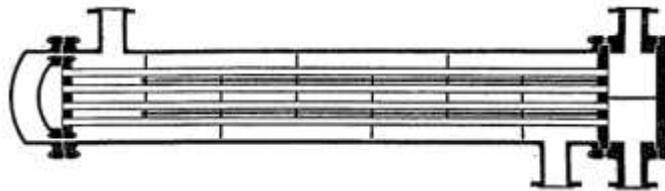


Figure 33 Pull-through Floating-Head 1-2 Heat Exchanger

Tube Side and Shell Side Fluid Arrangement and Tube Material – Usually, greater flow rate is kept on tube side to avoid greater pressure drop. However, the

flow rate of both hot fluid (20530 kg/h) and cold fluid (22850 kg/h) is comparable, so the arrangement of fluids does not need to depend on flow rates. Since stripped process condensate in stripper bottoms is low in impurities especially carbon dioxide, which is corrosive in nature, it is kept on the shell side. The process condensate feed is kept on the tube side and stainless-steel tube material is chosen to enhance the durability of our economizer.

Tube Dimensions

Number of tube passes (N)	2
Length of tube (m)	4.88 (16 ft)
Outer Diameter (OD) (m)	0.019 (0.75 in)
Tube Thickness (BWG)	18

Table 38 Specification Sheet for Heat Exchanger Tubes

For the tube ID and the flow area the following data table is referred to, using the outer diameter and BWG assumed.

Tube OD, in.	BWG	Wall thickness, in.	ID, in.	Flow area per tube, in. ²	Surface per lin ft, ft ²		Weight per lin ft, lb steel
					Outside	Inside	
3/8	12	0.109	0.282	0.0625	0.1309	0.0748	0.493
	14	0.083	0.334	0.0876		0.0874	0.403
	16	0.065	0.370	0.1076		0.0969	0.329
	18	0.049	0.402	0.127		0.1052	0.258
	20	0.035	0.430	0.145		0.1125	0.190
1/2	10	0.134	0.482	0.182	0.1963	0.1263	0.965
	11	0.120	0.510	0.204		0.1335	0.884
	12	0.109	0.532	0.223		0.1393	0.817
	13	0.095	0.560	0.247		0.1466	0.727
	14	0.083	0.584	0.268		0.1529	0.647
	15	0.072	0.606	0.289		0.1587	0.571
	16	0.065	0.620	0.302		0.1623	0.520
	17	0.058	0.634	0.314		0.1660	0.469
	18	0.049	0.652	0.334		0.1707	0.401

Figure 34 Heat Exchanger Tube Data

Internal diameter (ID) / m	0.0166
Flow area per tube / m²	0.000215

Table 39 Data from the Standard Ranges for Tube Specifications

Shell Dimensions

The following assumptions have been made for the calculation of shell dimensions:

Number of shell passes	1
Baffle cut percentage (%)	25
Baffle Spacing (m)	0.305
Shell Diameter (m)	0.387

Table 40 Shell Specifications

6.3.2 Thermo-Physical Properties

The values of thermo-physical properties for shell and tube side were retrieved from Aspen Hysys database. Since both fluids are water, the property values are similar.

Tube-side Fluid

Subject Stream: Process Condensate Feed

Properties	Symbol	Value	Units
Density of Fluid	ρ	939.5	kg/m ³
Viscosity	μ	0.0005	kg/ms
Thermal Conductivity	k	0.67	J/s m K
Specific Heat	c_p	4.3	KJ/kg K
Fouling Factor	R_d	0.0015	m ² K / W

Shell Side Fluid

Subject Stream: Stripped Process Condensate in Stripper Bottoms

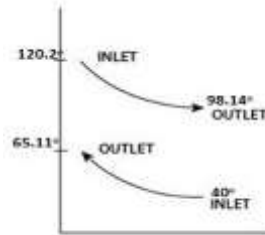
Properties	Symbol	Value	Units
Density of Fluid	ρ	939.5	kg/m ³
Viscosity	μ	0.00028	kg/ms
Thermal Conductivity	k	0.67	J/s m K
Specific Heat	c_p	4.4	KJ/kg K
Fouling Factor	R_d	0.0006	m ² K / W

6.3.3 Temperature Gradients

Hot Fluid (Shell-Side)		Cold Fluid (Tube-Side)
120.2°C	Higher Temp	40°C
98.14°C	Lower Temp	65.11°C
22.06°C	Difference	25.11°C
$(T_1 - T_2)$		$(t_2 - t_1)$

6.3.4 LMTD

$$LMTD = \frac{\Delta T_1 - \Delta T_2}{\ln \frac{\Delta T_1}{\Delta T_2}}$$



Flow Type	Counter-Current
Hot Ends Temp. Diff (ΔT_1)	55.09 K
Cold Ends Temp. Diff (ΔT_2)	58.14 K
Log Mean Temp. Diff (LMTD)	56.60 K

6.3.5 Corrected LMTD

$$R = \frac{T_{hot,in} - T_{hot,out}}{T_{cold,out} - T_{cold,in}}$$

$$R = \frac{120.2 - 98.14}{65.11 - 40} = 0.8785$$

$$S = \frac{T_{cold,out} - T_{cold,in}}{T_{hot,in} - T_{cold,in}}$$

$$S = \frac{65.11 - 40}{120.2 - 40} = 0.3131$$

Using these values of R & S, we find the value of the F_T from the graph correlation

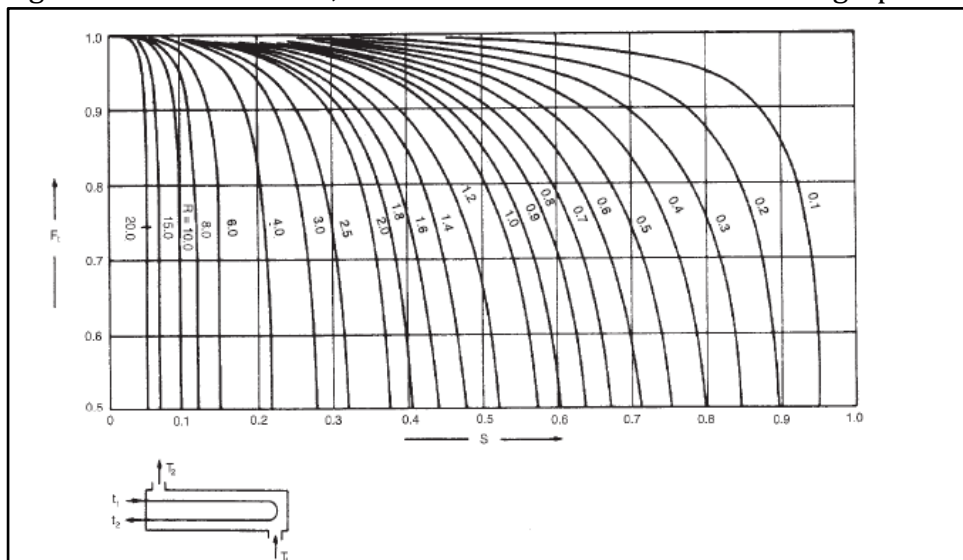


Figure 35 Temperature Correction Factor: One Shell Pass; Two Or More Even Tube Passes

$$F_T = 0.98$$

$$\Delta T_m = F_T \times LMTD = 0.98 \times 56.60 \text{ K} = 55.47 \text{ }^\circ\text{C}$$

6.3.6 Required UA

The value for the duty of the heat exchanger was already calculated and quoted in the energy balance chapter, and it was as follows.

Duty (Q) (kJ / hr)	2,217,906.49
ΔT_m ($^\circ\text{C}$)	55.47

Table 41 Data for UA calculation

And now, using the equation for basic heat transfer, we will be finding out the UA required in order for to design the heat exchanger.

$$Q = UA\Delta T_m$$

$$UA = Q/T = 2,217,906.49/55.47 = 39983.9 = 11106.64 \text{ W}/^\circ\text{C}$$

This is the value of the required UA, one of our design parameters is attaining a value of UA that is greater than or equal to this specific value.

Allowable range of U (W/m ² °C)	800 - 1500
Assumed value of U (W/m ² °C)	1200

Table 42 Estimated ed Range for Value of Overall Heat Transfer Coefficient U

6.3.7 No. of Tubes & Tube Pitch

For the number of tubes, we have to initially assume a value for U (the overall heat transfer coefficient), which can be found out from literature. The range of U found from literature allowed a range from 800 - 1500 W/m²°C when dealing with gases on both inlet and outlet streams. Hence, with the help of hit and trial, we were able to assume a value, which was most suitable for our design.

Using the value of UA as 11106.64 as mentioned above, the required area is calculated as follows:

$$A = \frac{UA}{U} = \frac{11106.64}{1200} = 9.07 \text{ m}^2$$

For the tube dimensions mentioned in initial design specifications, the area is calculated using the following formula:

$$\text{Area per tube} = \pi DL = \pi \times 0.019 \times 4.88 = 0.291 \text{ m}^2$$

The number of tubes required for this particular area would then be,

$$\text{Number of tubes} = \frac{\text{Total area required}}{\text{Area of one tube}} = \frac{9.07}{0.291} = 31.82 \approx 32 \text{ tubes}$$

The pitch chosen for this type of heat exchanger is triangular because of the higher rates of heat transfer provided by a triangular pitch along with the ease with which it can be cleaned. The tube pitch would be in accordance with the KERN method would be,

$$p_t = 1.25d_o = 1.25 \times 0.019 = 0.0238 \text{ m}$$

In addition, the tube clearance would then be calculated using the following formula,

$$\text{Clearance} = p_t - d_o = 0.0238 - 0.019 = 0.0048 \text{ m}$$

The tube dimensions so far are summarized in the table below.

Tube Specification Summary	
Specification Title	Specification Value
Number of passes (N_p)	2
Tube length (L)	4.88 m
Outer Diameter (d_o)	0.019 m
Inner Diameter (d_i)	0.0166 m
BWG	18
Number of tubes (N_t)	32
Flow area per tube	0.000215 m ²
Surface area per tube	0.291 m ²
Tube pitch (p_t)	Triangular & 0.0238 m
Tube clearance	0.0048 m

Table 43 Tube Specification Summary

6.3.8 Shell Diameter

$$D_H = 4 \frac{\left(\frac{P_t}{2} \times 0.86P_t - \frac{\pi d_o^2}{8} \right)}{\frac{\pi d_o}{2}} \text{ For Triangular Pitch}$$

$$D_H = 0.014 \text{ m}$$

6.3.9 Mass Velocities

Shell-Side

$$G_s = \frac{m_s}{a_s}$$

$$G_s = 199.81 \frac{\text{kg}}{\text{m}^2\text{s}}$$

Tube-Side

$$G_T = \frac{m_T}{a_T}$$

$$G_T = 1654.8 \frac{\text{kg}}{\text{m}^2 \text{s}}$$

6.3.10 Linear Velocity of Tube Side

$$U_T = \frac{q_T}{a_T}$$

$$U_T = 1.76 \text{ m/s}$$

6.3.11 Flow Areas

$$a_T = n \frac{\pi}{4} d_i^2 \text{ Tube Side}$$

Flow Area of Tube [inside]	a_T	0.00344645	m^2
Linear Velocity of Tube [inside]	u_T	1.76143792	m/s
Mass Velocity	G_T	1654.87093	$\text{kg/m}^2 \text{s}$

$$a_s = \frac{c_T B D_s}{P_t} \text{ Shell Side}$$

Flow Area	a_s	0.03175248	m^2
Mass Velocity	G_s	199.80944	$\text{kg/m}^2 \text{s}$

6.3.12 Reynolds's Number

Shell-Side

$$Re_s = \frac{D_H G_s}{\mu_s}$$

$$Re_s = 9640 \text{ Turbulent Flow}$$

Tube-Side

$$Re_T = \frac{\rho_T u_T d_T}{\mu_T}$$

$$Re_T = 54812 \text{ Turbulent Flow}$$

6.3.13 Prandtl Number

Shell-Side

$$Pr_s = \frac{C_{p,s}\mu_s}{k_s}$$

$$Pr_s = 1.839$$

Tube-Side

$$Pr_T = \frac{C_{p,T}\mu_T}{k_T}$$

$$Pr_T = 3.209$$

6.3.14 Nusselt's Number

$$Nu_s = 0.36Re_s^{0.55}Pr_s^{0.33} \quad .$$

Shell-Side

$$Nu_s = 68.3686$$

Tube-Side

$$Nu_T = 0.027Re_T^{0.8}Pr_T^{0.3}$$

$$Nu_T = 236.7964$$

6.3.15 Shell Side HT Co-efficient

$$h_{o,s} = Nu_s \frac{k_s}{D_H}$$

$$k_s = \text{thermal conductivity of shell side fluid} = 0.67$$

$$Nu_s = 68.3686$$

$$h_{o,s} = 3390.915803 \text{ W / m}^2 \text{ K}$$

6.3.16 Tube Side HT Co-efficient

$$h_{i,T} = Nu_T \frac{k_T}{d_i} \quad \text{Tube Side}$$

$$k_s = \text{thermal conductivity of tube side fluid} = 0.67$$

$$Nu_T = 236.7964$$

$$h_{i,T} = 9580.070143 \text{ W / m}^2 \text{ K}$$

6.3.17 Overall Heat Transfer coefficient

$$U_{do} = \frac{1}{\frac{1}{h_{o,s}} + R_{do,s} + \frac{d_o}{d_i} R_{do,T} + \frac{d_o}{d_i} \frac{1}{h_{i,T}}}$$

$$U_{do} = 364.905 \text{ W / m}^2 \text{ K}$$

6.3.18 Pressure Drop Calculations

$$\Delta P_s = \frac{f_s G_s^2 D_s (N_B + 1)}{2g\rho_s D_H} \text{ Shell Side}$$

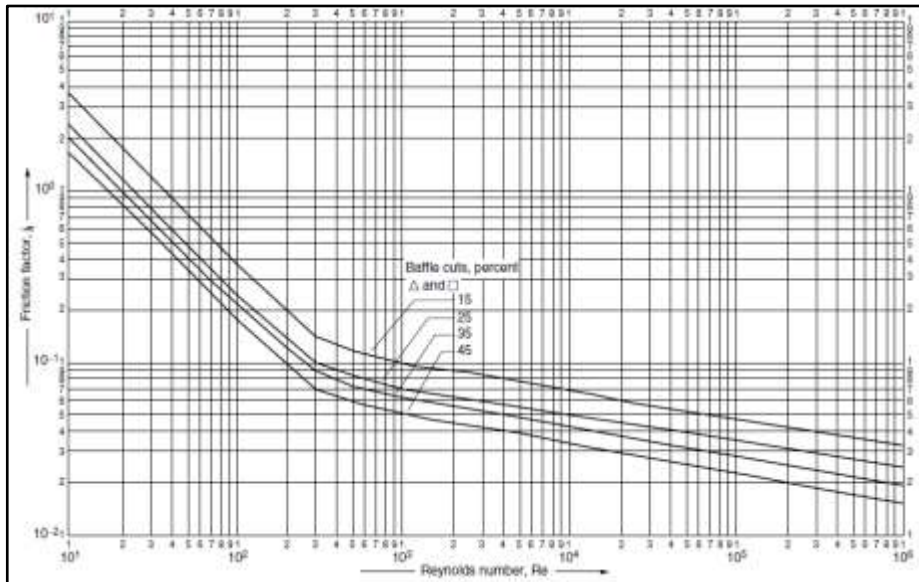


Figure 36 Shell Side Pressure Drop Graph

Friction Factor (use graph)	f_s	0.0023	From graph
Number of Baffles	N_b	15.003937	
Shell Side Pressure Drop	ΔP_s	2.2886469	kg / m²

Table 44 Shell-side Pressure Drop

$$\Delta P_T = \frac{f_T G_T^2 L N_P}{2g\rho_T d_i} \text{ Tube Side}$$

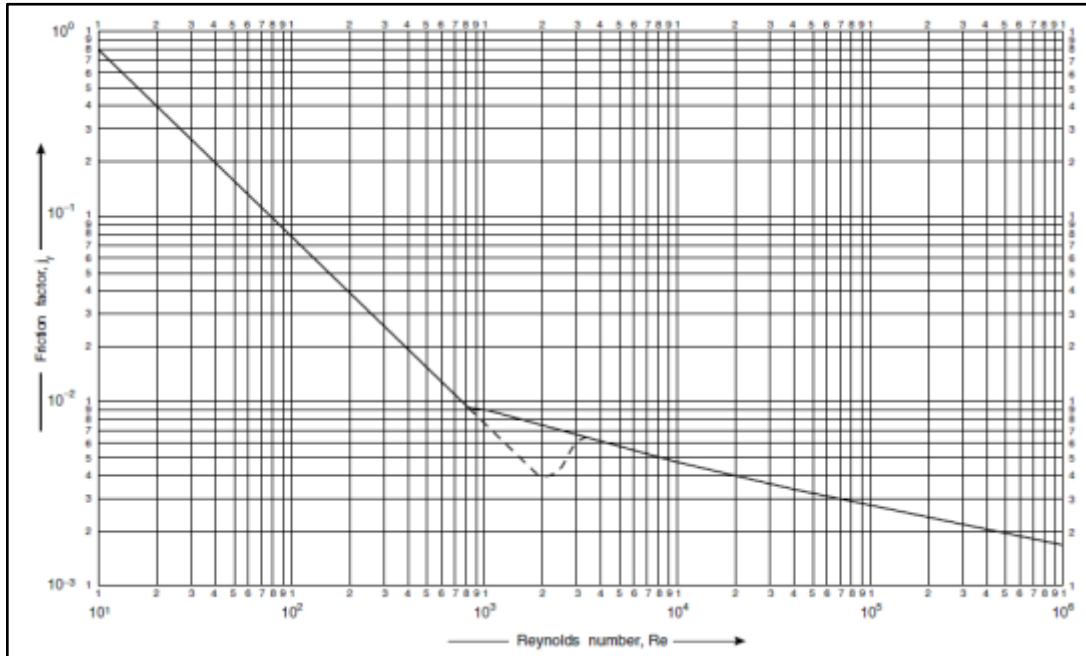


Figure 37 Tube Side Pressure Drop Graph

Friction Factor (use graph)	f_T	0.00018	From graph
Pressure Drop	ΔP_T	15.7702429	kg / m ²
Return Loss	ΔP_r	1189.77649	kg / m ²
Tube Side Pressure Drop	ΔP_T	1205.5467	kg / m²

Table 45 Tube Side Pressure Drop

In order to calculate the shell inside diameter (D_i), there are some parameters that need to be calculated namely the shell clearance and the bundle diameter.

SIMULATION

The simulation model for our plant was developed for the purpose of reaching the most optimum conditions that ensured high yield of our final.

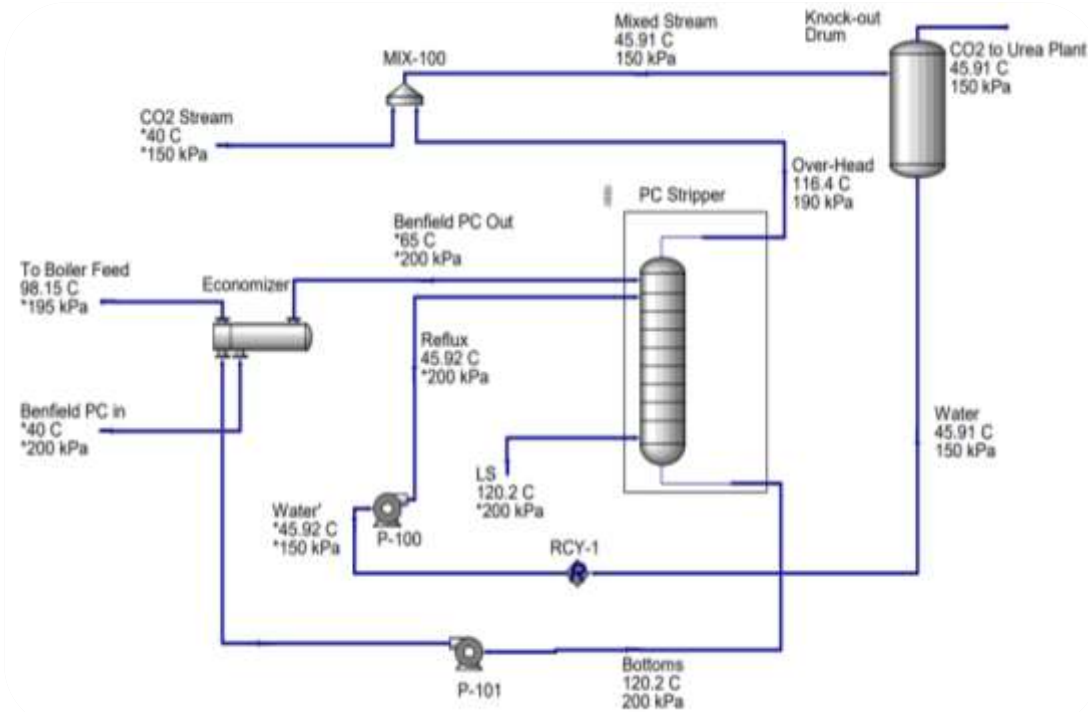


Figure 38 Benfield PC Stripping System Simulation Model

7.1 General Process Description

- A steam stripping system is designed operating at high temperature (120 °C) and low pressure (2 bar).
- We used economizer to preheat Benfield PC by heat recovery from bottoms.
- Solubility of NH₃ (180g/100g water at 25 °C) is considerably greater than CO₂ (1.4g/100g) hence requiring extensive separation.
- Heated Benfield PC enters the top stage of stripper with 12 actual stages of 1.06m diameter sieve type trays.
- Purified PC leaves at bottom and can be used as boiler feed water.

- Vapor overhead is not sent to atmosphere as it contains NH3 and CO2.
- It is combined with CO2 recovered from syngas being sent to Urea section. After mixing, the CO2 stream is sent to a knock-out drum and separated moisture is refluxed back to the steam stripper.

7.2 Components

The components were extracted from the Aspen Hysys. Following is the list of components that were used in our simulation model.

Component	Type	Group
H2O	Pure Component	
CO2	Pure Component	
Ammonia	Pure Component	
Argon	Pure Component	
CO	Pure Component	
Hydrogen	Pure Component	
Nitrogen	Pure Component	
Methane	Pure Component	

Figure 39 Simulation – Component List

7.3 Fluid Package

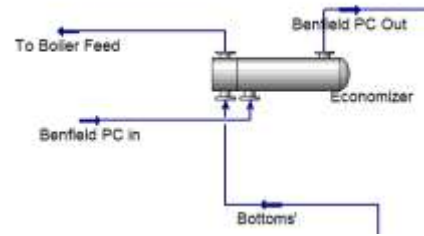
Fluid Package	Component List	Property Package
Basis-1	Component List - 1 [HYSYS Datab	Sour PR

Figure 40 Simulation - Fluid Package

7.4 Process Equipment

7.4.1 Economizer

It is the main equipment in simulation. To heat the feed stream at specific temperature, we use shell and tube heat exchanger as economizer. Hot bottom stream of stripper passes from the shell side and transfer its heat to Feed stream and raises its temperature up to 65 C°.



Design	Rating	Worksheet	Performance	Dynamics	Rigorous Shell&Tube	
Worksheet		Name	Bottoms'	To Boiler Feed	Benfield PC in	Benfield PC Out
Conditions		Vapour	0.0000	0.0000	0.0000	0.0000
Properties		Temperature [C]	120.2	98.25	40.00	65.00
Composition		Pressure [kPa]	200.0	195.0	200.0	200.0
PF Specs		Molar Flow [kgmole/h]	1268	1268	1137	1137
		Mass Flow [kg/h]	2.285e+004	2.285e+004	2.053e+004	2.053e+004
		Std Ideal Liq Vol Flow [m3/h]	22.90	22.90	20.61	20.61
		Molar Enthalpy [kJ/kgmole]	-2.787e+005	-2.805e+005	-2.849e+005	-2.830e+005
		Molar Entropy [kJ/kgmole-C]	75.44	70.87	57.78	63.77
		Heat Flow [kJ/h]	-3.536e+008	-3.558e+008	-3.240e+008	-3.218e+008

Figure 41 Simulation – Economizer Worksheet

Design	Rating	Worksheet	Performance	Dynamics	Rigorous Shell&Tube
Performance		Overall Performance			
Details		Duty	2.218e+06 kJ/h		
Plots		Heat Leak	0.000e-01 kJ/h		
Tables		Heat Loss	0.000e-01 kJ/h		
Setup		UA	4.03e+04 kJ/C-h		
Error Msg		Min. Approach	55.201 C		
		LMTD	55.06 C		
		Detailed Performance			
		UA Curvature Error	0.0000 kJ/C-h		
		Hot Pinch Temp	120.2013 C		
		Cold Pinch Temp	65.0000 C		
		Ft Factor	0.971		
		Uncorrected LMTD	56.713 C		

Figure 42 Simulation – Economizer Performance Sheet

7.4.2 Process Condensate Stripper

The stripper is a sieve trayed column with 13 stages. The heated process condensate feed enters the stripper at top stage and low-pressure steam enters at bottom stage. Due to heat and mass transfer between the two streams, carbon dioxide and ammonia is transferred into the vapor stream. The stripped process condensate leaves stripper in bottoms and is taken to boiler feed water after heat recovery in economizer.

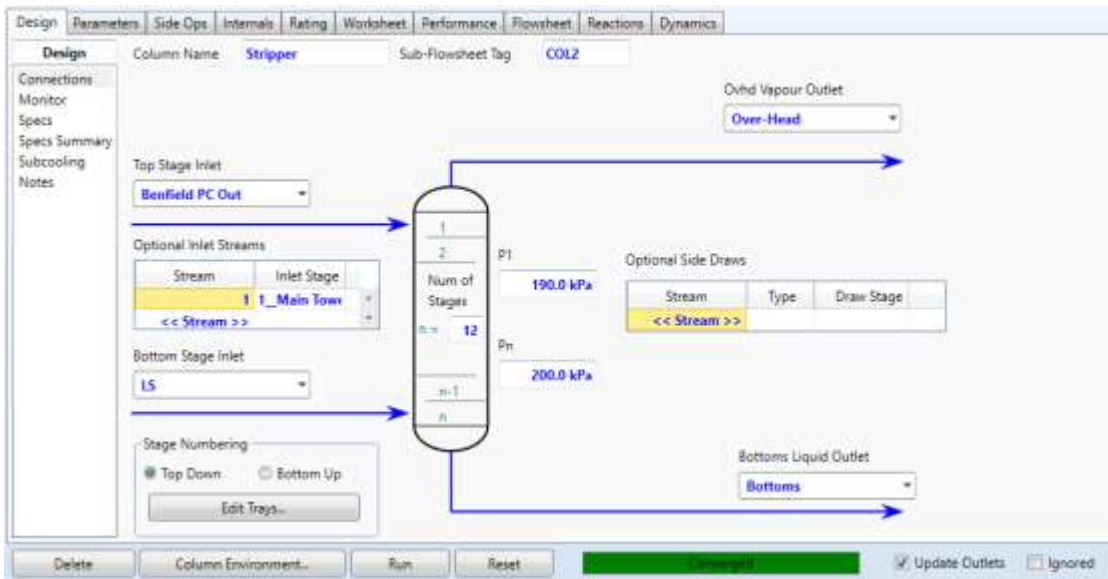
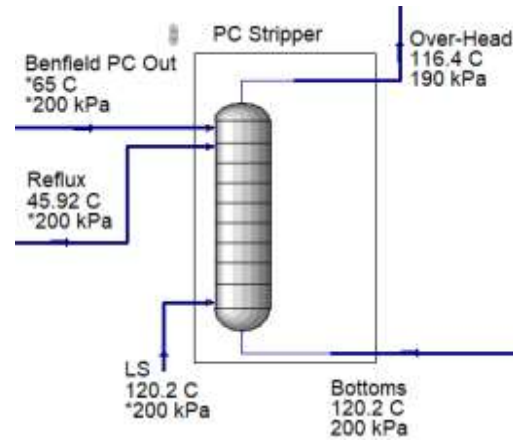


Figure 43 Simulation - Stripper Design Layout

		WASTE FEED-2 @COL2	1 @COL2	STEAM @COL2	VAPOUR IN @COL2	CONDENSATE @COL2
Conditions	Name					
Properties	Vapour	0.0000	0.0000	1.0000	1.0000	0.0000
Compositions	Temperature [C]	65.00	46.56	120.2	116.2	120.2
PF Specs	Pressure [kPa]	200.0	200.0	200.0	190.0	200.0
	Molar Flow [kgmole/h]	1137	8.864	172.1	49.63	1268
	Mass Flow [kg/h]	2.053e+004	163.6	3100	943.6	2.285e+004
	Std Ideal Liq Vol Flow [m3/h]	20.61	0.1670	3.106	0.9832	22.90
	Molar Enthalpy [kJ/kgmole]	-2.830e+005	-2.827e+005	-2.387e+005	-2.376e+005	-2.787e+005
	Molar Entropy [kJ/kgmole-C]	63.77	61.45	177.2	179.9	75.44
	Heat Flow [kJ/h]	-3.218e+008	-2.506e+006	-4.108e+007	-1.179e+007	-3.536e+008

Update Outlets

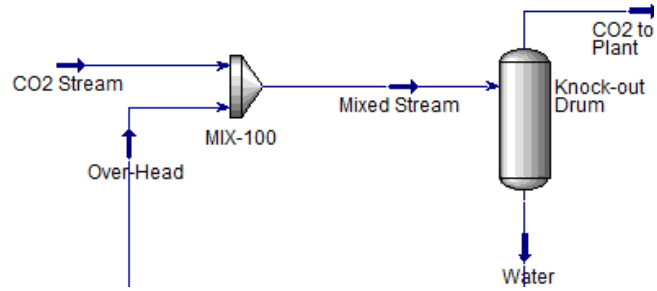
Figure 44 Simulation - Stripper Worksheet

		WASTE FEED-2	1	STEAM	VAPOUR IN	CONDENSATE
Conditions	H2O	0.9969	0.9641	1.0000	0.9230	1.0000
Properties	CO2	0.0016	0.0176	0.0000	0.0398	0.0000
Compositions	Ammonia	0.0015	0.0183	0.0000	0.0372	0.0000
PF Specs	Argon	0.0000	0.0000	0.0000	0.0000	0.0000
	CO	0.0000	0.0000	0.0000	0.0000	0.0000
	Hydrogen	0.0000	0.0000	0.0000	0.0000	0.0000
	Nitrogen	0.0000	0.0000	0.0000	0.0000	0.0000
	Methane	0.0000	0.0000	0.0000	0.0000	0.0000

Figure 45 Simulation - Stripper Streams Composition Sheet

7.4.3 Mixer and Knock-Out Drum

Vapor overhead from process condensate stripper contains carbon dioxide and ammonia, hence it is not environmentally friendly to discard it into the atmosphere. In our process we integrated this stream back into



the process by combining it with the carbon dioxide stream that is removed from syngas and is sent to the urea plant. After mixing both streams, excess moisture is removed from the carbon dioxide stream using the knock-out drum.

Mixer: MIX-100

Design Rating Worksheet Dynamics

Worksheet

	Over-Head	CO2 Stream	Mixed Stream
Conditions			
Properties			
Composition			
PF Specs			
H2O	0.9230	0.0490	0.0708
CO2	0.0398	0.9422	0.9198
Ammonia	0.0372	0.0000	0.0009
Argon	0.0000	0.0019	0.0019
CO	0.0000	0.0000	0.0000
Hydrogen	0.0000	0.0039	0.0038
Nitrogen	0.0000	0.0030	0.0029
Methane	0.0000	0.0000	0.0000

Delete OK Ignored

Figure 46 Simulation - Mixer Stream Compositions

Mixer: MIX-100

Design Rating Worksheet Dynamics

Worksheet

	Name	Over-Head	CO2 Stream	Mixed Stream
Conditions	Vapour	1.0000	1.0000	0.9956
Properties	Temperature [C]	116.2	40.00	46.55
Composition	Pressure [kPa]	190.0	150.0	150.0
PF Specs	Molar Flow [kgmole/h]	49.63	1946	1996
	Mass Flow [kg/h]	943.6	8.274e+004	8.369e+004
	Std Ideal Liq Vol Flow [m3/h]	0.9832	100.0	101.0
	Molar Enthalpy [kJ/kgmole]	-2.376e+005	-3.824e+005	-3.788e+005
	Molar Entropy [kJ/kgmole-C]	179.9	172.8	173.6
	Heat Flow [kJ/h]	-1.179e+007	-7.442e+008	-7.559e+008

Figure 47 Simulation - Mixer Worksheet

Separator: Knock-out Drum

Design Reactions Rating Worksheet Dynamics

Worksheet

		Mixed Stream	Water	CO2 to Urea Plan
Conditions	H2O	0.0708	0.9640	0.0668
Properties	CO2	0.9198	0.0176	0.9238
Composition	Ammonia	0.0009	0.0184	0.0008
PF Specs	Argon	0.0019	0.0000	0.0019
	CO	0.0000	0.0000	0.0000
	Hydrogen	0.0038	0.0000	0.0038
	Nitrogen	0.0029	0.0000	0.0029
	Methane	0.0000	0.0000	0.0000

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Figure 48 Simulation - Knock-out Drum Stream Compositions

Separator: Knock-out Drum

Reactions Rating Worksheet Dynamics

Worksheet	Name	Mixed Stream	Water	CO2 to Urea Plan
Compositions	Vapour	0.9956	0.0000	1.0000
Properties	Temperature [C]	46.55	46.55	46.55
Position	Pressure [kPa]	150.0	150.0	150.0
Specs	Molar Flow [kgmole/h]	1996	8.866	1987
	Mass Flow [kg/h]	8.369e+004	163.6	8.352e+004
	Std Ideal Liq Vol Flow [m3/h]	101.0	0.1671	100.8
	Molar Enthalpy [kJ/kgmole]	-3.788e+005	-2.827e+005	-3.792e+005
	Molar Entropy [kJ/kgmole-C]	173.6	61.46	174.1
	Heat Flow [kJ/h]	-7.559e+008	-2.506e+006	-7.534e+008

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Figure 49 Simulation - Knock-out Drum Worksheet

OPTIMIZATION

8.1 Optimizing Steam Requirement

Steam is our main utility and principal operating cost required for the treatment of Benfield process condensate. It is important to optimize steam usage to make our process cost-effective. We performed two parametric analyses for steam flow rate optimization.

8.1.1 Case Study of Stripper Bottoms vs. Steam Flow Rate

The separation of ammonia from Benfield condensate requires mass and heat transfer through steam. The steam increases the temperature of our PC feed and provides a medium for mass transfer. In our analysis we varied the steam flow rate from 1800 to 10,000

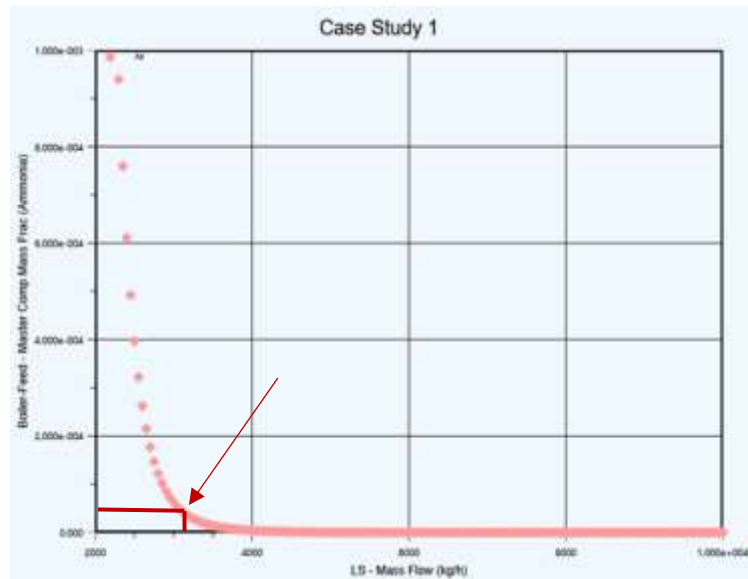


Figure 50 Case Study of Stripper Bottoms vs. Steam Flow Rate

kg/h. Since an ammonia concentration of less than 10 ppm is desired in stripped condensate as per BFW standards by ABMA, we chose a value of 3100 kg/hr which corresponds to our required purity. The case study simulation results are shown below, and operating point is marked.

Optimized Steam Flow Rate = 3100 kg/hr

Ammonia mass fraction in stripped PC = 0.5 ppm

8.1.2 Case Study of Moisture Content in Vapor Overhead vs. Steam Flow Rate

As steam flow rate is increased, the heat transfer to process condensate feed increases which results in vaporization of water along with the gases. Hence moisture entrainment on the vapor overhead is increased. On one hand increasing steam flow rate is necessary for good ammonia removal, whereas on the other hand it reduces our yield of BFW by

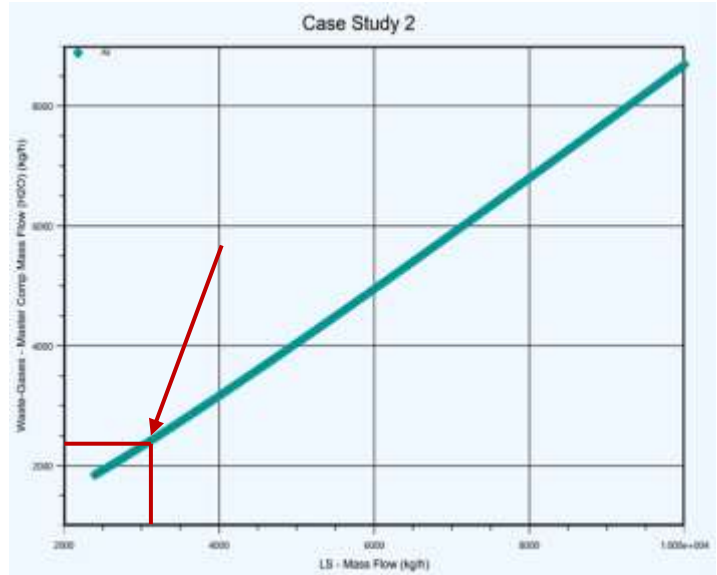


Figure 51 Case Study of Moisture Content in Vapor Overhead vs. Steam Flow Rate

causing greater water entrainment into the vapor overhead of stripper. Hence it is important to monitor the moisture entrainment with increase in steam flow rate before selecting an optimized steam flow rate. The case study shows that as steam flow rate increases, moisture entrainment increases linearly. The flow rate of water in vapor overhead at 3100 kg/hr LS flow rate as given by the case study is 2400 kg/hr which is almost 0.1% of stripped process condensate. The moisture entrainment obtained when the simulation was run on Sour PR fluid package is even less with a mass flow rate of water of 825 kg/hr.

Optimized Steam Flow Rate = 3100 kg/hr

Moisture Entrained in Vapor Overhead = 825 kg/h

8.2 Optimizing Feed Temperature

The temperature of steam plays a vital role in the efficiency of our process condensate stripper. The Henry's constant significantly increases with temperature as shown in the curve below. Hence it is favorable to enter the process condensate feed at a higher temperature to reduce steam requirement and facilitate ammonia separation. For this purpose, we performed a parametric analysis on Aspen Hysys by studying the change in ammonia concentration in stripper bottoms by varying process condensate feed temperature. The steam flow rate was set at its optimized value of 3100 kg/hr.

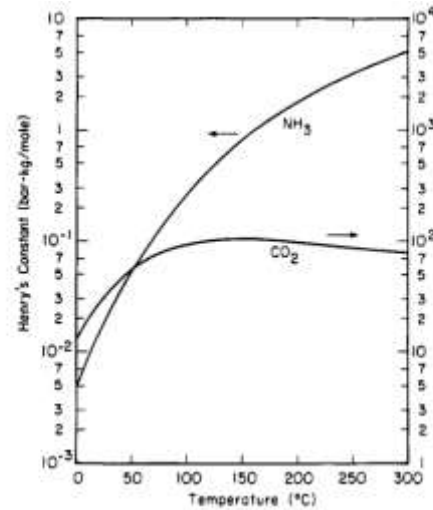


Figure 532 Henry's Constants for Ammonia and Carbon Dioxide

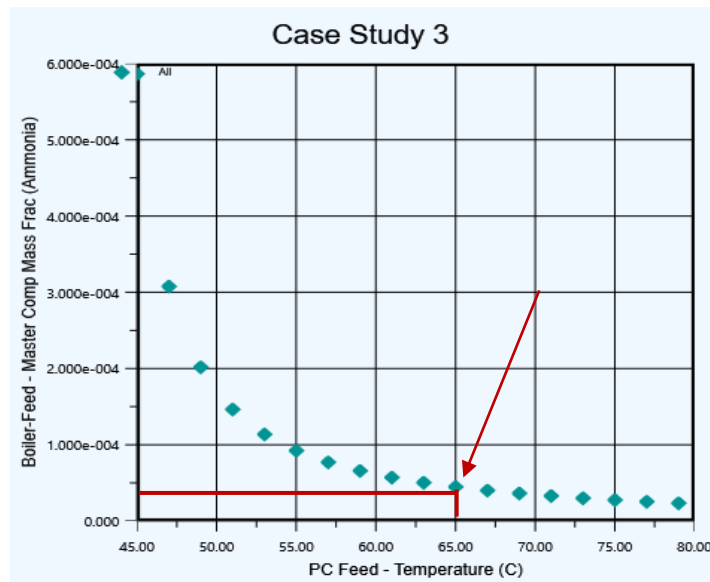


Figure 523 Case Study of Stripper Bottoms vs. PC Feed Temperature

Optimized PC Feed Temperature = 65°C

Ammonia mass fraction in stripped PC = 0.75 ppm

CHAPTER 9

ECONOMIC ANALYSIS

Cost estimation and economic analysis are the most important yet complicated aspects of setting up a plant. Industrial plants, like the one being discussed in our thesis are designed to produce products on a very large scale. To maximize the profits generated, the plant investment needs to be minimized. Hence, costing is of utmost importance especially because it is a crucial part of the feasibility analysis, which helps in the selection of the manufacturing process to be used.

The estimation of design cost is the fiscal description of the different costs related to the plant, based on total investment. Working capital and fixed capital make up the total investment. Fixed capital is the overall cost required for erection of a plant that is functional and ready to be set up. This cost is paid for the installation of equipment and will eventually wind up as the salvage value. Whereas, working capital is the additional funds that would be needed to ensure that the plant is up and running. All these values, along with the detailed economic analysis are shown in the tables given below.

9.1 Plant Equipment Cost (PCE)

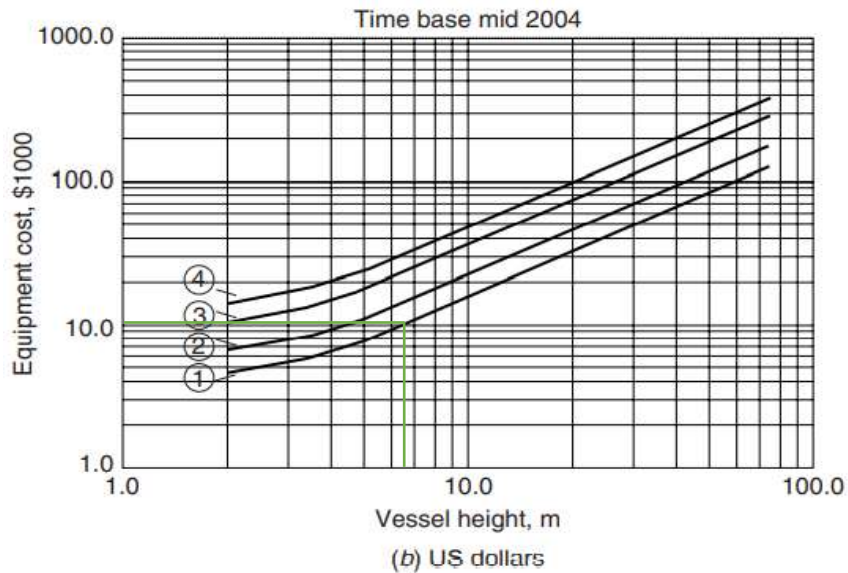
9.1.1 Process Condensate Stripper

Vessel Height = 6.5 m

Material = Stainless steel since our PC feed stream is corrosive to some extent

Pressure = 2 bar

$$\text{Vessel cost} = 10 \times 2.0 \times 1.0 = \$20,000$$



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

Temperature up to 300°C

Figure 54 Vertical Pressure Vessels Costing Data

No. of trays = 12

Type of trays = sieve type

Plate Diameter = 1.067 m

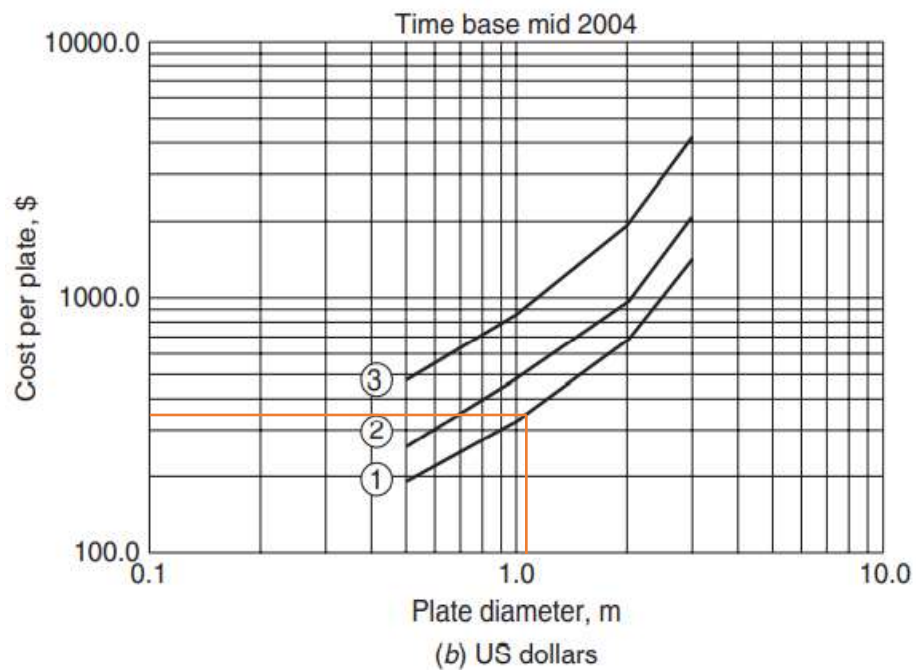
Plate Material = Stainless Steel

$$\text{Cost of trays} = 12 \times 360 \times 1.7 = \$7344$$

$$\text{Total cost of Stripper} = \text{vessel cost} + \text{cost of plates}$$

$$= \$20,000 + \$7344 = \$27,344$$

$$\text{Total cost of Stripper} = \$27,344$$



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

Figure 55 Column Plates Costing Data

9.1.2 Knockout Drum

$$\text{Drum Diameter} = 3.98\text{m}$$

$$\text{Drum Height} = 7.96\text{ m}$$

$$\text{Capacity} = \pi \frac{D^2}{4} \times h = 3.14 \times \frac{3.98^2}{4} \times 7.96 = 99\text{m}^3$$

Drum material = carbon steel

$$\text{Purchased equipment cost} = C_e = CS^n$$

Where C_e is the purchased equipment cost in \$,

C is a constant in \$,

S is size unit, and

n is the index.

From table 6.2 in Coulson Richardson's Chemical Engineering Series Volume 6,

$$C = \$2900$$

$$n = 0.6$$

$$S = \text{drum capacity in } \text{m}^3$$

$$= 99\text{m}^3$$

$$C_e = \$2900 \times 99^{0.6}$$

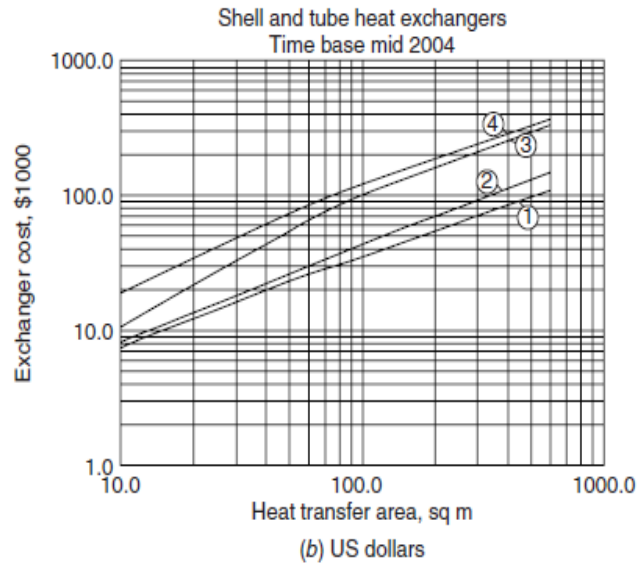
$$C_e = \$45685$$

Equipment	Size unit, S	Size range	Constant C, £	Constant C, \$	Index n	Comment
Agitators						
Propelled	driver	5-75	1200	1900	0.5	
Turbine	power, kW		1800	3000	0.5	
Boilers						
Packaged	kg/h steam	(5-50) × 10 ³	70	120	0.8	oil or gas fired
up to 10 bar			60	100	0.8	
10 to 60 bar						
Centrifuges						
Horizontal basket	dia., m	0.5-1.0	35,000	58,000	1.3	carbon steel
Vertical basket			35,000	58,000	1.0	× 1.7 for ss
Compressors						
Centrifugal	driver power, kW	20-500	1160	1920	0.8	electric, max. press. 50 bar
Reciprocating			1600	2700	0.8	
Conveyors						
Belt	length, m	2-40				
0.5 m wide			1200	1900	0.75	
1.0 m wide			1800	2900	0.75	
Crushers						
Cone	t/h	20-200	2300	3800	0.85	
Pulvisers	kg/h		2000	3400	0.35	
Dryers						
Rotary	area, m ²	5-30	21,000	35,000	0.45	direct
Pan		2-10	4700	7700	0.35	gas fired
Evaporators						
Vertical tube	area, m ²	10-100	12,000	20,000	0.53	carbon steel
Falling film			6500	10,000	0.52	
Filters						
Plate and frame	area, m ²	5-50	5400	8800	0.6	cast iron
Vacuum drum		1-10	21,000	34,000	0.6	carbon steel
Furnaces						
Process						
Cylindrical	heat abs, kW	10 ³ -10 ⁴	330	540	0.77	carbon steel
Box		10 ³ -10 ⁵	340	560	0.77	× 2.0 ss
Reactors						
Jacketed, agitated	capacity, m ³	3-30	9300	15,000	0.40	carbon steel
			18,500	31,000	0.45	glass lined
Tanks						
Process	capacity, m ³					
vertical		1-50	1450	2400	0.6	atmos. press.
horizontal		10-100	1750	2900	0.6	carbon steel
Storage						
floating roof		50-8000	2500	4350	0.55	× 2 for
cone roof		50-8000	1400	2300	0.55	stainless

Figure 56 Equipment Cost Index

9.1.3 Economiser

Economizer Specification			
Total Heat transfer Area	9m ²	Type	Floating Head
Pressure Factor	1.0	Material	CS (Shell): SS (tubes)



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 6.3a, b. Shell and tube heat exchangers. Time base mid-2004
Purchased cost = (bare cost from figure) × Type factor × Pressure factor

Figure 57 Shell and Tube HE Costing Data

Total Purchased Cost	10000\$
-----------------------------	----------------

9.1.4 Reflux Pump

$$\text{Purchased equipment cost} = C_e = CS^n$$

From table 6.2 in Coulson Richardson's Chemical Engineering Series Volume 6,

$$C = \$1920$$

$$n = 0.8$$

$$S = \text{driver power in kW} = 0.0025 \text{ kW}$$

$$C_e = \$1920 \times 0.0025^{0.8} = \$15.9$$

9.1.5 Bottoms Pump

$$\text{Purchased equipment cost} = C_e = CS^n$$

From table 6.2 in Coulson Richardson's Chemical Engineering Series Volume 6,

$$C = \$1920$$

$$n = 0.8$$

$$S = \text{driver power in kW} = 0.7726 \text{ kW}$$

$$C_e = \$1920 \times 0.773^{0.8} = \$1562$$

Pumps Specifications	Reflux Pump	Bottom Pump
C	1920\$	1920\$
n	0.8	0.8
S	0.0025kW	0.7726kW
C _e (Purchased Cost)	15.9\$	1562\$

9.1.6 Total purchase cost of major equipment items (PCE)

Equipment	PCE
Stripper column	\$27,344
Economizer	\$10000
Flash Vessel	\$45685
Reflux Pump	\$15.9
Bottom Pump	\$1562
Total PCE	\$84607

9.1.7 Inflation

Financial Year	Cost of Inflation Index (CII)
2001-02 (Base year)	100
2002-03	105
2003-04	109
2004-05	113
2005-06	117
2006-07	122
2007-08	129
2008-09	137
2009-10	148
2010-11	167
2011-12	184
2012-13	200
2013-14	220
2014-15	240
2015-16	254
2016-17	264
2017-18	272
2018-19	280
2019-20	289
2020-21	301

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Cost of inflation index in 2004 = 113

Cost of inflation index in 2021 = 301

Total PEC in 2004 = **\$84607**

Total PEC in 2021 = **\$84607** $\times \frac{301}{113}$

Total PEC in 2021 = **\$225369**

9.2 Estimation of Fixed Capital Cost

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

*Omitted for minor extensions or additions to existing sites.

Table 46 Typical Factors for Estimation of Project Fixed Capital Cost

PPC is calculated using the formula:

$$PPC = PCE(1 + f_1 + \dots + f_6)$$

Fluids have been selected as the process type because no solids are involved in our plant operation.

Item	PCE
Equipment erection (f1)	0.4
Piping (f2)	0.7
Instrumentation (f3)	0.2
PPC (\$)	\$292979.8

Table 47 Calculation of PPC

The calculation of Fixed Capital is done using the equation:

$$FC = PPC(1 + f_{10} + f_{11} + f_{12})$$

Item	PCE
Design and Engineering (f10)	0.3
Contractor's fee (f11)	0.05
Fixed capital (\$)	\$304248.3

Table 48 Calculation of Fixed Capital

Working Capital is estimated to be 10-20% of the fixed capital. We have assumed an average value of 15% to calculate the working capital. However, since our plant is just an extension of an existing plant, the working capital will be covered by the already present plant.

$$\text{Total investment} = \text{Fixed Capital} + \text{Working Capital}$$

Total Investment (\$)	\$304248.3
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9.3 Annual Operating Cost

9.3.1 Fixed Operating Cost

During plant operation, the operating costs are calculated using Table 6.6 of Coulson Richardson Chemical Engineering Design Volume 6.

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

Figure 59 Data for Calculation of Operating Cost

Utility	UK	USA
Mains water (process water)	60 p/t	50 c/t
Natural gas	0.4 p/MJ	0.7 c/MJ
Electricity	1.0 p/MJ	1.5 c/MJ
Fuel oil	65 £/t	100 \$/t
Cooling water (cooling towers)	1.5 p/t	1 c/t
Chilled water	5 p/t	8 c/t
Demineralsed water	90 p/t	90 c/t
Steam (from direct fired boilers)	7 £/t	12 \$/t
Compressed air (9 bar)	0.4 p/m ³ (Stp)	0.6 c/m ³
Instrument air (9 bar) (dry)	0.6 p/m ³ (Stp)	1 c/m ³
Refrigeration	1.0 p/MJ	1.5 c/MJ
Nitrogen	6 p/m ³ (Stp)	8 c/m ³

Figure 58 Data for calculation of Utility cost

The cost of utilities is calculated using Table 6.5 of Coulson Richardson Volume 6.

The fixed operating cost is the sum of all the items ranging from 5 to 12.

$$\text{Total base salary} = \text{Annual Base Salary} * \text{Number of operators}$$

$$\text{Total operating labor} = \text{Total base salary} + \text{Allowances}$$

$$\text{Supervision} = 0.2 * \text{Total Operating Labour}$$

$$\text{Plant Overheads} = 0.5 * \text{Total Operating Labour}$$

$$\text{Capital Charges} = 0.1 * \text{Fixed Capital}$$

$$\text{Local taxes} = 0.02 * \text{Fixed Capital}$$

Item	%	Cost
Maintenance = 0.05*FC	0.05	\$15212.4
Operating Labor		
Annual Base salary (\$)	5700	
Shifts	2	
Operators	1	
Total Base Salary (\$)	5700	
Total operating labor (\$)	5700	
Supervision (\$)	0.2	Not required
Plant Overheads (\$)	0.5	Not required
Capital Charges (\$)	0.1	\$30424.8
Local Taxes (\$)	0.02	\$6049
Total fixed operating cost (\$)		\$ 51722.2

Table 49 Calculation of Fixed Operating Cost

9.3.2 Variable Operating Cost

The variable operating cost covers the following items:

Raw materials	
Benfield Process Condensate	Free of cost as it is a waste product currently
$\text{Steam, at } 10.59 \text{ \$/ton} \times 8322\text{h} \times \frac{3100 \text{ kg/h}}{1000 \text{ kg/ton}} = \$273,202.9$	
$\text{Power, at } 1\text{\$/MJ} = 2.7 \text{ (MJ/hr)} \times 8322\text{h} = \23211	
Utilities	
Total cost of PW (\$)	33201
Steam Cost (\$)	\$273202.9
Total Utilities Cost (\$)	\$306404

Table 50 Calculation of Variable Operating Cost

$$\text{Total variable cost} = \text{Raw material} + \text{Utilities}$$

Total Variable costs (\$)	\$306404
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9.4 Annual Production Cost

Annual production cost = Fixed operating cost + Variable operating cost

Annual Production Cost (\$)	\$358126.2
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Table 51 Calculation of Annual Production Cost

The production cost per kg is calculated using the formula:

$$\text{Production cost} \left(\frac{\$}{\text{kg}} \right) = \frac{\text{Annual Production Cost}}{\text{Annual Production Rate}}$$

Production cost (\$/kg)	0.00188
-------------------------	---------

9.5 Annual Revenue

Revenue = Total selling of Stripped PC BFW + recovered gases

*= Selling price per unit mass * mass produced*

$$\text{Revenue} = \$3. \frac{24}{\text{ton}} \times 22.82 \text{ ton} + 27\text{kgNH}_3 \times \$0.3/\text{kg} + 79\text{kgCO}_2 \times \$0.05/\text{kg}]$$

$$* 8322\text{hr}$$

$$\text{Annual Revenue} = \$714027$$

9.6 Payback Period

Annual Cash Flow = Annual Revenue – Annual production cost

$$\text{Payback period} = \frac{\text{Total investment}}{\text{Annual Cash Flow}}$$

Payback Period	
Annual Revenue \$	714027
Annual Cash Flow (\$)	355900.8
Payback period (years)	0.855

Table 52 Calculation of Payback Period

CONTROL AND INSTRUMENTATION

10.1 Significance of Process Control

The physical conditions and raw materials are always changing in a plant environment. Sometimes, small changes in input can create a large impact on the output. Hence, it is important to regulate our process. This can be achieved through process control and instrumentation. Industries control their processes for three reasons:

- Enhance product quality by precise control of process parameters
- Ensure safety by operating within specified limits
- Maximize profitability by optimally using energy and raw material

10.2 Closed Control Loops

For this purpose, control loops are installed on equipment. There are two types of control loops, *open loop*, and *closed loop*. In open loop control the value of process variable is not compared to a setpoint and the manipulated variable is adjusted without taking feedback from the system.

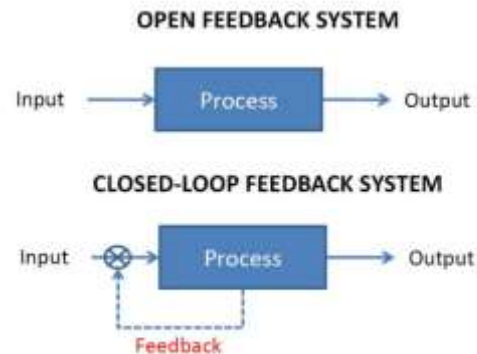


Figure 60 Feedback System

The *closed loop control* is generally applied in industries, and it functions by:

- Measuring the controlled variable,
- Comparing its value to a set point,
- Adjusting the manipulated variable.

A control loop consists of the following components,

10.2.1 Primary element/sensor

This is the first component of the loop which includes a sensor that shows a change in its physical property when the value of process variable changes. Hence, it converts the change in measured physical property to a mechanical signal.

10.2.2 Transducer

It converts the mechanical signal from the sensing element to an electrical signal.

10.2.3 Converter

A convertor is used to convert one type of signal to another type of signal. For example, an analog signal might be converted to a digital signal, or a current signal might be converted to a pneumatic signal.

10.2.4 Transmitter

A transmitter converts the sensor reading coming from the transducer to a standard signal and sends it to a controller or monitor.

10.2.5 Controller

Controller receives data from the measurement and compares it to the setpoint. There are three types of controllers that are discussed in the next section. A controller is usually present in a digital control system.

10.2.6 Final Control Element

This component of control loop physically changes the manipulated variable based on the signal received from the controller. For example, a valve acts as a final control element in many closed loop control systems.

10.3 Types of Controller Algorithms and their Applications

A controller can be designed to run on different mechanisms. [20]

- **Discrete controllers** have only two states of operation; on and off. They can only hold the controlled variable in a range of values which is known as dead zone.

- **Multi-step controllers** as the name suggests have one more state in addition to On and Off. Due to this the controlled variable response is less fluctuating as compared to discrete controllers.
- **Continuous Controllers** are more complex than the first two and provide a precise control over process variables. There are three modes of continuous controllers.

10.3.1 Proportional Action

In proportional Action control loop the controller receives the measured value and compares it with the set point to calculate the error. The output of proportional control is proportional to this error and can be defined by the following equation:

$$P(t) = K_p \varepsilon(t) + P(0)$$

Where,

$P(t)$ = output of controller

K_p = proportional controller gain

$P(0)$ = Controller output when $\varepsilon(t)$ is 0

Here K_p or proportional gain determines the sensitivity of the controller. It is the ratio of percentage change in controller output to the percentage change occurring in the input of controller. An inherent characteristic of proportional mode is offset. It is the deviation of the controlled variable from the set point after the control action has been applied.

10.3.2 Integral Action

In integral control mode, the Controller output is proportional to the amount of time the error exists. The integral control output can be mathematically expressed as,

$$P(t) = K_I \int_0^t \varepsilon dt + P(0)$$

The reciprocal of K_I is called integral time τ_i . Advantages of integral action include:

- Integral controllers give a slow response in the beginning, but over time they tend to eliminate errors.
- The integral controller removes the offset, but its slow transient response can lead to instability.

10.3.3 Derivative Action

The output of controller in derivative control action is proportional to the time rate of change of error. It can be mathematically expressed as,

$$P(t) = K_d \frac{d\varepsilon(t)}{dt} + P(0)$$

With sudden changes in the system the derivative controller will compensate the output fast. 2. The long-term effects the controller allows huge steady state errors. 3. A derivative controller will in general have the effect of increasing the stability of the system, reducing the overshoot, and improving the transient response. Advantages of derivative action include,

- When changes in the system are abrupt the derivative controller compensates the output fast.
- A derivative controller is used to increase the stability of the system by decreasing the overshoot and enhancing the transient response.

10.3.4 The Proportional Integral Derivative Action - Three Mode Controller

In PID control action the characteristics of proportional, integral, and derivative controls are combined to provide better control of the process variable. A PID controller can be mathematically expressed as,

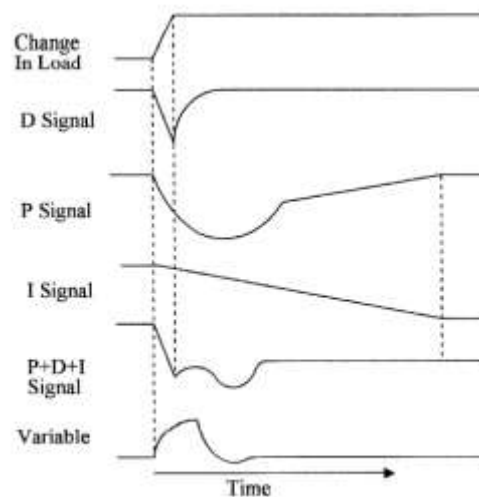


Figure 14.6 Waveforms for proportional plus integral action and waveforms for proportional plus derivative and integral action.

Figure 61 Comparison of PI and PID action [23]

$$P(t) = K_p \varepsilon(t) + K_I \int_0^t \varepsilon dt + K_d \frac{d\varepsilon(t)}{dt} + P(0)$$

Comparing the waveforms of different modes of control action above shows the advantages of PID control action. In PID control,

- The offset of proportional mode is eliminated.
- The control provided by this mode is most stable and accurate.
- The response time is as fast as derivative mode but with much smaller overshoot.
- For frequent changes in load, set point, and available energy; PID mode of action provides the best results.

Although PID controller is the best type of controller, but it is not always beneficial to apply all modes of action in every control loop. The table below shows the types of control actions suitable for control loops for different process variables.

10.3.5 The Appropriate Mode of Action for Different Control Loops

Different types of controllers are suitable for different types of loops. Some guiding rules are mentioned below and summarized in the table.

- Pressure control requires proportional and integral only. Derivative mode is normally not required.
- Level control requires proportional and occasionally integral mode. Derivative is unnecessary.
- Flow control can also be better carried out with proportional and integral modes without the derivative mode of action.
- Temperature control uses PID controller usually with a large value of integral time.

Controlled Variable	Proportional Control	PI Control	PID Control
Flow	Yes	Yes	No
Level	Yes	Yes	Rare
Temperature	Yes	Yes	Yes
Pressure	Yes	Yes	Rare
Analytical	Yes	Yes	Rare

Table 53 Appropriate mode of actions for different control loops^[23]

10.4 Process Control Loops on PC Stripping System

Four different control loops are required in our proposed Benfield condensate stripping system to precisely obtain the required purity of boiler feed water. The four control loops are summarized in the table below. They are discussed along with their ISA symbology-based Process and Instrumentation diagrams (P&ID) in the following sections.

Equipment	Controller	Type of Controller	Manipulated Variable	Controlled Variable
Economizer (Heat Exchanger)	Temperature Controller	PID	Heating Fluid (Bottoms) Flow Rate	Process Condensate Feed temperature
Stripping Column	Temperature Controller	PID	Steam Flow Rate	Bottoms Temperature
Stripping Column	Analytical Controller (Specific Conductance)	PI	Steam Flow Rate	Ammonia Content in Boiler Feed Water
Knock-out Drum	Level Controller	PI	Reflux Flow Rate	Liquid level

Table 54 Process Control Loops on PC Stripping System

10.4.1 Temperature Control Loop on Economizer (Feed Preheater)

Economizer is a heat exchanger that recovers thermal energy from bottom product of the stripper and uses it to raise the temperature of process condensate feed to the stripper. The temperature of the process condensate which also affects the temperature of stripping column is crucial for proper removal of ammonia as well as minimum moisture entrainment in vapor overhead. A temperature lower than the set point will increase ammonia content in boiler feed water, whereas a temperature much higher than the set point can cause loss of water by vaporization into the overhead.

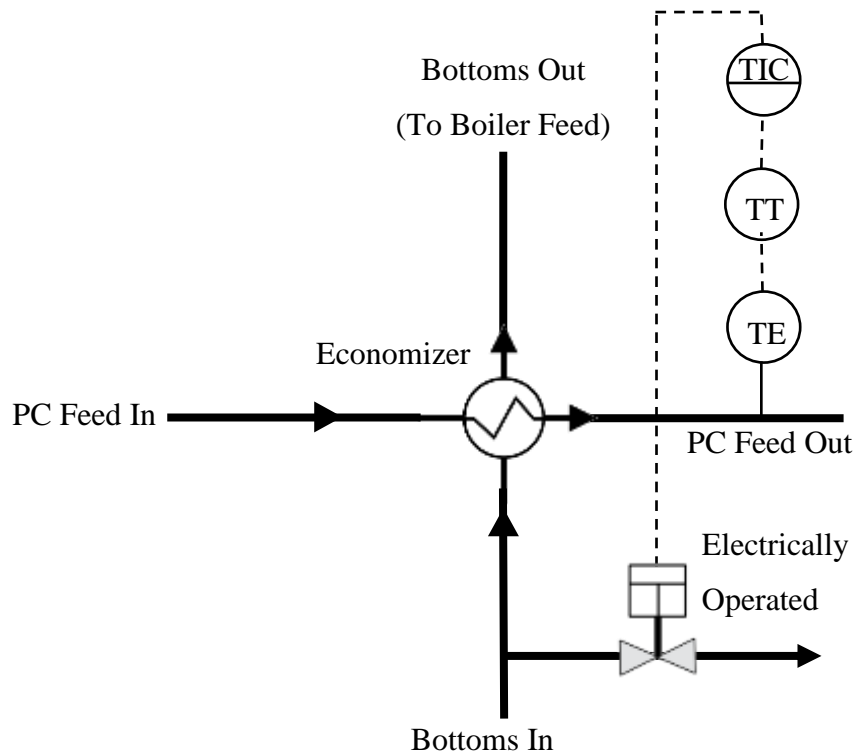


Figure 62 Temperature Control Loop on Economizer

A change in flowrate or temperature of the bottoms can influence the temperature of the process condensate feed. This change can be sensed by a temperature sensing element (TE). The measured signal is transferred to through a temperature transmitter (TT) to a temperature indicator controller (TIC) located in the control room. For temperature control, the proportional-integral-derivative PID control action is best suited. The TIC sends an electrical signal to the final control element (FCE). Here an electrically operated valve has been used as FCE. The FCE valve adjusts the flow rate of bottoms by allowing some quantity to flow through the bypass directly to boiler feed.

10.4.2 Temperature Control Loop on Stripping Column (Feed Preheater)

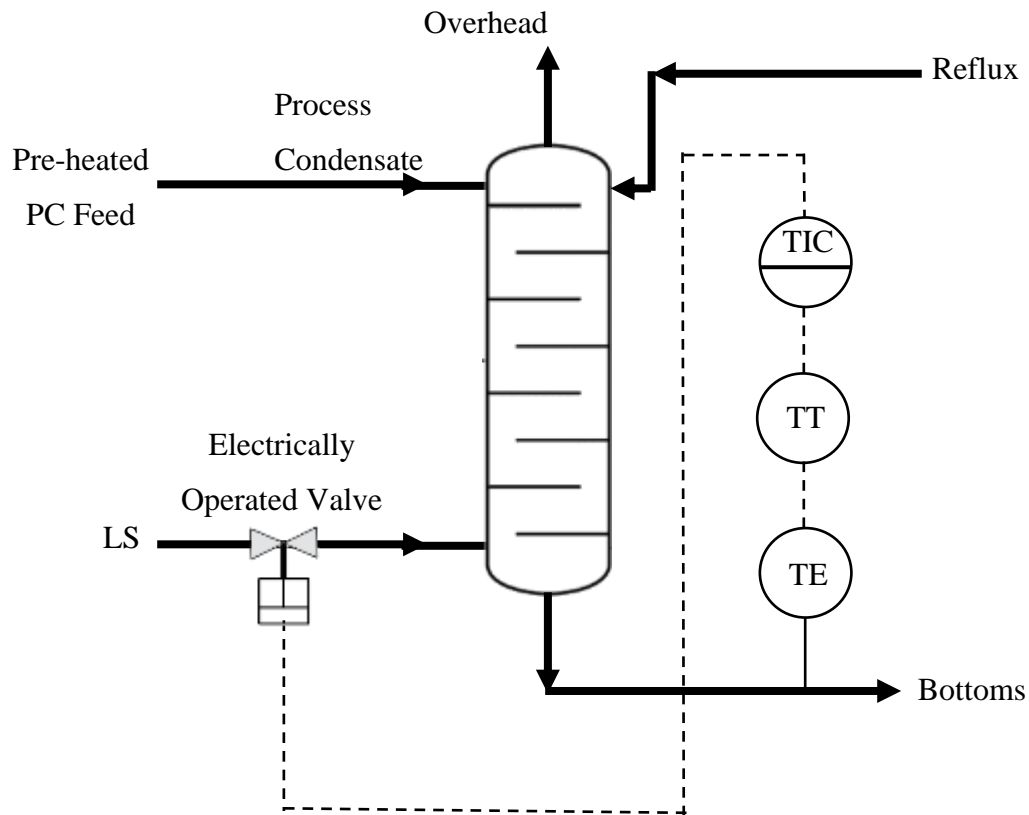


Figure 63 Temperature Control Loop on Stripping Column

Two types of control loops have been applied in the stripping column. The first one is temperature control loop which manipulates the steam flow rate to control the temperature of the stripper bottoms which is a parameter indicating the maximum temperature inside the stripper. The temperature sensing element (TE) measures the bottoms temperature and transmits the signal to temperature indicating controller (TIC). Furthermore, an electrically operated valve uses a PID controller to manipulate the steam flow rate according to the magnitude of error.

10.4.3 Specific Conductance Control Loop on Stripping Column (Feed Preheater)

Specific conductance or the ability to carry electric current is directly related to the dissolved ions in water. These ions can be contributed by both dissolved gases and dissolved solids. Due to their electrolytic nature, NH_3 and CO_2 are also present in water in their ionic form. Since our feed is essentially free of Total dissolved Solids TSS, we can rely on the specific conductance to study the NH_3 and CO_2 content in our bottom product. According to the Handbook of Industrial Water Treatment by Suez water technologies and solutions, conductivity of water increases

- 8.0 – 9.0 μS per ppm of NH_3
- 5.0 μS per ppm of CO_2

The conductivity meter acts as the primary sensing element and is denoted as CE (Conductivity sensing element). It measures the water conductivity and conductivity

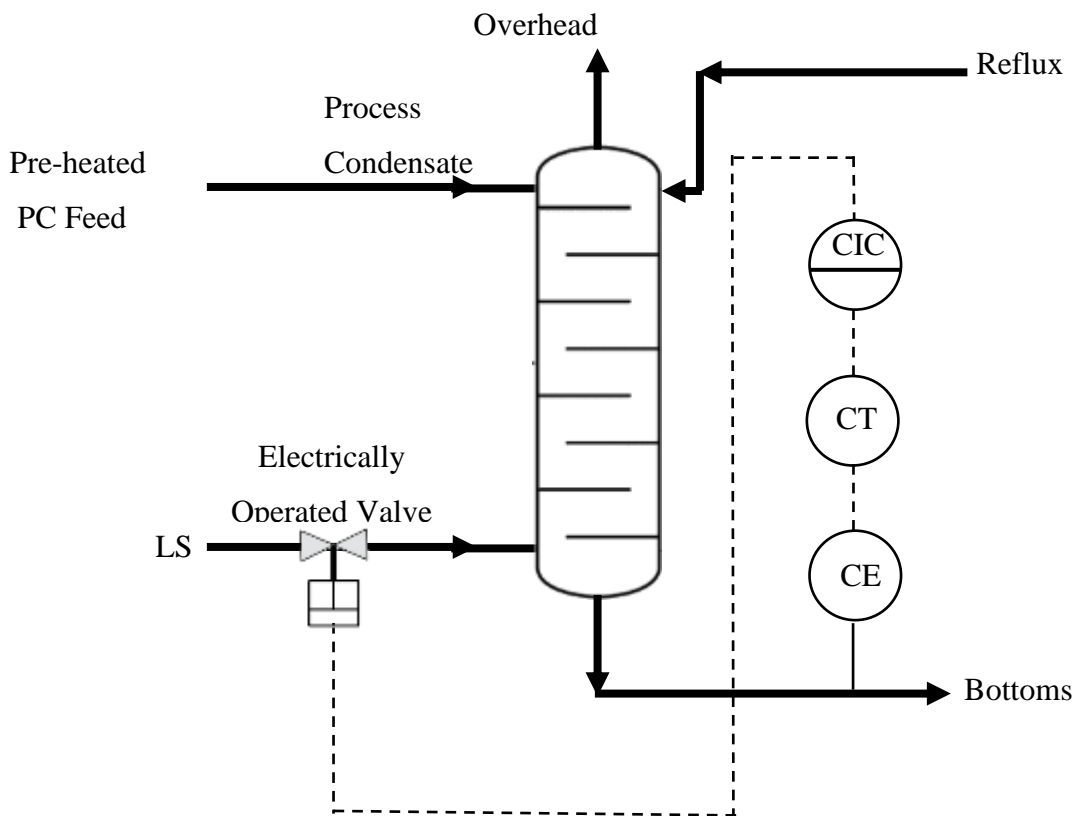


Figure 64 Specific Conductance Control Loop on Stripping Column

transmitter CT transmits the signal to conductivity indicator controller CIC. In analytical control, proportional integral controller action is used. The output of CIC manipulates the steam flow rate by adjusting the electrically operated valve which is our final control element.

10.4.4 Level Control Loop on Knock-out Vessel

Knock-out drum separates the excess moisture from CO₂ vapor stream that is being sent to the Urea section after mixing it with the PC stripper vapor overhead. To control the level of liquid in the knock-out drum, a level control loop is applied.

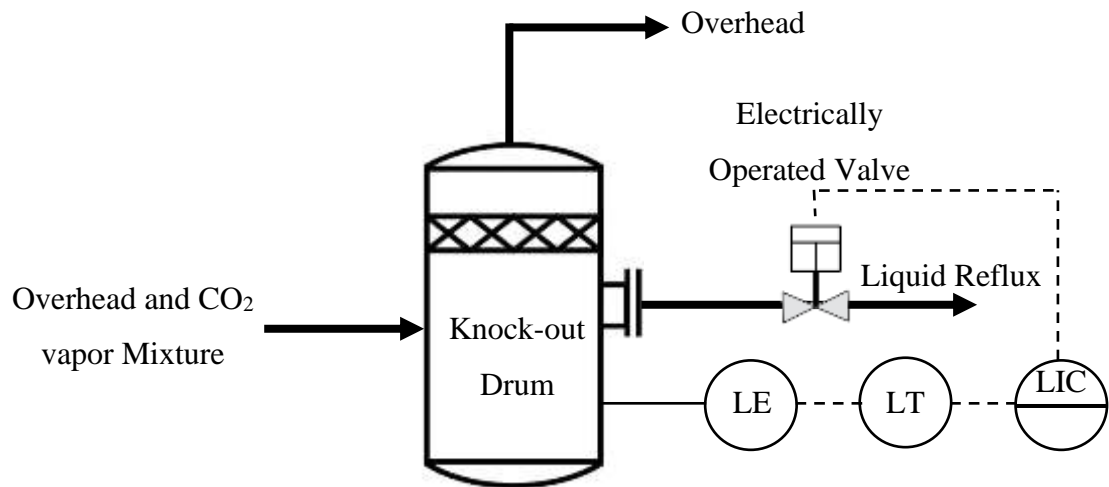


Figure 65 Level Control Loop on Knock-out Vessel

The level control loop consists of level sensing element LE, level transmitter LT and level indicator controller LIC. The proportional integral PI control is applied for level control. An electrically operated valve manipulates the reflux flow rate to maintain the liquid level at the given set point.

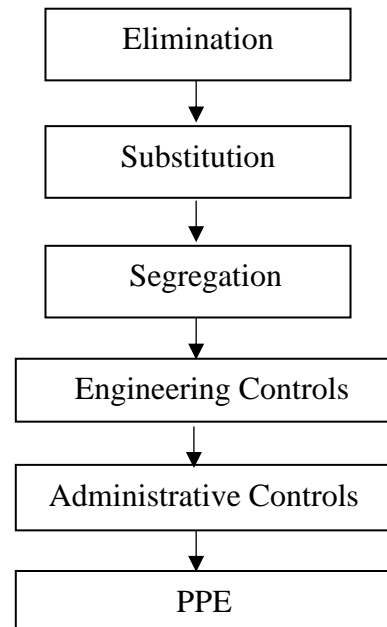
CHAPTER 11

HAZOP ANALYSIS

As we look at different manufacturing plants, including the ammonia and Urea Plant, it is important to adopt a wide range of procedures to ensure the smooth and reliable plant operation as well as the well-being and safety of the staff. It is also crucial to prevent the emergence of any hazards and operability issues. These safety measures are classified into four different categories as mentioned below:

11.1 Design strategies and continuous inspection

- Location of administration building away from the plant
- Plant boundaries consolidation with strong walls for safety and security
- Incorporation of noise controlling measures
- Proper illumination
- Proper ventilation for displacement of harmful oxides with air
- Minimization of fire risks by construction of roads
- Proper accessibility throughout the plant
- Easy accessibility of safety lifts, ladders and other instrumentation equipment
- Proper insulation of equipment for conservation of temperatures and safety of operators
- Proper guarding and inspection of rotating and moving parts
- Walkthrough audits to monitor the effectiveness of existing procedures and policies



11.2 Health and Safety Communication

- Periodic training of all employees regarding prevention of risks and vocational health and safety
- Induction of safety signs in compliance with national and international standards
- Proper tagging and labelling of all equipment
- Demarcation of different areas of the plant and walkways
- Provision of instruction manuals for different equipment
- Establishment of access point to material safety data sheet

11.3 Safety Practices and Personal Protective Equipment (PPE)

- Preparation of suitable work schedules to avoid fatigue
- Proper cleaning facilities for workers to use during and after shifts
- Medical testing for proper monitoring of employee health
- Usage of personal protective equipment
- Implementation of safety protocols other than personal protective equipment depending on likelihood and severity of the hazard

Type of PPE	Application
Hard Hat	Protects skull from impacts, penetration and electrical injuries especially in compact areas (converter and absorption column) and in multilevel units with restricted head space
Safety Shoes	Protection against rolling and falling objects, electrical hazards, hot surfaces and hot liquid splashes Recommended to be worn throughout the plant
Goggles	Protection against flying fragments, dusts and hot liquid splashes Recommended to be worn near high concentration oleum and Sulphuric acid production units
Respirators	Protection against inhalation of toxic vapors such as oxides of Sulphur, Sulphuric Acid mist etc.

Table 55 Types of PPE and its Application

11.4 Emergency Planning

- Allocation of assembly points near admin building in case of emergencies
- Incorporation of a functioning, frequently tested system of alarms
- Induction of fire extinguishers at different locations

- Sufficient stock of first aid material due to excessive exposure to sulphuric acid and oleum
- Directions for all first aid equipment (showers, eye-wash stations, stretchers, kits) throughout the plant
- Installation of health care centre near the plant
- Detailed procedures for immediate evacuation in case of emergency
- Back up services such as extra water tanks and fire ponds

11.5 Detailed HAZOP Analysis of Benfield PC Stripping System

A hazard and operability study is used for systematic, critical, examination of operability of a process. It can be applied to a process design or operating plant to find out the potential hazards that may arise from deviations that occur instead of intended design conditions.

11.5.1 HAZOP Analysis on Economizer (Shell and Tube Heat Exchanger)

Process Parameter	Guide Word	Possible Causes	Consequences	Actions
Temperature	Higher	<ul style="list-style-type: none"> Higher Shell-side inlet temperature Higher shell-side inlet flow rate Higher tube-side inlet temperature Decrease in tube side flow rate 	<ul style="list-style-type: none"> Increase in temperature of heated stream fed to stripping column Decrease in CO₂ solubility can increase CO₂ vapor formation which can cause an increase in pressure. 	<ul style="list-style-type: none"> Apply flow control to maintain heating fluid flow rate into economizer Apply ratio control loop to adjust heating fluid flow rate for fluctuations in cold feed stream flowrate.
	Lower	<ul style="list-style-type: none"> Decrease in shell side flow rate Increase in tube side flowrate Lower shell side inlet temperature Increase in fouling factor 	<ul style="list-style-type: none"> The feed stream to stripper will enter at temperature lower than designed value NH₃ and CO₂ separation in stripper would be insufficient. 	<ul style="list-style-type: none"> Check for fouling and clean tubes and shell Maintenance of flow control loops Apply ratio control loop to adjust heating fluid flow rate for fluctuations in cold feed stream flowrate.

Process Parameter	Guide Word	Possible Causes	Consequences	Actions
Flowrate	Higher	<ul style="list-style-type: none"> Decrease in flow of bottoms of PC stripper due to water entrainment in overhead Higher stripping temperature Leakage in pipelines or heat exchanger 	<ul style="list-style-type: none"> The desired temperature increase will not be achieved hence stripping operation will be affected 	<ul style="list-style-type: none"> Supply alternate source of heating fluid in case bottoms flow rate decreases Check for leakages
	lower	<ul style="list-style-type: none"> Increase in shell side inlet flowrate due to more steam condensing back into stripper bottom. Lower operating temperature of stripper Increase in flow of incoming process condensate feed from CO₂ removal unit than designed for. 	<ul style="list-style-type: none"> More flow rate, more turbulence, greater heat transfer, the temperature of the feed increases. Pressure drop increase along shell side Pressure drop increase along tube side 	<ul style="list-style-type: none"> Apply ratio control loop to adjust heating fluid flow rate for fluctuations in cold feed stream flowrate. Install a bypass for stripper bottoms incase bottoms flow rate is more than desired

11.5.2 HAZOP Analysis on Stripper (Distillation Column)

Process Parameter	Guide Word	Possible Causes	Consequences	Actions
Flowrate	Higher	<ul style="list-style-type: none"> • Increase in Benfield Process Condensate flow rate • Lower stripping temperature due to heat losses or insufficient feed preheating • Increase in steam flow rate 	<ul style="list-style-type: none"> • Flooding in the column • Product quality affected • Temperature decrease • Rise in bottom 	<ul style="list-style-type: none"> • Install high level alarm, check inlet valve • Check maintenance procedure • Apply ratio control to adjust steam flow rate for increase in Process Condensate flow rate • Make bypass for excess feed and bottom stream flows
	Lower	<ul style="list-style-type: none"> • Inlet pipe or valve blockage • Blockages in the plates • Decrease in PC Feed flowrate 	<ul style="list-style-type: none"> • Temperature increase • Increase in moisture entrainment in overhead 	<ul style="list-style-type: none"> • Ensure cleaning of plates to remove blockage • Clean inlet pipe and check inlet valve • Apply ratio control to adjust steam flow rate for increase in PC feed flow rate
	None	<ul style="list-style-type: none"> • Inlet pipe breakage • Pump failure • Control Valve fully closed 	<ul style="list-style-type: none"> • Column dry-out • Column damage due to direct contact with steam 	<ul style="list-style-type: none"> • Apply low flow alarm • check inlet pipe, valve and pump

Process Parameter	Guide Word	Possible Causes	Consequences	Actions
Temperature	Higher	<ul style="list-style-type: none"> • More than desired feed preheating • Increase in steam temperature or flowrate 	<ul style="list-style-type: none"> • Increase in stripper operating temperature • Increase in moisture entrainment in overhead • Decrease in bottoms flowrate 	<ul style="list-style-type: none"> • Apply temperature control loops to regulate feed preheating • Make bypass line for bottoms to avoid overheating of feed • Adjust reflux flowrate • Apply temperature and flow control loop on steam
	Lower	<ul style="list-style-type: none"> • Less than desired feed preheating • Increase in feed flow rate • Heat loss to surroundings • Decrease in steam temperature or flow rate 	<ul style="list-style-type: none"> • Decrease in column temperature • Desired separation of ammonia is not achieved 	<ul style="list-style-type: none"> • Adjust the quantity of steam for increase in feed flow rate. • Install low temperature alarm
Pressure	Higher	<ul style="list-style-type: none"> • More than desired feed preheating • Increase in inlet steam pressure • Outlet pipe or valve blockage 	<ul style="list-style-type: none"> • Increase in bottoms flow rate • Desired ammonia separation is not achieved • Rupture or breakage of the vessel 	<ul style="list-style-type: none"> • Apply pressure safety valve PSV • Apply high pressure alarm on stripper • Check steam source to regulate steam pressure
	Lower	<ul style="list-style-type: none"> • Decrease in steam pressure • Breakage in vessel wall • Heat loss to surroundings 	<ul style="list-style-type: none"> • Increase in moisture entrainment in overhead • Decrease in bottoms flowrate 	<ul style="list-style-type: none"> • Apply low pressure alarm • Apply pressure control loop on steam • Check for breakage in vessel wall • Reduce heat loss to surrounding by insulation

11.5.3 HAZOP Analysis on Knockout Drum

Process Parameter	Guide Word	Possible Causes	Consequences	Actions
Level	Higher	<ul style="list-style-type: none"> Higher moisture content in inlet vapor stream Blockage of outlet pipe or valve Decrease in reflux flow due to control valve malfunctioning 	<ul style="list-style-type: none"> Significant increase in liquid level in drum can cause overflow Separation will not be achieved Stripper operation affected due to decrease in reflux flow 	<ul style="list-style-type: none"> Apply level control loop on separator Check inlet valve and pipe for blockage Apply high level alarm Optimize stripping conditions to decrease moisture entrainment
	Lower	<ul style="list-style-type: none"> Decrease in moisture content of inlet vapor stream Decrease in stripper temperature causing inadequate heat transfer and all feed enters the bottoms 	<ul style="list-style-type: none"> Reflux drops to zero if required liquid level is not achieved and affects stripper operation 	<ul style="list-style-type: none"> Apply temperature control on stripper Adjust stripper parameters according to the reflux ratio
Pressure	Higher	<ul style="list-style-type: none"> Blockage of outlet pipe or valve 	<ul style="list-style-type: none"> Improper separation Vessel might leak or blast 	<ul style="list-style-type: none"> Install high pressure alarm Install PSVs
	Lower	<ul style="list-style-type: none"> Leakage in the lines or vessel 	<ul style="list-style-type: none"> Improper separation 	<ul style="list-style-type: none"> Install low pressure indicator Check for leakages

RESULTS AND FUTURE PROSPECTS

After describing our methodology, process description, and profitability analysis of our proposed process condensate stripping system in our thesis, pertinent results and future prospects are appended below.

Reliability

We have designed a process condensate treatment system based on a proven and commercially used methodology of steam stripping that is not present at FFBL's ammonia plant. Application of this method has shown to yield the desired purity of condensate product as discussed in our literature review. In addition, our case studies predict that steam stripping can be optimized for changes in feed conditions by varying the flow rate of steam.

Scale-up

Out of the 84000 kg/h of process condensate produced at FFBL, 22530 kg/h waste Benfield condensate is currently wasted. We have taken this waste as feed for our stripping system. The remaining process condensate which is reused without treatment and has an unfavorable effect on the process can also be combined with this feed however the steam flow rate will have to be adjusted accordingly.

Feasible

The principal utility required for our proposed system is low pressure steam which is already available at plant. The feed to steam ratio is approximately 7:1.

Profitability

The cost effectiveness and payback duration of the proposed design brings the conclusion that this is a viable and profitable addition to the plant which will enhance water economy.

Sustainability

The initiative of utilizing process condensate as boiler feed water is in line with United Nation's Sustainable Development Goals and will help avoid the depletion of natural resources in order to maintain an ecological balance.

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