Manufacturing of Urea By

Snamprogetti Process for 1900 MTPD

Muhammad Sharjeel Khan	CHEN18111042
Muhammad Naveed	CHEN18111015
Muhammad Gulfam	CHEN18111022

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Date

22-June-2022

Approved by
Dr. Asim Riaz

Supervisor name
Engr. Afaq

Major Advisor name
Engr. Muhammad Suleman

This certifies that the dissertation

of (Names of candidates) is approved

Thesis Supervisor: Engr. Muhammad Suleman

_____

Thesis Committee Chairman

_____

Thesis Committee Member #1:

_____

Thesis Committee Member #2:

University Name

Khwaja Fareed University of Engineering &

Information Technology, Rahim Yar Khan

Date

22-January-2022

Acknowledgement

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Abstract

In this thesis we are manufacturing urea by Snamprogetti process, we chose this process over other methods is because this process allows a total stripping process along the PFD process. What happens in a total stripping process is that extra ammonia and evaporated water and CO_2 recycled back to the reactor. We will further elaborate on this process in the chapter 2. PFD of the process is also shown to better understand the process. Next, we will apply mass balance to the process, we have already decided to we want 1900 MT per day with 99% purity of urea product with biuret less than 1%. Similarly, next we will apply energy balance to the process and calculate heat and energy input and output.   The further intention of the project was to gain an insight into the operations of the plant and designs the equipment. Equipment parameters also important to understand, whether the material selection is accurate and is viable for that given process.

Next is the Instrumentation and control of the process, control loop is used on all single equipment's to control the parameters like Temperature, Pressure, Level, Concentration. Manipulated variables are controlled by FM, Pressure gauges, valves and coolants. Further the cost estimation of the whole process is also calculated, while the prices of the equipment is calculated by cost index method by comparing prices to prior years to current year. Also, by calculating TPC and TCC, we can obtain Profit analysis of the process. In the end we are able to achieve 5700MTPY yield with purity of 98%.

1.Chapter 1 – Introduction

In regards to fertilizers and nitrogen fertilizers urea is an important fertilizer. Especially in Pakistan the utilization is steadily increasing, also it's the preferred nitrogen fertilizer worldwide. It can be used in forms such as solid, liquid and other adhesives.

Within the 20th century urea was found to be best nitrogenous fertilizer for plants and then was later proved as an excellent fertilizer. Urea is produced on large scale and also utilized in other industries. For industrial use it is produced from carbon dioxide and synthetic ammonia. Urea can also be produced as granules, crystals, pellets and pills. In 1773 Rouelle first discovered urea in urine. The research was further continued by Friedrich Wohler by 1828. Because of urea cycle in human beings it was mainly relevant to human health. For use in fertilizer solution urea is very suitable and it is also water soluble. Urea plays important role in plants growth. Phosphorus, Potassium and nitrogen are the primary nutrients.

Fertilizer is generally defined as "any material, organic or inorganic, natural or synthetic, which supplies one or more of the chemical elements required for the plant growth". The main purpose of the fertilizer industry is to provide the primary and secondary nutrients which are required in macro quantities. While Primary nutrients are typically supplied through chemical fertilizers. It has chemical compounds containing one or more of the primary nutrients and are generally produced in a chemical reaction. Whatever the chemical compounds are present, its most important ingredient for plant growth is the nutrient content. The primary and other most important nutrients are Nitrogen, Phosphorus and Potassium. However, their concentration in a chemical fertilizer is expressed as a percentage of total nitrogen (N), available phosphate (P_2O_5) and soluble K_2O. The grade of a fertilizer is expressed as a set of three numbers in the order of percent N, P_2O_5 and K_2O. If a nutrient is missing in a fertilizer, it is represented by a zero. (CHAUHAN) (DAS)

1.1. Commercial Production of Urea

It is produced from these raw materials ammonia and carbon dioxide. When ammonia is manufactured from natural gas and air. There are many processes for urea production and unconverted reactants are then used for other products manufacturing. Two main reactions take place, one is exothermic and other endothermic.

1.2. Urea Demand

Urea's interest in world overview of 2011 is some 170.348 million MT and it is required to 180million MT in 2012.
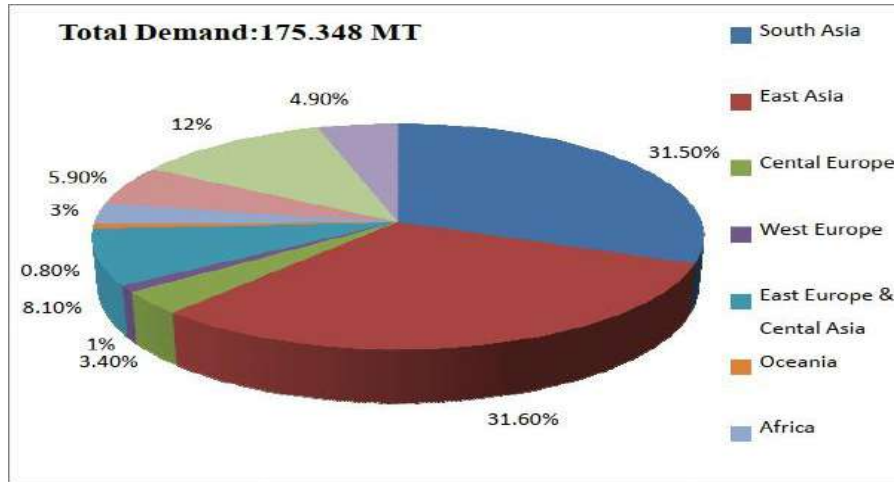
Figure 1:Urea Demand

Asia would contribute 68% in this augmentation India 25%, China 24% and Pakistan 5%. Then again, in America the significant commitment would be from Brazil (6%) and USA 6%. In Europe striking additionis normal in Russia (4%). To satisfy this interest, financial backers are attempting to introduce urea plantsin exceptionally populated nations where food request is at its pinnacle.

1.3. Urea Plants in Pakistan

Pakistan is also one of these locations where urea demand exists and is increasing day by day. The sources of fertilizer supply in Pakistan are domestic production and imports. Different firm share to the local market is as given:
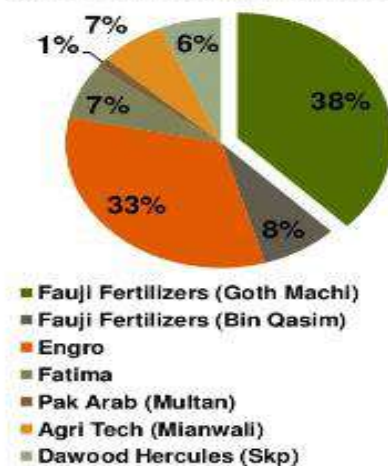
Figure 2:Production in pakistan

1.4. Physical & Chemical Properties Of Urea

1) Urea is odorless white solid and also has the property that it is non-corrosive.

2) Urea is less soluble in alcohol

3) Urea is highly soluble in water

4) Urea reacts with acid to form salts

Table 1 Chemical Properties of urea

Molecular Weight	60.05
Relative Index	60%
Minimum Nitrogen Content	46.60%
Specific Gravity	1.335
Heat Of Fusion	60 Cal/gm (endothermic)
Heat Of Solution in Water	58 Cal/gm (endothermic)
Bulk Density	0.74 gm/cc

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Table 2 Specific Heat of urea at different temperatures

Temperature (^{0}C)	Specific Heat (Kj/Kg)
0	1.398
50	1.66
100	1.89
150	2.11

1.5. Uses of Urea

1   It is mainly used as a nitrogen release fertilizer to make the product water-soluble.
2   Urea is used as a stabilizer in most of the nitrocellulose explosive products
3   It is used in manufacturing high explosive materials like urea nitrate ($CH_5N_3O_4$)
4   It is used as an important reagent in lanthanide chemistry.
5   It is used in creams/ointments that are used for rehydration.
6   It is used in hair removal creams and dish soaps.
7   It is used as a browning agent for pretzels.
8   It is used in manufacturing melamine. It is used to detect the bacteria in the stomach through the urea breath test.

1.6. Process Technology

There are many techniques used to produce urea all having different efficencies. The two main reactions involved are:

$$CO_2 + 2NH_3 \longrightarrow NH_2COONH_4 \qquad \Delta H = -37.4 Kcal/gm\ mol$$

$$NH_2COONH_4 \longrightarrow NH_2CONH_2 + H_2O \qquad \Delta H = +6.3\ Kcal/\ gm\ mol$$

Undesirable side reaction take place is

$$2NH_2CONH_2 \longrightarrow NH_2CONHCONH_2 + NH_3$$
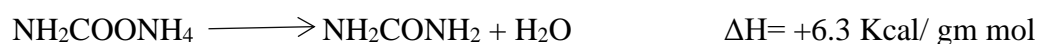
<div align="center">(Biuret)</div>

Both of these above reactions are equilibrium reactions. The first reaction completed on the temperature of 185 to 190°C and 180 to 200 atm pressure. The second reaction is slow and determines the rate of reaction. Unconverted carbon di oxide and NH_3 along with undecomposed carbamate must recycle and used again. This is tough step. The synthesis is further difficult by formation of biuret, $NH_2CONHCONH_2$, which must be kept low because its effect plant growth.

1.7. Process Description for Urea Production

Ammonia and CO_2 are basic raw material for urea. These two materials compressed first separately and fed to reactor at pressure of 180 atm which must be water cooled due to exothermic reaction. A mixture of urea, ammonium carbamate, water and unreacted carbon dioxide and ammonia is produced. This liquid is let down to 27 atm and fed to flash evaporator which contain gas liquid separator and condenser. Unreacted NH_3, CO_2 and water are recycled or removed. An aqueous solution of carbamate passed through atmospheric flash drum. The further decomposition of carbamate occurs in it. The off gases from this decomposition either recycled or moved NH_3 process for manufacturing chemical fertilizer.

The 80% of urea aqueous solution is sent to vacuum evaporator to get molten urea which contains less than 1% water in it. Then this solution sprayed into pilling tower. To avoid formation of biuret and keep it less than 1 % the temperature should be just kept above the melting point.

1.8. Urea Supply & Demand in Pakistan

Overview for last 20 years of urea demand and production is given in following graph. Pakistan has been lagging demand forever. Now days a new plant of capacity 1.3 million ton has been installed which enabled the country to meet its requirement. But demand is increasing very rapidly due to high population concentration and agriculture dependence of country's GDP. It is expected that in 2018 demand of Pakistan will reach some 9 million MT and installed capacity will remain only 6.9million ton. So, a huge demand is there for new capacity installing's

Although the production of urea is high at high pressure and high temperature, the reaction is not operated at maximum temperature and pressure because:

1. Increased pressure increases capital & operating costs of compression and reaction equipment.
2. Increased temperature accelerates decomposition of urea to biuret, a compound that adversely affects the growth of some plants.
3. The above stipulated conditions produce intolerable corrosion rates, and a compromise design must be chosen.

2. Chapter 2 – Process Description

2.1. Manufacturing Process

The snamprogetti process is based on the principle of the internal carbamate recycle technique and is commonly called the snam NH_3 stripping process. The basic difference between the snam process & the conventional carbamate solution recycle urea processes is the fact that in this case the unconverted carbamate is removed/stripped and recovered from the urea synthesis reactor effluent solution at reactor pressure, condensed to an aqueous solution in a steam producing high pressure condenser, & recycle back to the reactor by gravity. Part of the liquid NH_3 reactor feed, vaporized in a steam heated exchanger, is used as inert gas to decompose & strip ammonium carbamate in the steam heated high pressure stripper.

The reactor operates at about 130 atm & 180-190°C. The stripper operates at about 130 atm & 190°C. The stripper off-gas is condensed in a vertical shell & tube condenser, operating at about 130 atm & 148-160°C. Low pressure steam is produced in the high-pressure carbamate condenser. The urea product solution, leaving the stripper & still containing 2-3% of residual unreacted carbamate, is further degassed in a low-pressure decomposition-absorption system. The recovered ammoniacal solution of ammonium carbamate is pumped back to the reactor.

Snamprogetti ammonia-stripping urea process is selected because it involves a high NH_3 to CO_2 ratio in the reactor, ensuring the high conversion of carbamate to urea. The highly efficient ammonia stripping operation drastically reduces the recycling of carbamate and the size of equipment in the carbamate decomposition. Snamprogetti technology differs from competitors in being based on the use of excess ammonia to avoid corrosion as well as promote the decomposition of unconverted carbamate into urea. (Nigam, 2019) (DAS)

$$2NH_3 + CO_2 \longrightarrow NH_2COONH_4 \qquad\qquad \text{Exothermic}$$

$$NH_2COONH_4 \longrightarrow NH_2CONH_2 + H_2O \qquad\qquad \text{Endothermic}$$
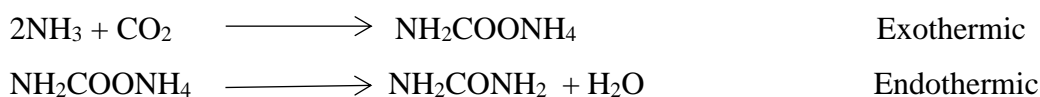
As the reactions are reversible in nature only partial conversion occurs in the reactor. Urea solution consisting of urea, carbamate, water & unconverted CO_2 & NH_3 are fed into the stripper where stripping action of NH_3 favors decomposition of carbamate, and hence 80% of carbamate is decomposed here. Pressure in the stripper is same as that of the reactor.

Urea solution from the stripper is sent to medium pressure decomposer where urea purification takes place by the dehydration of the carbamate. Urea solution is further purified in Low pressure decomposer. Off gases from the medium pressure decomposer & low pressure decomposer are sent

to the medium pressure condenser & medium pressure absorber for the recovery of unconverted Ammonia. This way 71.12% of Urea solution resulting from low pressure decomposer is sent to vacuum concentrators in operating in two stages:

1.   1st Vacuum Evaporator
2.   2nd Vacuum Evaporator

Finally, 98% molten urea is sent to the prilling towers where urea prills are formed by passing a current of cold air in the tower from the bottom. Proper size urea prills are sent to bagging section through belt conveyors. In bagging section, coating of urea prills may be done if required. Oversized urea prills or lumps are sent to lump dissolving tank.

Urea Synthesis

NH_3 & CO_2 react under specific concentration, temperature & pressure conditions to form urea as per the following reactions:

1) $CO_2(g) + 2NH_3(g) \longrightarrow NH_2COONH_4 (s)$ $\Delta H = -37.64$ kcal/gm mol

2) $NH_2COONH_4(s) \longrightarrow NH_2CONH_2 (s) + H_2O (l)$ $\Delta H = 6.32$ kcal/gm mol

$CO_2 (g) + 2NH_3(g) \longrightarrow NH_2CONH_2 (s) + H_2O (l)$ $\Delta H = -31.32$ kcal/gm mol
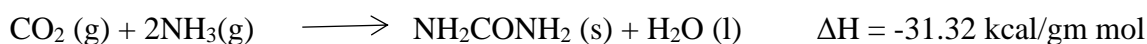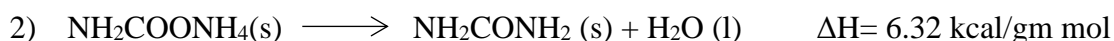
So, overall urea synthesis is exothermic, releasing heat of 31.32 kcal/gm mol at standard conditions of 1 atm pressure & 25°C. But actual heat available in a urea synthesis reaction will be only 5.74 kcal/gm mol because of the heat lost in evaporation of liquid NH_3, evaporation of water & melting of urea. This is based on the actual plant data.

Further energy is consumed in feeding CO_2 & NH_3 at high temperature & pressure, in recycling of carbamate, in vacuum concentration of urea, for operating different pumps & compressors etc. which altogether makes the urea production energy consuming.

2.2. Effect Of Various Parameters

Temperature

In above reactions 1st reaction is exothermic & 2nd one is endothermic. So, according to Le Chatelier's Principle 1st reaction is favored at low temp & 2nd one at high temp. Further, reaction no. 1 is fast & reaches to completion but reaction no. 2 is slow & determines the overall rate of urea production. For sufficient completion of reaction no.2 optimum temp is maintained. It is observed that max equilibrium conversion occurs between 190 to 200°C. If temp is increased beyond 200°C corrosion rate increases.

$$NH_2COONH_4(s) \longrightarrow 2NH_3(g)+CO_2(g) \qquad\qquad \Delta H=\text{-ve}(K1)$$

$$NH_2COONH_4(s) \longrightarrow NH_2CONH_2(s)+H_2O(l) \qquad\qquad \Delta H=\text{-ve}(K2)$$

Both the reactions are favored at high temp but our objective is to maximize 2nd reaction as 1st reaction is undesirable in the reactor. So, our operating zone should be in the region where K2 > K1.

Pressure

Overall urea synthesis reaction is given below:

$$CO_2(g) + 2NH_3(g) \longrightarrow NH_2CONH_2(s) + H_2O(l)$$
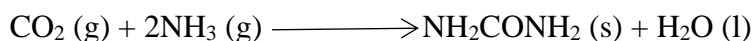
There is reduction in volume in the overall reaction & so high pressure favors the forward reaction. This pressure is selected according to the temp to be maintained & $NH_3:CO_2$ ratio.

Concentration

Higher the concentration of the reactants, higher will be the forward reaction according to the law of mass action. CO_2 being limiting reagent higher $NH_3:CO_2$ ratio favors conversion. Since, dehydration of carbamate results in urea production, lesser $H_2O:CO_2$ ratio favors conversion , water intake to the reactor should be therefore min.

Residence Time

Since, urea conversion reaction is slow, sufficient time is to be provided to get higher conversion. Reactor is designed to accommodate this with respect to the other parameters of temperature, pressure & concentration.

2.3 Snamprogetti Stripping Process

Formation of urea from ammonia & carbon-dioxide takes place through reversible reactions with formation of ammonium carbamate as intermediate product. Now, success of any urea manufacturing process depends on how economically we can recycle carbamate to the reactor. Snamprogetti process of urea manufacturing accomplishes the above task by stripping process.

$$NH_2COONH_4 \text{ (s)} \longrightarrow 2NH_3 \text{ (g)} + CO_3 \text{ (g)} \qquad \Delta H = + 37.4 \text{ Kcal/gm-mole}$$
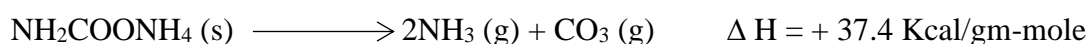
This reaction involves increase in volume & absorption of heat. Thus, this reaction will be favored by decrease in pressure & increase in temperature. Moreover, decreasing the partial pressure of either of the products will also favor the forward reaction. Process based on first principle of decrease in pressure & decrease in temp is called conventional process, whereas process based on increase/decrease of partial pressures of NH_3 or CO_2 is called stripping process. According to above equation we have:

$$K = (pNH_3)2 \text{ x } (pCO_2) \quad \text{[where K= equilibrium constant]}$$

The stripping is effected at synthesis pressure itself using CO_2 or NH_3 as stripping agent. If CO_2 is selected, it is to be supplied to the decomposers/stripper as in stamicarbon CO_2 stripping process. While if NH_3 is selected, it is to be obtained from the system itself because excess NH_3 is present in the reactor as in Snam's process. CO_2 stripping is advantageous because introducing CO_2 increase pCO_2 (partial pressure of CO_2). So pNH_3 will be reduced to maintain P constant as $P = pCO_2 + pNH_3$.

At a particular temp K is constant so when pNH_3 is reduced to keep K constant, carbamate will be reduced much faster by decomposition as pNH_3 appears in the equilibrium equation with a power of two. Selection of 1st stage decomposition should be in such a way that min water evaporates because the recovered gases go along with the carbamate to reactor again & if water enters reactor, production

will be affected adversely due to hydrolysis of urea. So, stage wise decomposition of carbamate is done. Second consideration in favor of isobaric stripping is that higher carbamate recycles pressure results in condensation at higher temp & that recovery in the form of low-pressure steam. This is why stage wise reduction in pressure is practiced.

Process Description

The urea production process takes place through the following main operations:

1. Urea synthesis & high-pressure recovery.
2. Urea purification & low-pressure recovery.
3. Urea concentration.
4. Urea prilling.

Urea Synthesis & High-Pressure Recovery

Urea is synthesized from liquid ammonia & gaseous carbon-di-oxide. . The carbon dioxide drawn from battery limits at about 1.6 atm pressure & about 40°C temp is compressed in a centrifugal compressor up to 162 atm. A small quantity of air is added to the CO_2 compressor suction in order to passivate the stainless steel surfaces. Thus, protecting them from corrosion due both to the reagent & the reaction product.

The liquid ammonia coming directly from battery limits is collected in the ammonia receiver tank from where it is drawn to & compressed at about 23atm pressure by means of centrifugal pump. Part of this ammonia is sent to medium pressure absorber & remaining part enters the high pressure synthesis loop. The NH_3 of this synthesis loop is compressed to a pressure of about 240atm. Before entering the reactor, it is used as a driving fluid in the carbamate ejector, where the carbamate coming from carbamate separator is compressed upto synthesis pressure. The liquid mixture of ammonia & carbamate enters the reactor where it reacts with compressed CO_2.

In the reactor the NH_3 & gaseous CO_2 react to form ammonia carbamate, a portion of which dehydrates to form urea & water. The fraction of carbamate that dehydrates is determined by the ratios of various reactants, operating temp, the residence time in the reactor & reaction pressure. The mole ratio of NH_3 / CO_2 is around 2:1; the mole ratio of water to CO_2 is around 0.67: 1.

$$2NH_3 \text{ (g)} + CO_2 \text{ (g)} \longrightarrow NH_2COONH_4 \text{ (s)} \qquad \text{Exothermic}$$

$$NH_2COONH_4 \text{ (s)} \longrightarrow NH_2CONH_2 \text{ (s)} + H_2O \text{ (l)} \qquad \text{Endothermic}$$

In the synthesis conditions (T= 190°C, P= 154 atm), the 1st reaction occurs rapidly & is completed. The 2nd reaction occurs slowly & determines the reactor volume.

Urea reactor is a plug flow type with 10 number of sieve trays to avoid back mixing & to avoid escape of gaseous CO_2 which must react in the lower part of the reactor. Stage wise decomposition is carried out to reduce water carry over to the reactor which could adversely affect conversion.

Urea solution containing urea, carbamate, H_2O & unconverted CO_2 & NH_3 enters the high-pressure stripper where the pressure is same as that of the reactor. The mixture is heated as it flows down the falling film exchangers. The CO_2 content of the solution is reduced by the stripping action of NH_3 as it boils out of the solution. The carbamate decomposition heat is supplied by 24atm steam. The overhead gases from stripper and the recovered solution from the MP absorber, all flow to the high-pressure carbamate condenser through mixer, where total mixture, except for a few inerts is condensed & recycled to the reactor by means of carbamate ejector. Condensing the gases at high temp & pressure permits the recovery of condensation heat in the production of steam at 4.5atm in the high-pressure carbamate condenser.

From the top of the carbamate separator the incondensable gases come out consisting of inerts & a little quantity of NH_3 & CO_2 unreacted in the condenser. These are sent to the bottom of medium pressure decomposer.

Urea Purification & Low-Pressure Recovery

Urea purification takes place in two stages at decreasing pressure as follows:

 1st stage at 18 atm pressure, i.e., Medium Pressure Decomposer

 2nd stage at 4.5 atm pressure, i.e., Low Pressure Decomposer

1st Purification & Recovery Stage at 18atm

It is falling film type medium pressure decomposer. It is divided into 2 parts: Top separator, where the released flash gases, the solution enters the tube bundle & decomposition section where the

residual carbamate is decomposed & required heat is supplied by means of 24atm steam condensate flowing out of the stripper.

2nd Purification & Recovery Stage at 4.5atm

The solution leaving the bottom of medium pressure decomposer is expanded at 4.5atm pressure& enters the low-pressure decomposer (falling film type). This is again divided in to two parts: top separator where the released flash gases are removed before the solution enters the tube bundle. Decomposition section where the last residual carbamate is decomposed & the required heat is supplied by means of steam saturated at 4.5atm.

Urea Concentration

Next section is urea concentration & objective is to reduce water content of urea to as low as 1 %. For the purpose a vacuum concentrator in two stages is provided. The solution leaving the low-pressure decomposer bottom with about 72% urea is sent to the 1st vacuum concentrator operating at a pressure of 0.23atm. The mixed phase coming out enters the gas liquid separator, wherefrom the vapors are extracted by the 1st vacuum system, while the solution enters the 2nd vacuum concentrator operating at a pressure of 0.03atm. The two concentrators are fed by saturated steam at 4.5atm. The mixed phase coming out enters the gas liquid separator, wherefrom the vapors are extracted by the 2nd vacuum system.

Urea Prilling

The molten urea leaving the 2nd vacuum separator is sent to the prilling bucket by means of a centrifugal pump. The urea coming out of the bucket in the form of drops fall along the prilling tower & encounters a cold air flow which causes its solidification. The solid prills falling to the bottom of the prilling tower are sent through the screeners to retain lumps only, & then to belt conveyor which carries the product to the automatic weighing machine & to the urea storage sections. Urea lumps by the means of belt conveyor are recycled to the underground tank, where they are dissolved.

Major Engineering Problems

1) Carbamate Decomposition & Recycle:

There are many processes that can be used for the manufacture of urea. Main difference in competing processes is in the recycle design. Since, conversion is only 40-50 % per pass, the unreacted off gases must be recirculated or used economically elsewhere. Recompression of off gases is virtually impossible because of corrosion and formation of solid carbamate in compressors.

2) Production Of Granular Urea:

Biuret formation is another problem. Vacuum evaporation of urea from 80% to about 99%, spraying to air cool and solidification must be done just above the melting point of urea and with a minimum residence time in the range of several seconds.

3) Heat Dissipation In The Autoclave:

The exothermic heat of reaction can be removed by coils or wall cooling.

4) Corrosion:

This has been the major reason why the NH_3-CO_2 process was slow to develop. High cost silver or tantalum liners are used in the autoclaves with hastealloy C, titanium, stainless steel (321 SS), and aluminum alloys used in other parts of the plant. Minimum pressure and temperature conditions with excess NH_3 are desirable to reduce the severe corrosion rates. Under these conditions, stainless steel can be used in the autoclave.
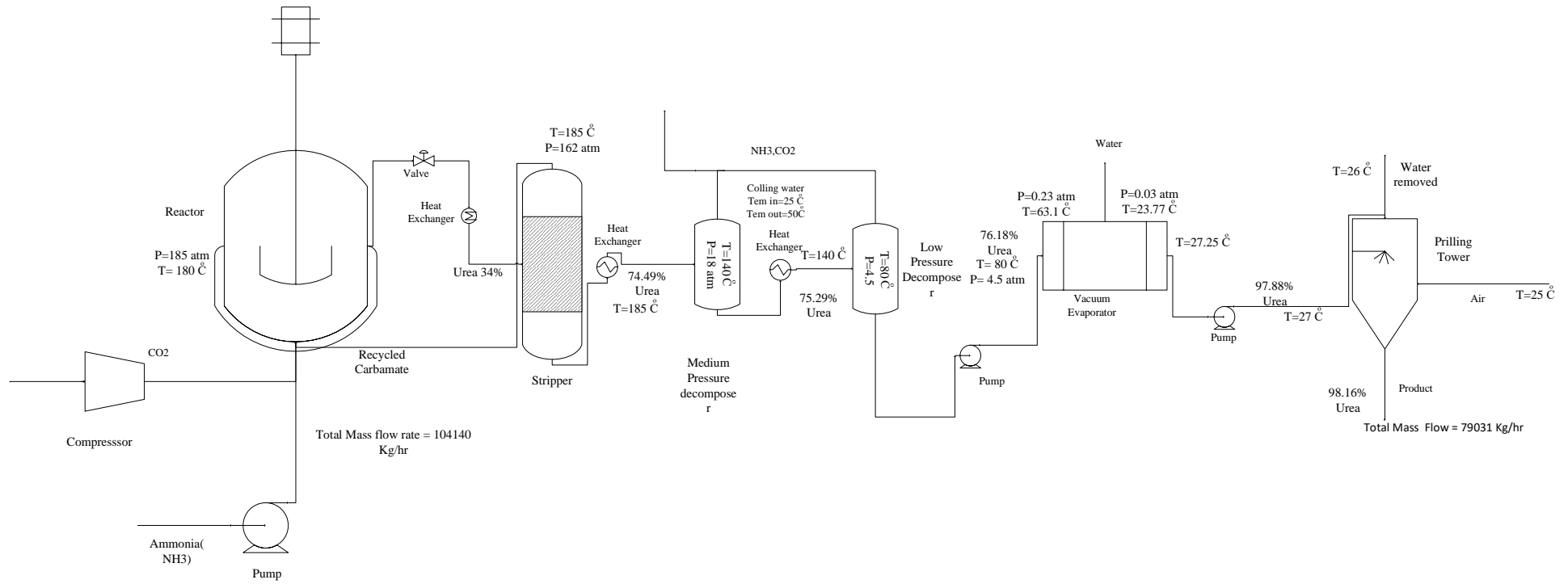
Figure 3. Process Flow Diagram of snamprogetti urea process

CO2(G)   NH3(L)

Reactor

Urea 34%
Water   Carbamate
Carbamate
NH3
CO2
Biuret

Stripper   Off gasses NH3 ,CO2

Urea 75.49%

Medium Pressure
Decomposer   Low Pressure Decomposer

NH3,CO2,Urea Biuret,Water

Water removed

WATER

Biuret,Water,Urea97.88%   Air supplied

Biuret,Water,Urea 76.16%

Prilling Tower

Vacuum Evaporator

Product
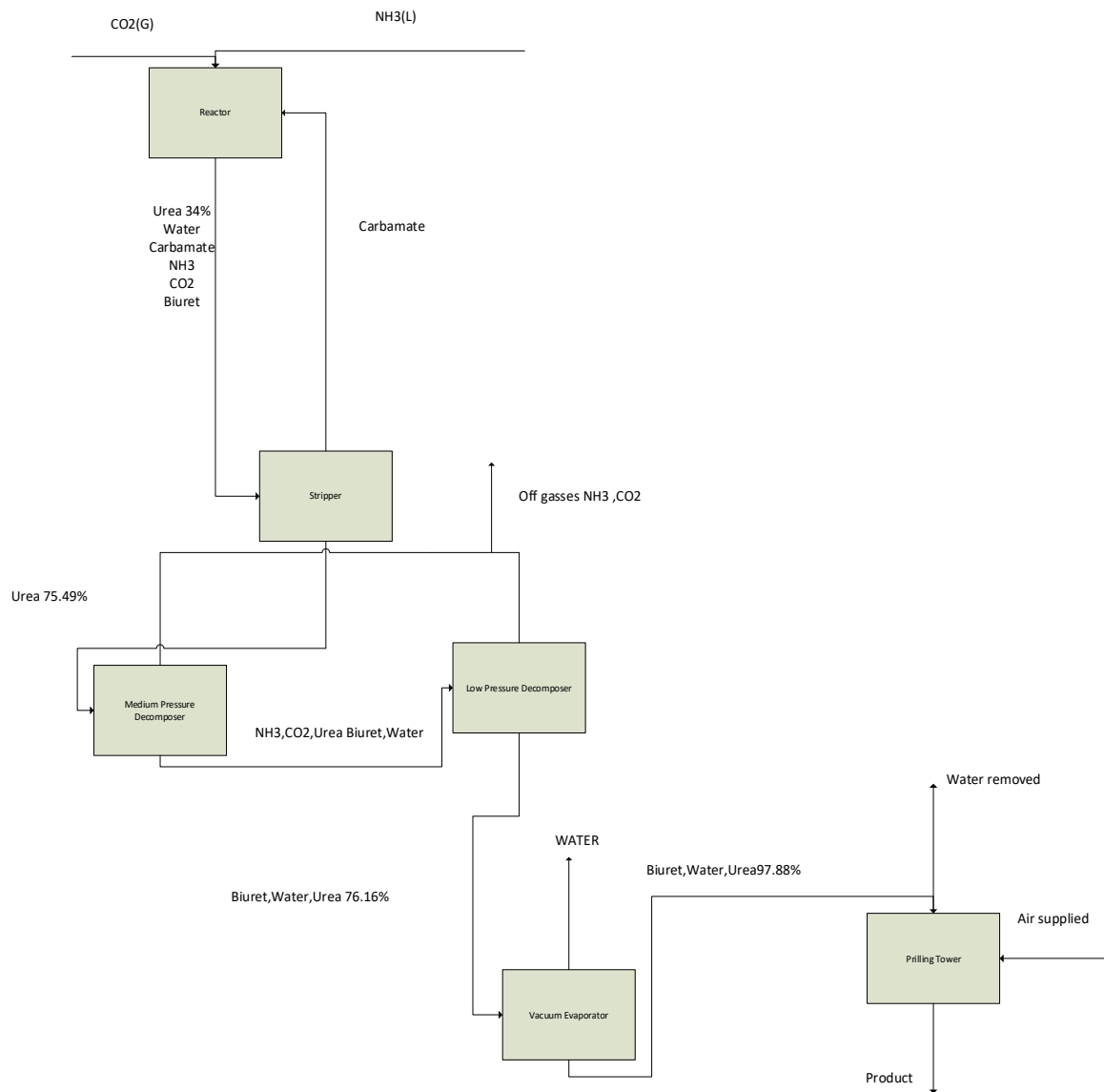
Figure 2.2 Block Diagram of snamprogetti urea process

3.Chapter 3 – Material Balance

Capacity per day   =   1900 MTPD

No. of working days   =   300 days

Capacity per year   =   5700 MTPY

1 Metric ton   =   1000 kg

1 day   =   24 hours

Capacity per day   =   79166.7 Kg/ hr

Composition of final product

Urea   =   98%

          =   79166.7 x 0.98   =   77583.33

Biuret   =   1%

          =   79166.7 x 0.01   =   791.667

Water   =   1%

          =   79166.7 x0.01

urea	=	98%	0.98	77583.33333	kg/hr
biuret	=	1%	0.01	791.6666667	kg/hr
water	=	1%	0.01	791.6666667	kg/hr

Overall conversion to urea is assumed be 98%

$$CO_2 + 2NH_3 \longrightarrow NH_2COONH_4 \qquad \Delta H = -37.4 \text{Kcal/gm mol} \quad \text{Reaction 1}$$

$$NH_2COONH_4 \longrightarrow NH_2CONH_2 + H_2O \qquad \Delta H = +6.3 \text{ Kcal/ gm mol} \quad \text{Reaction 2}$$

Undesirable side reaction take place is
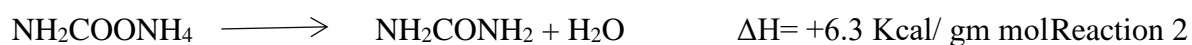
$$2NH_2CONH_2 \longrightarrow NH_2CONHCONH_2 \text{ (Biuret)} + NH_3$$

Molecular weights:

CO_2	44
NH_3	17
NH_2COONH_4	78
NH_2CONH_2	60
$NH_2CONHCONH_2$	103

Urea produced in side reaction $\quad$ = mol. weight of biuret/total molecular weight x biuret produced

$\qquad$ = $\quad$ 120/103 x 791

$\qquad$ = $\quad$ 922.3300971 kg /hr

Total production of urea in reaction 2 $\quad$ = urea produced daily + urea produced in side reaction

$\qquad$ = $\quad$ 77583.33 + 922.3301

$\qquad$ = $\quad$ 78505.66

NH_3 reacted in production of 78505 urea $\qquad$ = mol.weight of N/mol.weight of urea x total production of urea in reacton 2

$\qquad$ = $\quad$ 2 x 14/60 x 78505.66

$\qquad$ = $\quad$ 44486.54

Similarly, CO_2 reacted in the production of urea 78505 $\quad$ = $\quad$ mol.weight of CO_2/mol.weight of urea x total production of urea produced

$\qquad$ = $\quad$ 44/60 x 78505.66

$\qquad$ = $\quad$ 57570.81985 kg/hr

Assuming conversion we get $\quad$ = $\quad$ 98%

NH_3 actually required $\qquad$ = $\quad$ NH_3 reacted in production of urea/ coversion

$\qquad$ = $\quad$ 44486.54/0.98

$\qquad$ = $\quad$ 45394.43 $\quad$ kg/hr

CO_2 actually required $\qquad$ = $\quad$ CO_2 reacted in production of urea/conversion

$$= \quad 57570.72/0.98$$

$$= \quad 58745.73 \quad \text{kg/hr}$$

considering if reaction occurs at 100% then

urea produced from CO_2 at 100% = mol.weight of urea/mol.weight of CO_2 x CO_2 actually required

$$= \quad 60/44 \text{ x } 58745.73$$

$$= \quad 80107.81983 \text{ kg/hr}$$

For 98% urea produced = 80107.81983 x 0.98

$$= \quad 78505.66343 \text{ kg/hr}$$

Urea from NH_3 and biuret = urea produced from CO_2-urea produced in reaction 2

$$= \quad 78505.47\text{-}77583.33$$

$$= \quad 922.3300971 \text{ kg/hr}$$

biuret produced = 1% of total production

$$= \quad 791.6666667 \text{ kg/hr}$$

water produced in reaction 2 = mol.weight of water/mol.weight of urea x total urea produced in reaction 2

$$= \quad 23551.69903 \text{ kg/hr}$$

3.1. Mass balance on reactor (34% urea)
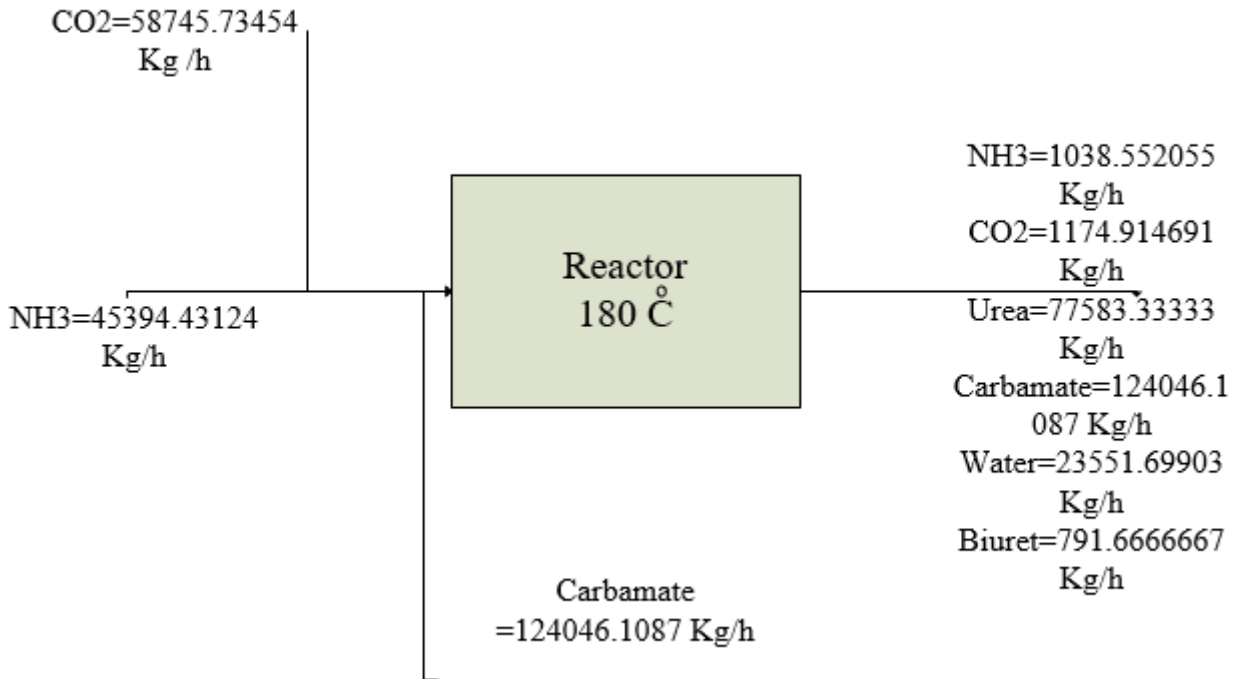
Figure 4 Flow of material across reactor

Flow rate of stream	=	daily urea production/0.34	=	77583.33/0.34

$$= \quad 228186.2745 \text{ kg/hr}$$

NH$_3$ reacted in reaction 1 = mol.weight of NH$_3$/mol.weight of urea x urea produced from CO$_2$

$$= \quad 2*17/60*78505.66$$

$$= \quad 44486.54261 \text{ kg/hr}$$

NH$_3$ produced in reaction 4 = mol.weight of NH$_3$/mol.weight of biuret+NH$_3$ x urea produced form NH$_3$ and biuret

$$= \quad 34/60 \text{ x } 922.33$$

		=	130.6634304 kg/hr

then, NH_3 unreacted　　　　　　　 = 　NH_3 actually required-NH-NH_3 reacted in reaction 1+NH_3 produced in reaction 4

　　　　　　　　　　 = 　1038.552055 kg/hr

CO_2 reacted in reaction 1　　 = 　mol.weight CO_2/mol.weight of urea x urea produced from CO_2

　　　　　　　　　　 = 　57570.81985 kg/hr

CO_2 unreacted　　　　　　　　 = 　CO_2 actually required – CO_2 reacted in reaction 1

　　　　　　　　　　 = 　1174.914691 kg/hr

Now,

Flow rate of carbamate　　　 = 　flow rate of stream at reactors exit – (flow rate of urea + CO_2 + NH_3 + Water+ Biuret)

　　　　　　　　　　 = 　228186.3-(77505.66+1174.91+1038.55+23551.66

　　　　　　　　　　 = 　124046.10 Kg/hr

Mass Balance

Table 3 Flow of material across reactor

Input				Output		
Material	Flow rate	Flow Rate %		Material	flow rate	Flow Rate %
Feed	(Kg/hr)			product	(Kg/hr)	
NH_3	45394.43124	43.58974359		NH_3	1038.552055	0.455133446
CO_2	58745.73454	56.41025641		CO_2	1174.914691	0.514892797

Total	104140.1658	100		Urea	77583.33333	34
Recycled carbamate	124046.1087	100		Carbamate	124046.1087	54.36177483
Total Input	228185.294			Water	23551.69903	10.32126015
				Biuret	791.6666667	0.346938776
				Total	228186.2745	100

3.2.Mass balance on stripper

Carbamate=124046 Kg/h

NH3=1038.552055 Kg/h
CO2=1174.914691 Kg/h
Urea=77583.33333 Kg/h
Carbamate=124046.1 087 Kg/h
Water=23551.69903 Kg/h
Biuret=791.6666667 Kg/h

Stripper
185 Č

NH3=1038.55Kg/h
CO2=1174.915Kg/h
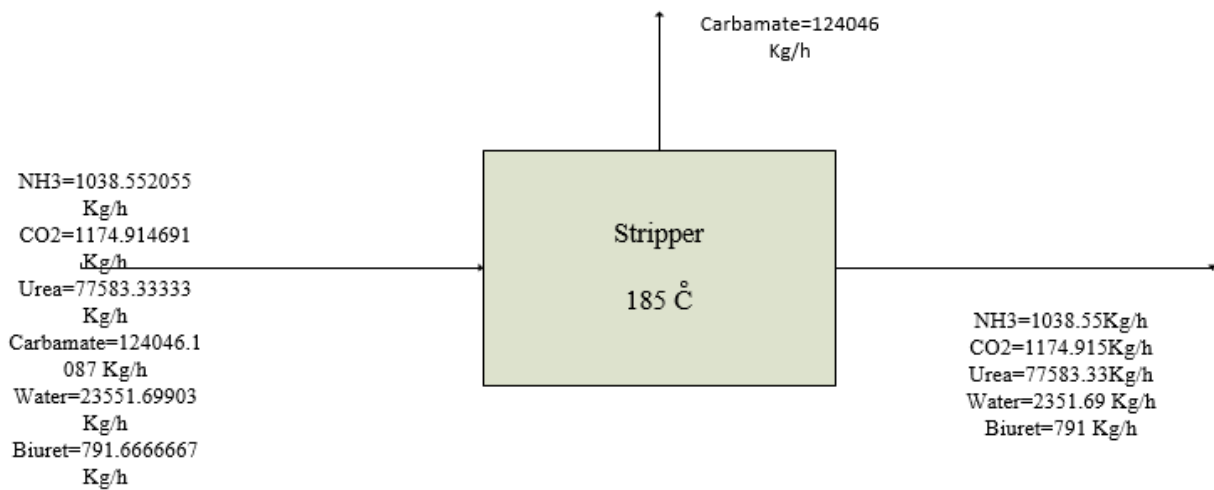Urea=77583.33Kg/h
Water=2351.69 Kg/h
Biuret=791 Kg/h

Figure 5 Flow of material across stripper

Since, no reaction takes place in the stripper & only carbamate gets recycled back to the reactor. Therefore, the amount of ammonia ,carbon-di-oxide ,water & biuret in the outlet stream of stripper will be same as it was in the inlet stream.

Input is the output from the reactor

NH_3 1038.552055

CO$_2$	1174.914691
Urea	77583.33333
Carbamate	124046.1087
Water	23551.69903
Biuret	791.6666667

Mass Balance

Table 4 Flow of material across stripper

Input				Output		
Material	Flow rate	Flow rate %		Material	Flow rate	Flow rate %
Feed	(Kg/hr)			Product	(Kg/hr)	
NH$_3$	1038.552055	0.455133446		carbamate(TOP product)	124046.1087	100
CO$_2$	1174.914691	0.514892797		NH$_3$	1038.552055	0.997263685
Urea	77583.33333	34		CO$_2$	1174.914691	1.128205128
Carbamate	124046.1087	54.36177483		Urea	77583.33333	74.49895317
Water	23551.69903	10.32126015		water	23551.69903	22.61538462
Biuret	791.6666667	0.346938776		Biuret	791.6666667	0.7601934
Total	228186.2745	100		total	104140.1658	100

				Total Output	228186.272	

3.3. Mass balance on medium pressure decomposer

NH3=519.276 Kg/h
CO2=587.4573Kg.h

Medium Pressure Evaporator

140 $\overset{\circ}{C}$

NH3=1038.55Kg/h
CO2=1174.915Kg/h
Urea=77583.33Kg/h
Water=2351.69 Kg/h
Biuret=791 Kg/h

NH3=519.276Kg/h
CO2=587.4573 Kg/h
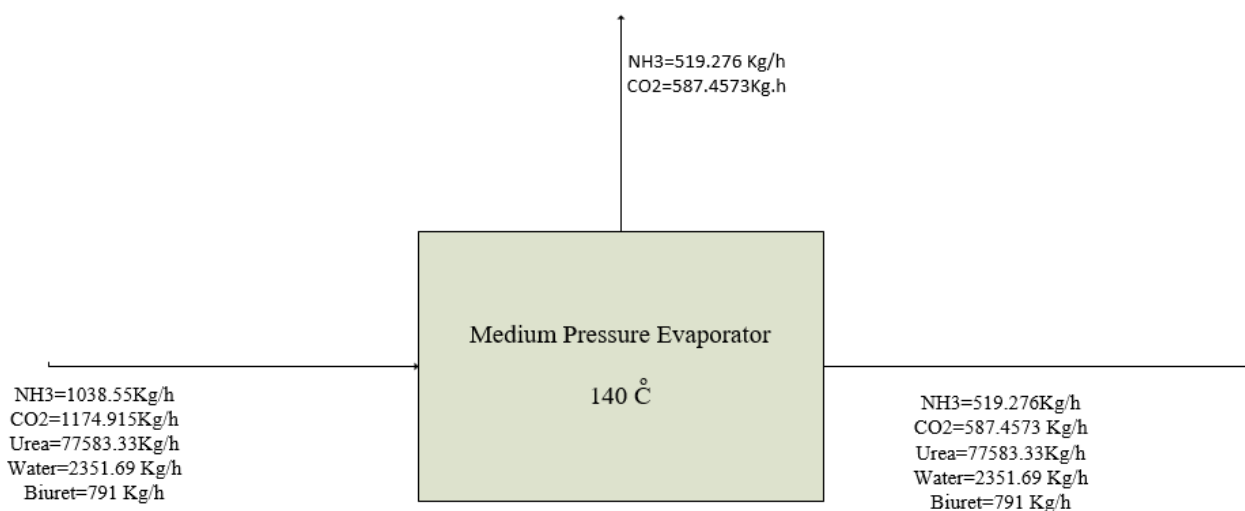Urea=77583.33Kg/h
Water=2351.69 Kg/h
Biuret=791 Kg/h

Figure 6 Flow of material across medium pressure decomposer

The amount of ammonia, carbon-dioxide, water & biuret will remain constant as no reaction is taking place. 50 % of ammonia & carbon-dioxide are assumed to escape from the top of the separator & rest goes with the bottom product. Amount of water & biuret remains constant as no reaction takes place.

Mass Balance

Table 5 Flow of material across medium pressure decomposer

Input				Output		
Material	Flow rate	Flow Rate %				
NH_3	1038.552055	0.997263685		Top product	50	Flow Rate %
CO_2	1174.914691	1.128205128		material	flowrate	%
Urea	77583.33333	74.49895317		NH_3	519.2760276	46.91970444

Water	23551.69903	22.61538462		CO_2	587.4573454	53.08029556
Biuret	791.6666667	0.7601934		Bottom product		
total	104140.1658	100		NH_3	519.2760276	0.503987895
Total top product	1106.73			CO_2	587.4573454	0.57016187
				Urea	77583.33333	75.2991835
				water	23551.69903	22.85830772
				Biuret	791.6666667	0.768359015
				Total	103033.4324	100
				Total Ouput	104140.166	

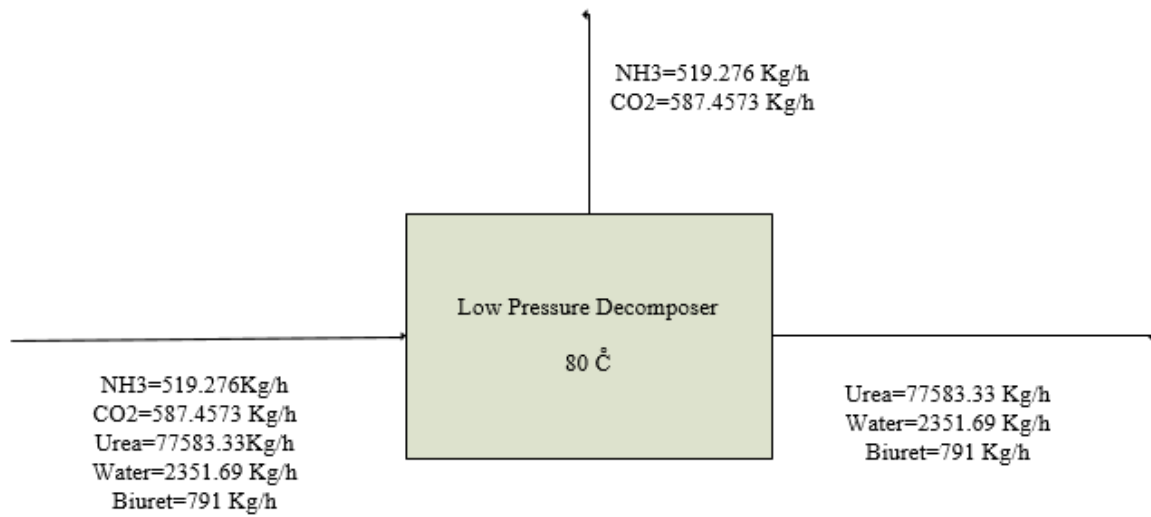## 3.4. Mass balance on low pressure decomposer

Figure 7 Flow of material across low pressure decomposer

Remaining ammonia & carbon-dioxide are assumed to escape from the top.

Mass balance on low pressure decomposer

Table 6 Flow of material across low pressure decomposer

Input				Output		
Material	Flow rate	%		Material	Flowrate	%
feed				top product (50%)		
NH₃	519.2760276	0.503987895		NH₃	519.2760276	46.91970444
CO₂	587.4573454	0.57016187		CO₂	587.4573454	53.08029556
Urea	77583.33333	75.2991835		Bottom product		

water	23551.69903	22.85830772		Urea	77583.33333	76.11679184
Biuret	791.6666667	0.768359015		Water	23551.69903	23.1065062
total	103033.4324	100		Biuret	791.6666667	0.776701958
				total	101926.699	100

Total top product =   1106.73
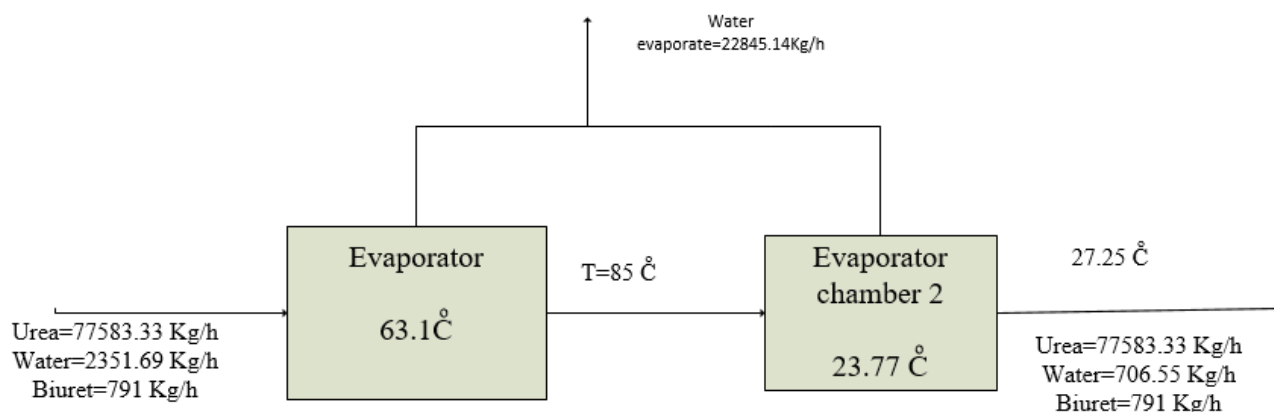
3.5.Mass balance on vacuum evaporator

Figure 8 Flow of material across vacuum evaporator

On this step of the reaction the 97% of water is vaporized.

Mass Balance

Table 7 Flow of material across evaporator

Input

Input	Flow rate	%
Urea	77583.333	76.1167
Water	23551.699	23.106
Biuret	791.667	0.7767
Total	101926.699	100

Output,

assume water evaporation is 97%

Water removed from top   =   22845.148

Bottom products

Output	Flow rate	%
Urea	77583.333	98.105
Water	706.5509	0.8934
Biuret	791.667	1.00107
Total	79081.55	100

3.6. Mass balance on prilling tower

Remove 50 Kg/h water
Air tem =26 Č

27 Č

Urea=77583.33 Kg/h
Water=706.55 Kg/h
Biuret=791 Kg/h

Air
25 Č

Water=656 Kg/h
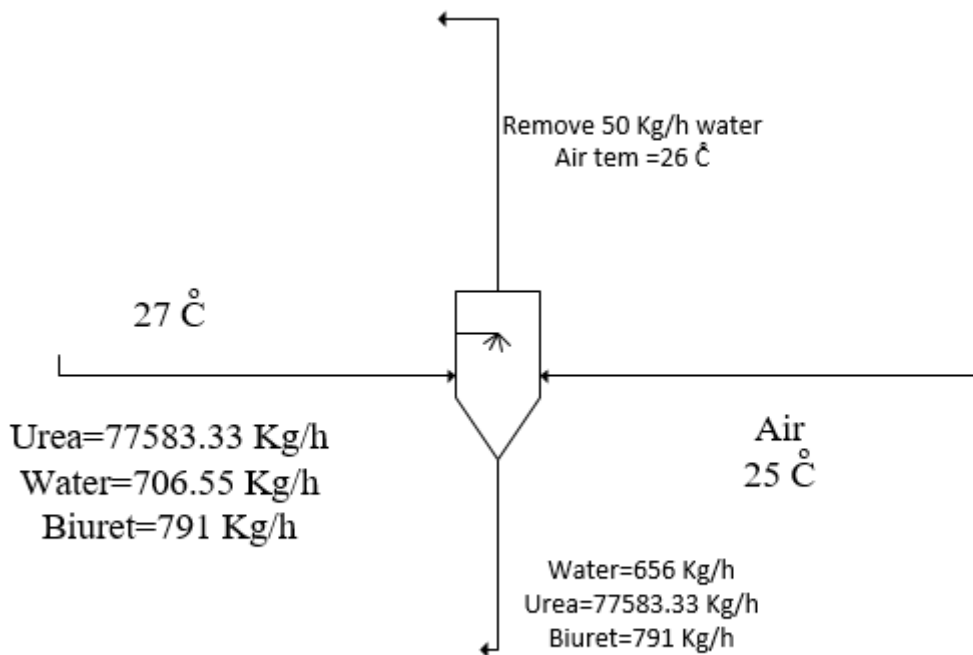Urea=77583.33 Kg/h
Biuret=791 Kg/h

Figure 9 Flow of material across prilling tower

Prilling Towers where Urea prills are formed by passing a current of cold air in the tower from the bottom. Proper size Urea prills are sent to bagging section through belt conveyors. In bagging section , coating of Urea prills may be done if required. Oversized Urea prills or lumps are sent to lump dissolving tank.

Mass Balance

Table 8 Flow of material across prilling tower

Input	Flow Rate	%		Output	Flow Rate	%
Urea	77583.33333	98.10547768		Urea	77583.33333	98.16754496
Water	706.5509709	0.893446021		Water	656.5509709	0.830745396

43

Biuret	791.6666667	1.001076303		Biuret	791.6666667	1.001709642
Total	79081.55097	100		Total	79031.55097	100

4.Chapter 4 – Energy Balance

4.1. Energy balance on reactor
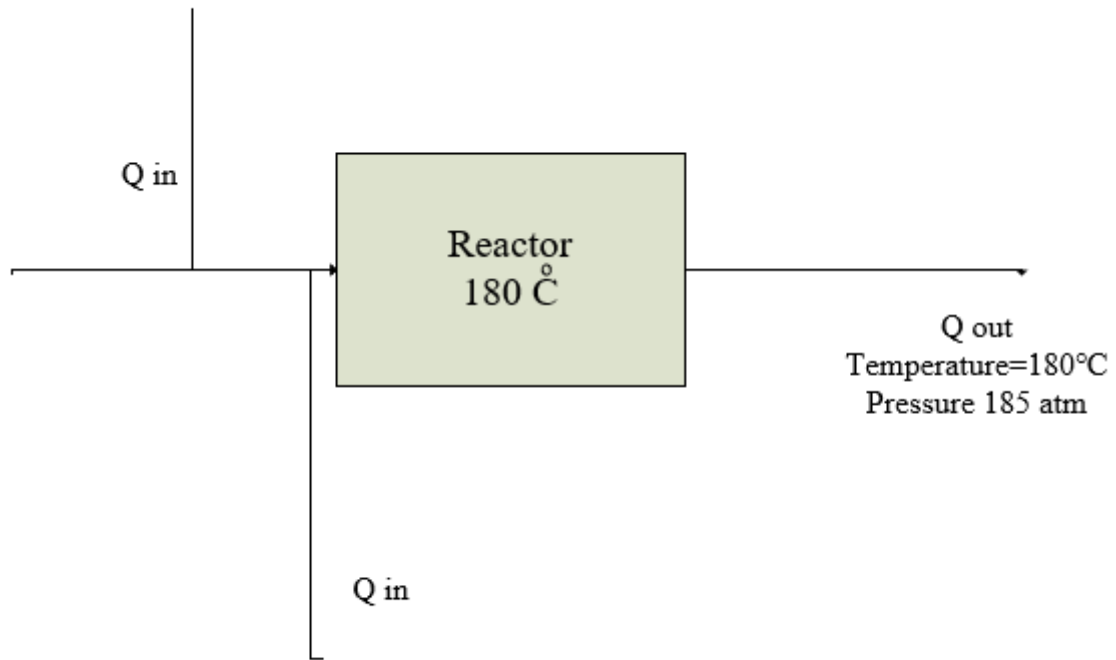
Figure 10 flow rate across reactor

Inlet Stream

Material   Specific heat at 40 °C

NH_3 =   0.53 cal/gm.°C = 2.2119Kj/Kg.°C

CO_2 =   0.22 cal/gm.°C = 0.9211 Kj/kg.°C

                Specific Heat at 180 °C

Carbamate   =   0.62 cal/gm.°C   =   2.596 Kj/kh.°C

Heat Input

$$(mC_p\Delta T)$$

NH_3 = 45394.4 x 2.219Kj/kg x 40

= 0.40292×10^7

CO_2 = 58745.7 x .9211Kj/kg x 40

= 0.21644×10^7

Carbamate = 124046.10 x 2.56Kj/kg x 40

= 5.7964×10^7

Heat input = $NH_3+CO_2+Carbamate$

= $40292 \times 10^7+0.21 \times 10^7+5.7964 \times 10^7$

= 6.4156×10^7 Kj/h

Heat input = 6.4156×10^7 Kj/hr

Enthalpy of Reaction ΔH = -31.32 Kcal gm.mol

ΔH_R = 0.013×10^7 Kj/kmol

Amount of Urea formed during reaction = 1293.05

Now,

ΔHR = 0.013×10^7 Kj/h x 1293.05

ΔHR = 16.81×10^7 Kj/h

Outlet Stream

Materials	Heat Capacity		Mole Fraction	Flow Rate
NH_3	39.15	Kj/Kmol C	0.012822458	61.09129736
CO_2	42.37	Kj/Kmol C	0.005604612	26.70260661
Carbamate	202.49	Kj/Kmol C	0.433934144	2067.435146
Urea	121.32	Kj/Kmol C	0.271399544	1293.055556
Water	75.37	Kj/Kmol C	0.274626009	1308.427724
Biuret	183.8	Kj/Kmol C	0.001613233	7.686084142
Total	**589.13**		**1.00**	**4764.398413**

Cp of mixture = $\sum x_i Cp_i$

= (39.15*0.0128)+(42.37*0.0056)+(202.49*0.433)+(121.34*0.271)+(75.37*0.27)+(183.8*0.00161)

Cp of mixture   =   142.52

Total heat of outlet   =   mCpΔT

                     =   (4764.2) x (142.52) x 180

Total heat outlet   = 12.22×10^7 Kj/h

Rate of heat accumulation   =   heat input + enthalpy of reaction - heat output

Heat accumulation   = $6.41 \times 10^{\wedge}7 + 16.81 \times 10^7 - 12.22 \times 10^7$

Heat accumulation   = 11.00×10^7

Assumption: Cooling water at 25 °C is used to remove heat from reactor. The outlet stream is at an absolute pressure 4.5 bar

Given values are:

Pressure   =   4.5 bar

Temperature of stream   =   147.9 ^{0}C

ΔT = 147.9 – 25 = 122.9 ^{0}C

λ =   2120.6 Kj/Kg

So,

Heat gained by water   =   11.00 x 10^7

$mCp\Delta T + m\lambda$ =   12.22 x 10^7

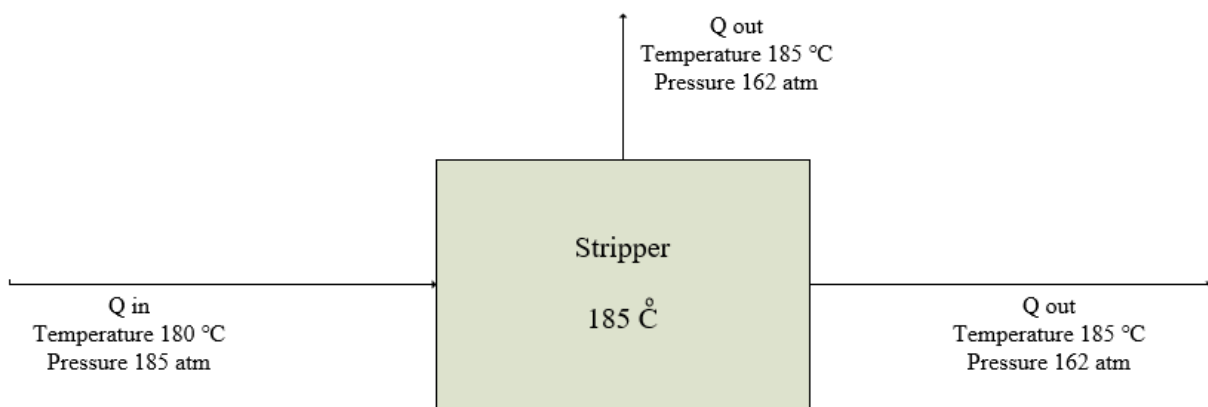m   =   41742.879 Kg/h

4.2. Energy balance on stripper

Figure 11 flow rate across stripper

Input	Flow rate	Output (liquid)	Flow rate
NH$_3$	1038.552055	NH$_3$	1038.55Kg/hr

CO$_2$	1174.914691	CO$_2$	1174.915Kg/hr
Urea	77583.33333	Urea	77583Kg/hr
Carbamate	124046.108	Water	23551.69Kg/hr
water	23551.69903	Biuret	791Kg/hr
biuret	791.6666667		

Output(vapor)

Carbamate   =   124046Kg/h

Heat at input   =   12.22 x 10^7 Kj/h

Now for heat output

Materials	specific Heat at 185C	Mole Fraction	Flow Rate
NH$_3$	41.31KJ/Kmol C	0.022651945	61.09117647
CO$_2$	44.22KJ/Kmol C	0.009901039	26.70261364
Urea	135.3KJ/Kmol C	0.479448896	1293.05
Water	75.37KJ/Kmol C	0.485150603	1308.427222
Biuret	183.8KJ/Kmol C	0.002847517	7.67961165
Total		1	2696.950624

Cp of outlet mixture   = $\sum x_i Cp_i$

Cp of mixture   =
(41.31*0.022)+(44.22*0.0099)+(135.3*0.4794)+(75.37*0.4851)+(183.8*0.0028)

Cp of mixture   =   103.3321Kj/Kmol.°C

Now,

Total heat by outlet = $mCp\Delta T$

Total heat by outlet = 2696.95 x 103.33 x 185

= 5.15×10^7

Now for Vapor phase output

Materials	Specific heat	flow rate at 185°C
Carbamate	209.49 Kj/kg.°C	2067.4333

For carbamate λ = 210 Kj/kg. ℃

Heat is $mCP\Delta T$ + m (flow in rate of carbamate) λ

Heat = 10.34×10^7 Kj/hr

Assumption : Cooling water at 25 ^{0}C is used to remove heat from the reactor. The outlet is steam at an absolute pressure of 24 atm (Ts = 221.8 °C).
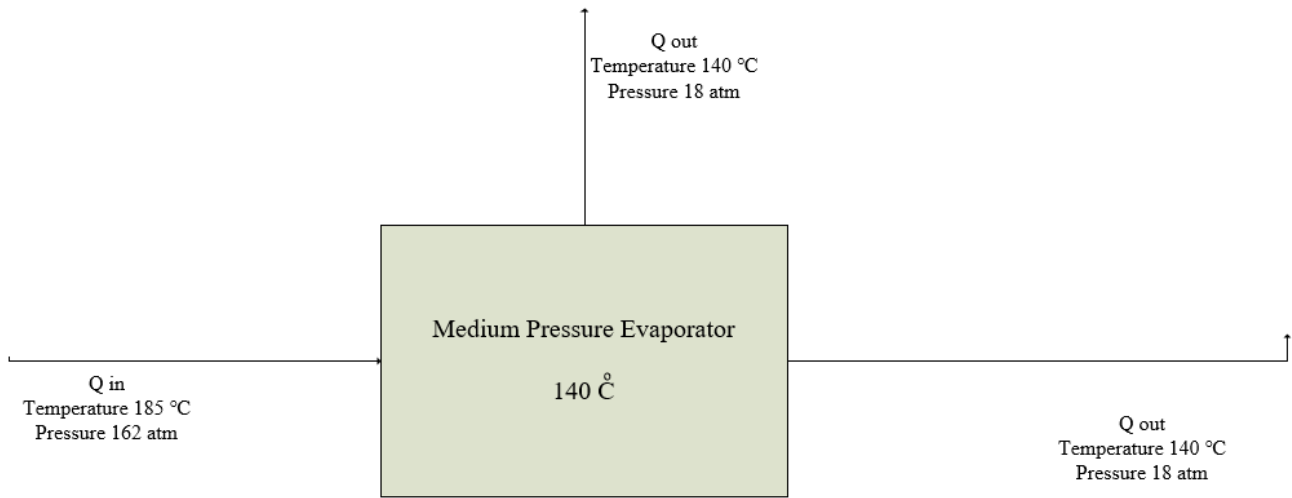
λ of stream = 1855.3 Kj/hr

heat supplied by steam = heat output - heat input

m λ = $10.37 \times 10^7 + 5.15 \times 10^7 - 12.22 \times 10^7$

m = 17691.10549 kg/hr

4.3 Medium Pressure Separator

Heat input $= 5.15 \times 10^7$

1) liquid

Outlet Stream

Material	Specific heat [140^0C]	Mol. Fraction	Flow rate
NH$_3$	38.4 Kj/Kmol ^{0}C	0.11516	30.54
CO$_2$	42.37 Kj/Kmol ^{0}C	0.00503	13.3513
Urea	123.84 Kj/Kmol ^{0}C	0.487	1291.76
Water	75.37 Kj/Kmol ^{0}C	0,4934	1308.42
Biuret	170.92 Kj/Kmol ^{0}C	0.00289	7.673
			Total = 2651.7449

Cp of mixture = $\sum x_i Cp_i$

So, Cp = (0.11516 x 38.4) + (0.00503 x 42.37) + (0.487 x 123.84) + (0.4934 x 170.92) + (0.00289 x75.37) Kj/Kmol°C

Cp = 102.624 Kj/Kmol ^{0}C

heat output   = 2652.7449 x 102.624 x140

= 3.8098 x 10^7 Kj/hr

2) **For gases escaping from the top**

Material	λ at 140 ^{0}C	Mol.Fraction	Flow rate
NH$_3$	22.77x 10^3 Kj/Kmol ^{0}C	0.6958	30.54
CO$_2$	20.265x 10^3 Kj/Kmol ^{0}C	0.3041	13.3513
			Total = 43.8913

λ of mixture = $\sum$ xiλi

So, λ = (0.6958 x 22.777 + 0.3041 x 20.265) x 10^3 Kj/Kmol ^{0}C

= 22.0108 x 103 Kj/Kmol ^{0}C

NH$_3$ specific heat at 140 ^{0}C = 38.4 Kj/Kmol ^{0}C

CO$_2$ specific heat at 140 ^{0}C = 42.37

Similarly Cp of gas mixture, Cp = (0.6958 x 38.4) + (0.3041 x 42.37)

$$= 39.603 \ Kj/Kmol \ ^{0}C$$

Heat escaping from the top $= m \ (Cp\Delta t + \lambda)$

$$= 43.89 \ (39.603 \times 140 + 22.0108 \times 10^{3})$$

$$= 0.12093 \times 10^{7} \ Kj/hr$$

Assumption : Cooling water enters at 25 ^{0}C & leaves at 50 ^{0}C.

So , heat gained by cooling water = heat input – heat output
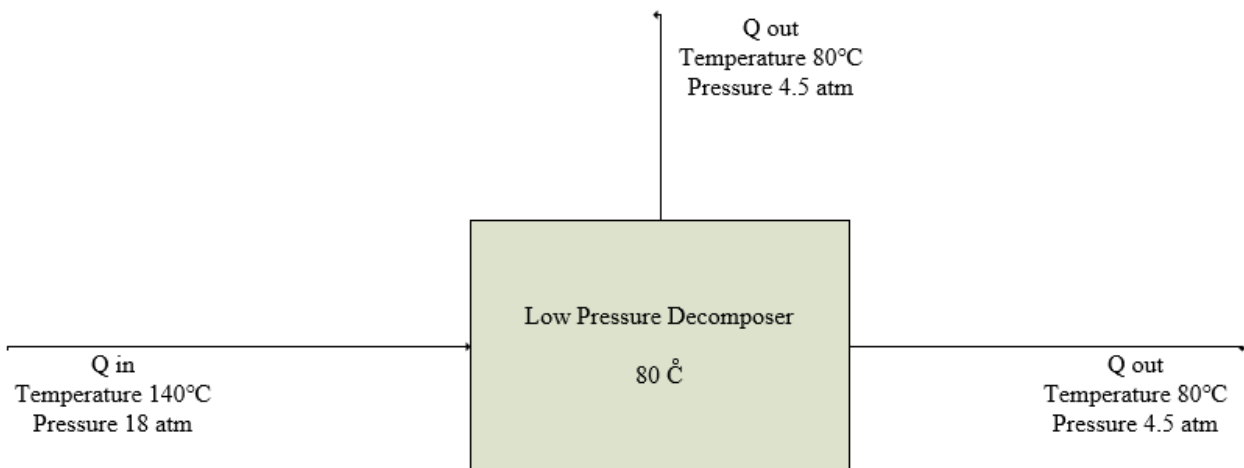
$$= (5.15 - 3.8098 - 0.12093) \times 10^{7} \ Kj/hr$$

$$mCp\Delta t \ = 1.219 \times 10^{7} \ Kj/hr$$

$$m = 1.219 \times 10^{7} \ / \ (\ 4.187 \times 25)$$

$$m = \ 116483.5 \ Kg/hr$$

4.4 Low Pressure Separator

Heat Input $= \ 3.8098 \times 10^{7} \ Kj/hr$

1) liquid

Outlet Seram

Material	Specific Heat at 80 °C	Mol fraction	Flow rate
Urea	107.76 Kj/Kmol °C	0.4953	1291.76
Water	75.37 Kj/Kmol °C	0.5017	1308.42
Biuret	149 Kj/Kmol °C	0.0029	7.671
			Total = 2607.85

Cp of mixture = $\sum$ xiCpi

So, Cp   = (0.5017 x 107.76) + (0.5 x 75.37) + (0.476 x 123.84) + (0.004 x 149) Kj/KmoloC

   =   91.62 Kj/Kmol oC

heat output = 2607.85 x 91.63 x 80

      = 1.9114 x 10^7 Kj/hr

3) For gases from top

Material	λ at 140°C	Mol fraction	Flow rate
NH_3	18.51 x 10^3	0.695	30.545
CO_2	15.66 x 10^3	0.304	13.35
Total			43.896

λ of mixture = $\sum x_i \lambda_i$

So, λ = (0.695 x 18.51 + 0.304 x 15.66) x 103 Kj/Kmol ^{0}C

       = 17.625 x 10^3 Kj/Kmol

Similarly,

Material	Specific heat at 140 ^{0}C	Mol fraction	Flow rate
NH$_3$	37.013 Kj/Kmol ^{0}C	0.695	30.545
CO$_2$	38.69 Kj/Kmol ^{0}C	0.304	13.35
Total			43.896

Cp of mixture = $\sum x_i Cp_i$

So, Cp = 0.695 x 37.013 + 0.304 x 38.69 Kj/Kmol^{o}C
       = 37.48 Kj/Kmol^{o}C

Heat escaping from the top = m (CpΔt + λ)

                 = 84.59 (37.48 x 80 + 17.625 x 10^3)

                 = 0.09528 x 10^7 Kj/hr

Assumption : Cooling water enters at 25 0C & leaves at 50 ^{0}C.

So , heat gained by cooling water = Heat input – heat output

                 = (3.9098 – 1.9114 – 0.09528) 10^7

mCpΔt   = 1.903 x 10^7

m   = 1.903 x 10^7 / 4.187 x 25

m   = 181809.49 Kg/hr

4.5 Evaporator
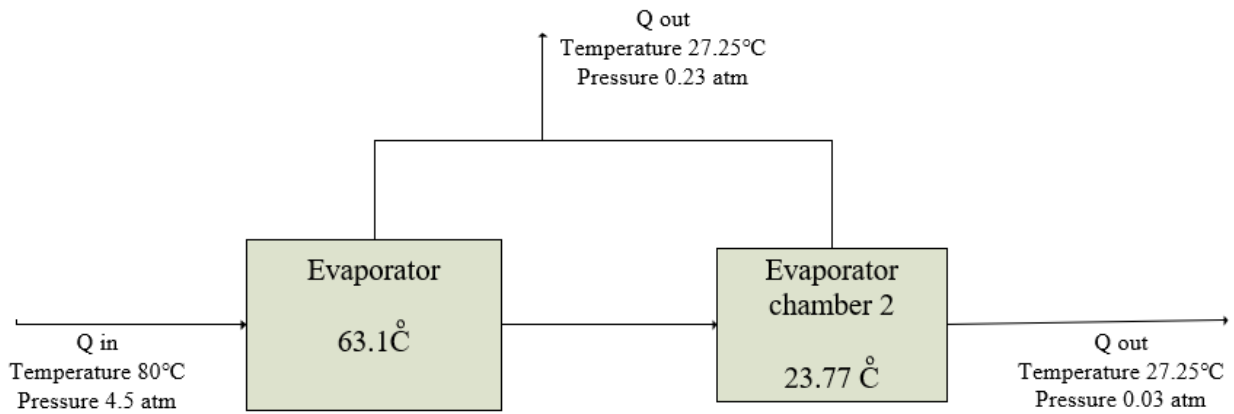
Product steam coming out of evaporator

Material	Specific heat at 85 ^{0}C	Mol fraction	Flow rate
Urea	109.28 Kj/Kmol ^{0}C	0.964	1291.76
Water	75.37 Kj/Kmol ^{0}C	0.00292	39.22
Biuret	149 Kj/Kmol ^{0}C	0.00573	7.673
Total			1338.65

Cp of mixture = $\sum$ xiCpi

So, Cp = 0.964 x 109.28 + 0.00292 x 75.37 + 0.00573 x 149 Kj/KmoloC

= 106.41 Kj/Kmol ^{0}C

mCpΔt = 1338.65 x 106.41 x 85

$= 1.210$ x 10^7 Kj/hr

Heat balance on 1^s evaporator:

Heat input (feed) + Heat input by steam = heat carried by water vapour + energy of the bottom product

Heat input (feed) + S_1 λs_1 = $E_1 H_{E1}$ + energy of the bottom product

1.537 x10^7 + S1 x 2123.2 = 12,593 x 2614.97 + 1.210 x 10^7

$S1 = 13969.6$ Kg/hr

2nd evaporator:

Heat input (feed) + Heat input by steam = heat carried by water vapour + energy of the bottom product

Heat input (feed) + S_1 λs_1 = $E_1 H_{E1}$ + energy of the bottom product

1.210 x10^7 + S1 x 2123.2 = 5300 x 2545.7 + 1065.8 x 96.84 x 27

$S1 = 1969.7$ Kg/hr

4.6 Prilling tower

Heat input = 0.279 x 10^7 Kj/hr

Outlet stream

Material	Specific heat at 30^0C	Mol fraction	Flow rate

Urea	94.41 Kj/Kmol ^{0}C	0.966	1291.76
Water	75.37 Kj/Kmol ^{0}C	0.0272	36.44
Biuret	133.02 Kj/Kmol ^{0}C	0.00574	7.67
Total			1335.87

Cp of mixture = $\sum$ xiCpi

So, Cp = 0.966 x 94.41 + 0.0272 x 75.37 + 0.00574 x 133.02 Kj/Kmol ^{0}C

= 94.01 Kj/Kmol ^{0}C

Heat output = 1335.89 x 94.01 x 25

= 0.3139 x 10^7 Kj/hr

Assuming, humidity of air at 25 ^{0}C = 0.01

Heat carried away by air = heat input – heat output

(mCpΔt)$_{dry\ air}$ = (0.279 – 0.3139) x 10^7

m = .029 x 107 / (1.009 x 1)

m = 345887 Kg/hr

So, flow rate of air   =   345887 Kg/hr

5. Chapter 5 – Plant Design

5.1. Reactor Design

Reactor is the core of the synthetic or chemical reaction process. Synthetic reactors are vessels that are intended for a substance response to happen within them. The presentation of the reactor impacts by and large execution of the entire urea plant underway limit and utilities utilization.

Reactor Selection Criteria:

- ➢ Operating conditions (Temperature and pressure)

- ➢ Feed condition and phase

- ➢ Rector types (batch, continuous, plug flow)

- ➢ Catalyst

Table 9 Comparison of reactors

| Batch Reactors	CSTR	Plug Flow Reactor
Used in comparatively small-scale operation. ➢ For testing new process that have not been fully developed. ➢ Manufacturing of expensive products. ➢ Pharmaceutical industries.	➢ Used mostly for liquid phase reactions. ➢ Used in organic industries	➢ Normally used for gas phase reactions. ➢ High pressure conditions. ➢ No back mixing required due to reversible reactions. ➢ Careful control of residence time.
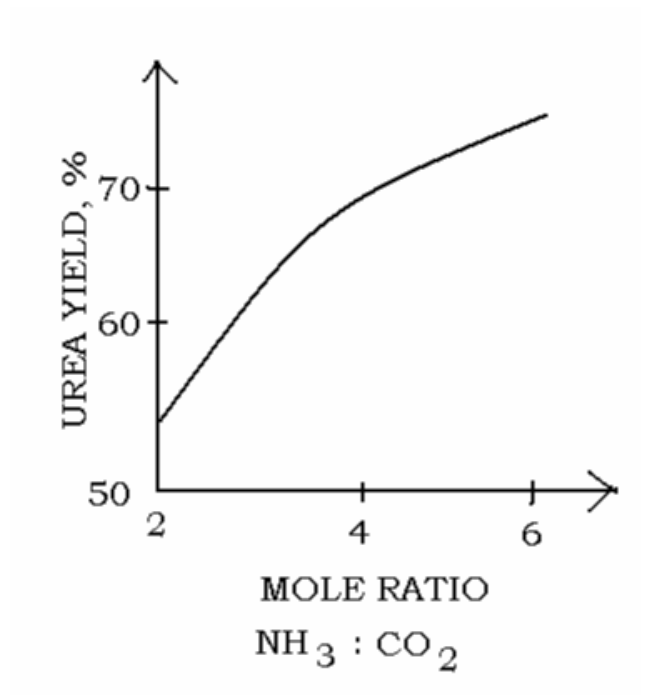
Figure 12 Graph of % urea yield Vs molar ratio of NH₃ Vs CO₂
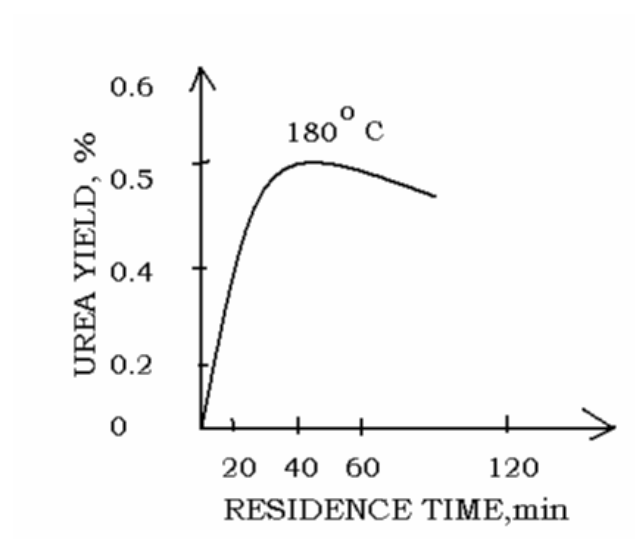
Figure 13 Graph of % urea yield Vs residence time.

Now, t = V/F   (Ref: Chemical Reactor Design Harriot, Pg-90)

Where   t = residence time

F = Volumetric flow rate into the reactor in m^3/hr

V = Volume of the reactor m^3

Now,

Density of liquid NH_3 =   618 Kg/m^3

Density of CO_2 gas at 40^oC = 277.38 Kg/m^3 (density=PM/RT; P=162atm, T=313K)

So.

Volumetric flow rate of NH_3 = 45394.43124 / 618   = 73.4537 m^3/hr

Volumetric flow rate of CO_2 =58745.73454 / 277.38   = 211.7879 m^3/hr

Volumetric flow rate of carbamate = 124046.1087 / 1600   =77.5288 m^3/hr

Total Volumetric flow rate in the reactor   = 362.7704 m^3/hr

Since, t = V/F

Where, t = residence time

V = Volume of the reactor

F = Volumetric flow rate

Then V = t x F

= (40 x 362.7704) / 60

V = 241.8 m^3

Giving 10% allowance   =   241.8 + 10%

=   265 m^3

Length & diameter

L/D   =   10 (for high pressure vessel)

Reactor is cylindrical, volume of cylinder = $\pi/4 \, D^2 \times L$

L   =   10D

Then,   V   = $\pi/4 \, D^2 \times 10D$

D   =   3.23 m

Then,   L   =   32.3 m

5.1.1 Thickness of reactor wall

Temperature inside the reactor = $180^0 C$

Pressure inside the reactor = 125 atm

Design pressure = 1.393 x 10^7 N/m^2

Thickness is given by, $T_h = \dfrac{PD}{2fj-P} + C$

Where,

T_h =   thickness

P   =   Design Pressure   =   1.393 x10^7 N/m^2

D   =   Diameter of reactor   =   3.23 m

f   =   Maximum allowance stress   =

C   =   Corrosion allowance   =   2 mm

J   =   Joint Factor   =   1 (for HP vessel)

Then, T_h = $\dfrac{(1.393*10\text{\textasciicircum}7)*(3.23)}{(2*1*180*10\text{\textasciicircum}8)-1.393*10\text{\textasciicircum}7}$

Allowable Stress = 1.8×10^8 N/m^2

Now, to calculate thickness we use:

So, t = $1.393 \times 10^7 \times 3.23/ ((2 \times 1.8 \times 10^8 \times 1) - 1.393 \times 10^7)) + 2mm = 0.196$ m

   = 196 mm

or, t = 200 mm

5.1.2 Head Design

For 2: 1 ellipsoidal dished head

T_h = p DV/2fJ   [ref: Equipment design- M V Joshi, Pg-106, Eq-5.24]

where, p = internal design pressure

       D = major axis of ellipse

       V = stress intensification factor = $(2 + k^2)/4$ k = major axis/minor axis

So, t_h = 1.716 x107 x 2.5 x 1.5 / (2 x 1.18 x108 x 1) t = 0.273 m

or, t = 273 mm

or, t = 300 mm

5.1.3 Diameter of pipes

We know that,

$$(D_i)_{opt} = 0.0144 \times (m`)^{0.45}/(\rho)^{0.32}$$

For inlet pipes:

$$(D_i)_{NH_3} = 0.0144 \times (36968)^{0.45}/(618)^{0.32}$$

$$= 0.2093 \text{ m}$$

$$= 8.24 \text{ inch}$$

Standardizing using Table-11, PHT, D Q Kern we get:

$$NPS = 10$$

$$\text{Schedule no.} = 60$$

$$OD = 10.75 \text{ inch}$$

$$ID = 9.75 \text{ inch}$$

$$(D_i)_{CO_2} = 0.0144 \times (47842)^{0.45}/(277.38)^{0.32}$$

$$= 0.3037 \text{ m}$$

$$= 11.95 \text{ inch}$$

Standardizing using Table-11,PHT,D Q Kern we get:

$$NPS = 12$$

$$\text{Schedule no.} = 30$$

$$OD = 12.75 \text{ inch}$$

ID = 12.09 inch

$(Di)_{carbamate} = 0.0144 \times (95337)^{0.45}/(1600)^{0.32}$

$= 0.2364$ m

$= 9.307$ inch

Standardizing using Table-11, PHT, D Q Kern we get:

NPS = 10

Schedule no. = 60

OD = 10.75 inch

ID = 9.75 inch

$(Di)_{outlet\ stream} = 0.0144 \times (1,80,147)^{0.45}/(1283.97)^{0.32}$

$= 0.3378$ m

$= 13.29$ inch

Standardizing using Table-11, PHT, D Q Kern we get:

NPS = 16

Schedule no. = 30

OD = 16 inches

ID = 15.25 inch

5.2 Evaporator Design

Vapor space pressure = 0.23 atm

Vapor space temperature = 63.1 ^{0}C

BPR = 21.9 ^{0}C

Boiling point of liquid = 85 ^{0}C

From above energy balance we know:

Material	Specific heat at 85 ^{0}C	Mol fraction	Flow rate
Urea	109.28 Kj/Kmol ^{0}C	0.964	1291.76
Water	75.37 Kj/Kmol ^{0}C	0.00292	39.22
Biuret	149 Kj/Kmol ^{0}C	0.00573	7.673
Total			1338.65

Cp of mixture = $\sum$ xiCpi

So, Cp = 0.964 x 109.28 + 0.00292 x 75.37 + 0.00573 x 149 Kj/KmoloC

$\qquad$ = 106.41 Kj/Kmol ^{0}C

mCpΔt = 1338.65 x 106.41 x 85

$\qquad$ = 1.210 x 10^7 Kj/hr

Heat balance on 1st evaporator:

Heat input (feed) + Heat input by steam = heat carried by water vapour + energy of the bottom product

Heat input (feed) + $S_1 \lambda s_1$ = $E_1 H_{E1}$ + energy of the bottom product

$1.537 \times 10^7 + S1 \times 2123.2$ = $12,593 \times 2614.97 + 1.210 \times 10^7$

$$S1 = 21208.7 \text{ Kg/hr}$$

Steam Economy = E / S

Steam economy = 12593 / 21208.7

$$= 0.5937$$

Now we will get UI value from fig-6.2.2.At 63.1oC (145.58 ^{0}F) the value of U1 is 270 Btu/hr.sq.ft. ^{0}F. Multiplying this value by 5.6783 gives the value of U1 in W/m^{2}K.

$A_1 = S_1 \lambda_{s1} / U_1 \Delta T_1$

First $\Delta T_1 = (\Delta T)app - BPR_1$

$= 147.165 - 63.1 - 21.9$

$= 62.165\ ^oC$

So, $A_1 = 13969.6 \times 2123.2 / 1533 \times 62.165$

$= 311.2\ m^2$

Similarly,

$A_2 = S_2 \lambda_{s2} / U_2 \Delta T_2$

$\Delta T_2 = (\Delta T)app - BPR_2$

$= 147.165 - 23.77 - 3.48$

$\Delta T_2 = 119.915\ ^oC$

So, $A_2 = 1969.7 \times 2123.2 / 738 \times 119.915$

$= 47.25\ m^2$

(Ref: values of U_1 & U_2 from Perry's handbook)

5.2.1 Design

Length = 6m

Tube OD = 1 inch

Tube ID = 0.834 inch

Minimum pitch = 1.25 x OD

$= 31.25\ mm$

Let the pitch = 32 mm square pitch

Area = 311.2 m^2

No. of tubes N:

$A = \Pi \times 0.025 \times 6 \times N$

311.2 = 3.14 x 0.025 x 6 x N

N = 660.7

Let OTL = D

We know, $(\Pi/4) \times D^2 = N \times (pitch)^2$

So, $(\Pi/4) \times D^2 = 660 \times (0.032)^2$

D = 927 cm = 0.927m

Now,   Ddi = OTL + 2C

        = 0.927 + 2 x 0.075

        Ddi = 1.077 m

Standardizing D, = 1100 mm

5.2.2 Wall thickness Calculation

Material of construction: Mild steel

Formula for wall thickness, $t = p_d D_i / (2fj - p) + C$

where, t = thickness of the shell

        D_i = internal diameter

        J = joint efficiency

        pd= design pressure

f = permissible stress

C=Corrosion allowance

pd = 1.1 x ps

ps = 4.5 atm = 4.413 bar

pd = 4.854 x 10^5 N/m^2

t = (4.854 x 10^5 x 1.1) / (2 x 0.93 x 10^8 x 0.85 – 4.854x 10^5)

=   3.3876 mm

To calculate critical external bucking pressure,

Using, Pc = [2.42E / (1-μ^2)$^{3/4}$] x [(t/Do)$^{5/2}$/ { L/Do – 0.45 x (t/Do)$^{1/2}$}]

5.3 Stripper

Top head (Elliptical head)

For 2 : 1 ellipoisdal dished head

D_i = 1 m

L = 4m

T_h = pDV/2fJ

  where, p = internal design pressure

    D = major axis of ellipse **Graph to find out heat transfer co-efficient**

    V = stress intensification factor = $(2 + k^2)/4$

    k = major axis/minor axis

    p = 0.23 ata = 0.226 x 105 N/m^2

    j = 0.85

    Di = 1.6 m

    k = 2

    V = 1.5

For internal pressure :

th = (0.226 x 105 x 1.5 x 1) / (2 x 0.93 x108 x 0.85)

   = 2.144 x 10^{-4} m

   = 0.214 mm

For external pressure:

P_{ext} = 1 Kg/cm2

Corresponding internal pressure to be used to calculate th = 1.67 x P_{ext}

So, P_{int} = 1.67 Kg/cm2

So,

$$t_h = 4.4 \times Rc[3 \times (1 - \mu^2)]^{1/2} \times (p/2E)^{1/2}$$

where, p = Design external pressure

Rc = Crown radius for tori spherical & hemispherical heads and equivalent crown radius for elliptical head.

E = modulus of elasticity

μ = Poisson's ratio

Putting the values we get,

$t_h = 0.0061$ mm

So, $t_h =$

5.3.1 Bottom head design

Assuming an apex angle of 90^0

For , conical head

$D = 1$ m

$t_h = pDV/2fJ\cos \alpha$

Here, $\alpha =$ Half the apex angle

For, internal pressure :

$p = 0.226 \times 105$

$t_h = (0.226 \times 105 \times 1.5 \times 1) / (2 \times 0.93 \times 108 \times 0.85 \times 0.707)$

$= 0.3032$ mm

For, external pressure :

$$p = 1.67 \times P_{ext}$$

$$= 1.67 \times 1 \ kg/cm2$$

So,

$$t_h = (1.67 \times 10^5 \times 1.5 \times 1) / (2 \times 0.93 \times 10^8 \times 0.85 \times 0.707)$$

$$= 2.241 \ mm$$

6.Chapter – 6 Instrumentation and Control

6.1 Overview

A synthetic plant is blend of many incorporated units which are answerable for changing over crude materials into valuable items. During it's anything but a plant needs to fulfill many working conditions and prerequisites determined by fashioner. Such necessities request consistent checking framework for a compound plant. This framework ensures the fulfillment of operational destinations. Framework is made out of types of gear and accommodating help and known as instrumentation and cycle control.

6.1.1 Safer Control Operation

- ➢ To keep the interaction factors inside known safe working cutoff points.
- ➢ To recognize risky circumstances as they create and to give cautions and programmed shut-down frameworks.
- ➢ To give entomb bolts and alerts to forestall perilous working methods.

6.1.2 Production rate & Quality Control

- ➢ To accomplish the plan item yield.
- ➢ To keep up the item organization inside the predetermined quality guidelines.

6.1.3 Cost

- ➢ To work at the most reduced creation cost, similar with the other.

6.1.4 Objectives

These are not independent targets and should be thought about together. The request in which they are recorded isn't intended to suggest the priority of any goal over another, other than that of putting wellbeing first. Item quality, creation rate and the expense of creation will be subject to deals necessities. For instance, it could be a superior system to deliver a superior quality item at a greater expense.

In a run of the mill substance handling plant these destinations are accomplished by a mix of programmed control, manual observing and research facility examination.

6.2 Components of Control System Process

Cycle is characterized as arrangement of steps which are utilized to deliver last wanted item from accessible crude materials. In this conversation the cycle is blend of urea.

6.2.1 Control

Control is characterized as utilization of procedures for fixing the given boundaries of interaction around required worth. Now and then we likewise use term measure control for same implications Boundary which is to be kept fixed is estimated by some yield factors going from temperature to time and distance estimations. This sign is sent to a gadget which analyzes the two qualities and sends mistake to definite control component to make a move. There are three fundamental pieces of a control framework [15].

- ➢ Sensor (sensor, transducer and transmitter)
- ➢ Controller (processor, memory and adding circuit)
- ➢ Final control element (actuator, power control circuit)

6.2.2 Control Loop

A control circle is an interaction the board situation which is intended to control input variable around set point. A control circle is supposed to be made of three or four stages recorded underneath:

(i)   sense

(ii) compare

(iii) respond

(iv) action Control circles utilized in enterprises are essentially of two sorts:

- ➢ Feed forward control loop
- ➢ Feed backward control loop

Feed forward control circle estimates unsettling influences and controls the controlled variable straightforwardly. While in feed in reverse control circle, we measure controlled variable and control the unsettling influences.

6.2.3 Classification of Controllers

Overall the cycle regulators can be delegated:

- ➢ Pneumatic regulators
- ➢ Electronic regulators
- ➢ Hydraulic regulators

In the urea producing, the regulator and the last control component might be pneumatically worked because of the accompanying reasons.

6.3 Control system

Estimation is characterized as the extraction from physical and compound frameworks or cycles of signs, which address boundaries or variable. The exhibition of a robotization framework can never outperform that the related estimating gadgets. An essential model is an individual.

6.3.1 Temperature measurement and control

Temperature estimation is utilized to control the temperature of outlet and bay streams in heat exchangers, reactors, and so forth most temperature estimations in the business are made through thermo-couples to work with carrying the estimations to concentrated area. For neighborhood estimations at the hardware bi-metallic or filled framework thermometers are utilized less significantly. Typically, for high estimation precision, obstruction thermometers are utilized.

6.3.2 Pressure measurement and control

Like temperature, pressure is a significant sign of material state and sythesis. Indeed, these two estimations considered together are the essential assessing gadgets of modern materials. Siphons, blower and other cycle hardware related with pressure changes

In the process material are outfitted with pressure estimating gadgets. Subsequently pressure estimation turns into a sign of energy increment or reduction.

Most pressing factor estimation in industry is flexible component gadgets, either straightforwardly associated for nearby use or transmission type to brought together area. Most broadly utilized mechanical pressing factor component is the bourden tube or a diaphragm or bellows measures.

6.3.3 Level measurement

A sensor estimates a degree of responding liquids inside pfr and signs to regulator. Onthe off chance that level is beneath or over the given level sign is sent towards the leave stream control component at urea arrangement. Accordingly, it chokes between open-shut circumstances.

6.3.4 Composition measurement

A sensor for estimating piece is likewise brought into framework as higher n/c proportion will cause energy load on plant. Likewise, lower proportion will cause less change driving towards in-monetary interaction.

6.4 Control Loop on PFR
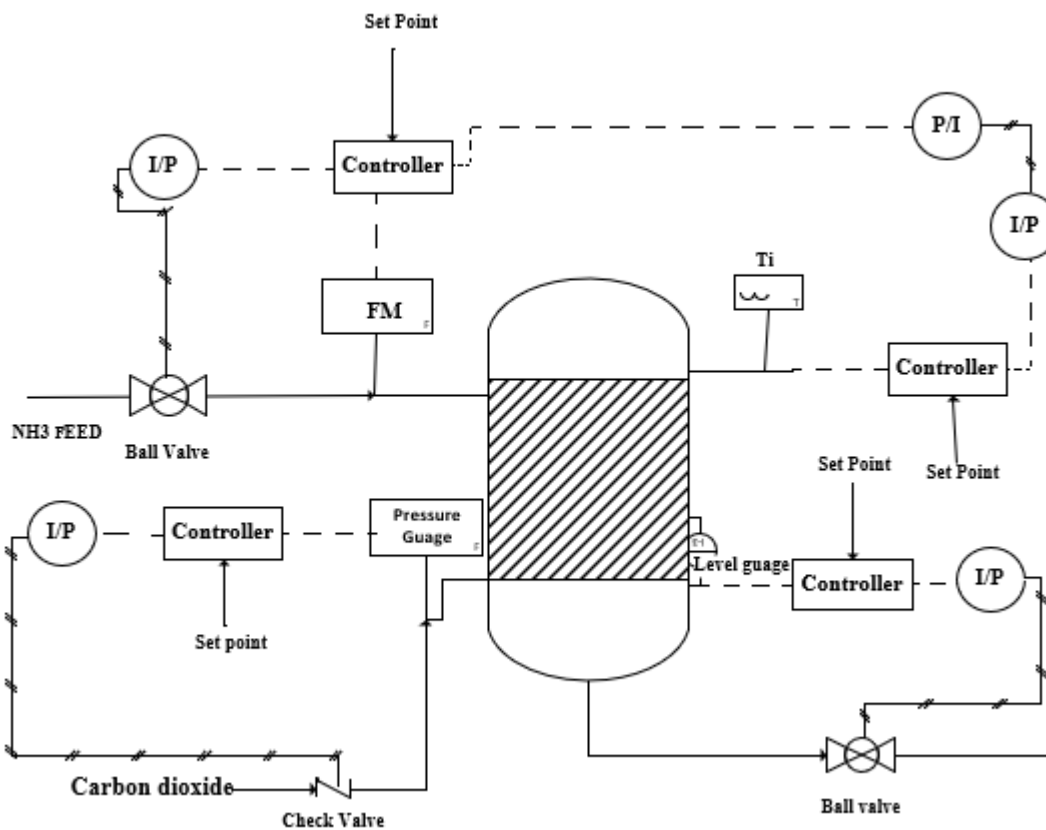
6.4.1 Objectives

Following are the control objectives for the reactor

➢ To maintain the temperature of the reactor as this is a exothermic reaction.

- ➢ Keeping the pressure in check in the reactor.
- ➢ Controlling the level of the mixture in the reactor to avoid any kind of spills.

6.4.2 Disturbances

- ➢ Influent flowrate
- ➢ Feed temperature
- ➢ Level disturbances

6.4.3 Manipulated variables

- ➢ Cooling water flow rate
- ➢ Effluent flow rate

6.4.4 Control Variables

- ➢ Temperature of reactor, as the reaction is exothermic
- ➢ Level of the reactor mixture inside the reactor otherwise spillage may occur

7. Chapter 7 – Cost Estimation

7.1 Economics Analysis:

This section deals with the overall economics of the suggested plant. The plant installation cost, operational cost, raw material costs and all the taxes have been incorporated into the economic study. All the costs are taken in USD because it is the standard currency, and all the economic evaluations are dealt in USD.

7.1.1 Chemical Engineering Plant Cost Index (Cepce):

According to CEPCE, the preliminary value for the CE Plant Cost Index (CEPCI) for March 2022 (the most recent available) rose compared to the previous month. However, the final CEPCI value for February 2022 was revised downwardly from its initial value. In March, all four of the major sub-indices (Equipment, Construction Labor, Buildings and Engineering & Supervision) were slightly higher in March than then previous month. The current CEPCI value now sits at 23.0% higher than the corresponding value from March 2021. Meanwhile, the Current Business Indicators show small increases in the CPI output index and the CPI operating rate for March 2022, and a small increase in the CPI value of output for February 2022.

7.1.2 Pressure Factor:

The Pressure Factor Adjustment is a factor by which the metered usage is multiplied to determine the actual usage. The Pressure Factor Adjustment will be applied to any meter with a deliver pressure above 0.625 psig (or gauge pressure).

$PF = ([GP_0] - [GP_1])/GP_2)$

PF = Pressure Factor Adjustment

GP_0 = Gauge Pressure (psi)

GP_1 = Atmospheric Pressure of 14.10 psi

GP_2 = Base Pressure of 14.73 psi

7.1.3 Purchase Equipment Cost:

PE is the purchased price of equipment from a vendor (someone selling the equipment). It is one of the major factors in the TCI direct costs. It includes the cost to build the equipment but does not include the cost associated with transportation of that equipment to the site, and installation, etc. The cost of each equipment must be calculated to find the purchase equipment cost. Bare module cost method is used for each equipment.

7.2 Cost Estimation of Plug Flow Reactor (PFR):

Reactors*
304 stainless steel; jacketed; no agitation
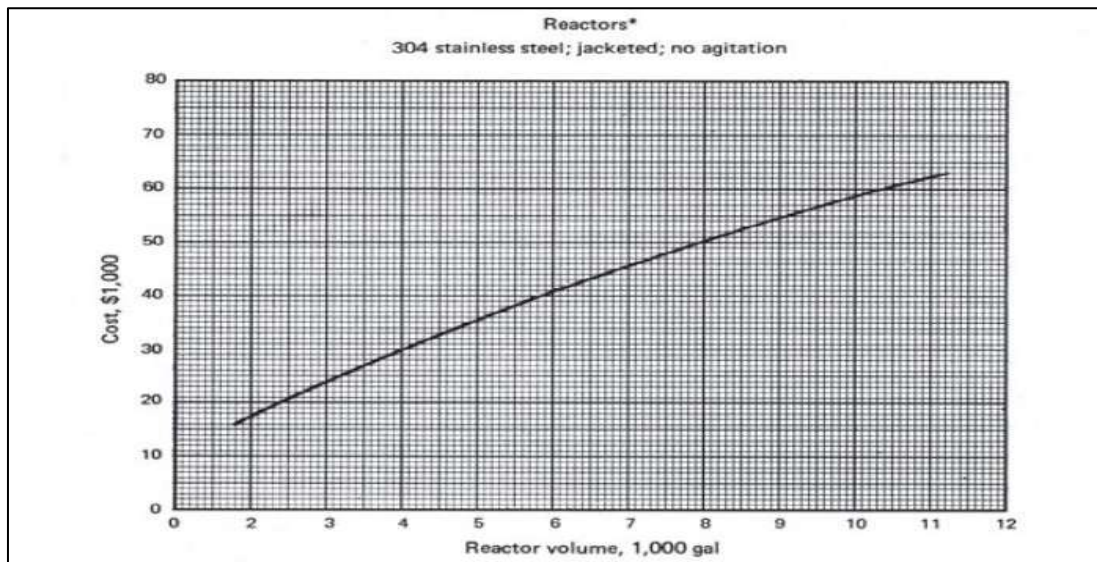
Pressure factor =1

Material factor =1.6

To get an estimate of the average cost for a Plug flow reactor, we use the following relation, (Turton et al….2009)

$$C_{BM} = C^{\circ}{}_{P} \times F_{BM}$$

Where, $C^{\circ}{}_{P}$ is purchasing cost found by formulae given below (Turton et al….2009)

$$log_{10}C^{\circ}{}_{P} = K_1 + K_2 log_{10}A + K_3[log_{10}A]^2$$

Here, K_1, K_2 and K_3 could be found from literature while V is the volume of the cyclone $V = 242 \; m^3$

Where F_{BM} is bare module factor.

K_1, K_2, and K_3 values are get from Table 41 (Turton et al….2009)

$$log_{10}C^{\circ}_{P} = K_1 + K_2 log_{10}A + K_3[log_{10}A]^2$$

$$log_{10}C^{\circ}_{P} = 3.4974 + 0.4485\ log_{10}242 + 0.1074[log_{10}242]^2$$

$$C^{\circ}_{P} = 138,038\$$$

Cost index in 2001 = 400

Cost index in 2021 = 701.4

Cost in 2001   = \$898,630

Then, cost in 2021 = cost in 2001 x cost index in 2021/cost index in 2001

$$= \$1,575,747.89$$

Now, Similarly we can calculate all the prices for the equipment's

7.3 Cost Estimation of Stripper:

Pressure factor= 1

Material factor = 1.6

$$C_{BM} = C^{\circ}_{P} \times F_{BM}$$

Where, C°_{P} is purchasing cost found by formulae given below (Turton et al….2009)

$$log_{10}C^{\circ}_{P} = K_1 + K_2 log_{10}A + K_3[log_{10}A]^2$$

Here, K_1, K_2 and K_3 could be found from literature while V is the volume of the Stripper $A =$

2.0096 m^2

Where F_{BM} is bare module factor.

K_1, K_2, and K_3 values are get from Table 41 (Turton et al….2009)

$$log_{10}C^{\circ}{}_{P} = K_1 + K_2 log_{10}A + K_3[log_{10}A]^2$$

$$log_{10}C^{\circ}{}_{P} = 3.4974 + 0.4485 \ log_{10}2.0096 + 0.1074[log_{10}2.0096]^2$$

$$C^{\circ}{}_{P} = 4265.8\$$$

Cost index in 2001 = 400

Cost index in 2021 = 701.4

Cost in 2001   = $22020.05

Then, cost in 2021 = cost in 2001 x cost index in 2021/cost index in 2001

$$= \$ \ \ 38612.15$$

7.4 Cost Estimation of Vacuum Evaporator:

Pressure factor =1

Material factor = 2

$$C_{BM} = C°_P \times F_{BM}$$

Where, $C°_P$ is purchasing cost found by formulae given below (Turton et al….2009)

$$log_{10}C°_P = K_1 + K_2 log_{10}A + K_3[log_{10}A]^2$$

Here, $K_1, K_2 \, and \, K_3$ could be found from literature while V is the volume of the Evaporator $A =$

$311.2 \, m^2$

Where F_{BM} is bare module factor.

K_1, K_2, and K_3 values are get from Table 41 (Turton et al….2009)

$$log_{10}C°_P = K_1 + K_2 log_{10}A + K_3[log_{10}A]^2$$

$$log_{10}C°_P = 4.6420 + 0.3698 \, log_{10}311.2 + 0.0025[log_{10}311.2]^2$$

$$C°_P = 371535.22\$$$

Cost index in 2001 = 400

Cost index in 2021 = 701.4

Cost in 2001   = \$2419437.35

Then, cost in 2021 = cost in 2001 x cost index in 2021/cost index in 2001

= \$ 4234015.36

7.5 Cost Estimation of Decomposer:

Pressure factor = 1

Material factor = 2

Purchase Cost in 2003 =

7.6 Cost Estimation of NH₃ (centrifugal) Pump:

Pressure factor = 1

Cost of pump in 2003 = $115,00

7.7 Cost Estimation of Prilling Tower:

Pressure factor = 1

Cost of prilling tower = $20,500

Sr.No	Equipment	No. of equipment	Cost in 2003	Total cost in 2003
1	Reactor	1	$125,500	$125,500
2	Stripper	1	$16,400	$16400
3	Vacuum Evaporator	2	$23,900	$47,800
4	Decomposer	2	$25,000	$50,000
5	NH₃ Pump	1	$11,500	$11,500
6	Prilling tower	1	$20,500	$20,500

				Total = $559,700

Total Cost of Purchased Equipment = $559,700

7.9 Capital Cost

Capital costs are fixed, one-time expenses incurred on the purchase of land, buildings, construction, and equipment used in the production of goods or in the rendering of services. In other words, it is the total cost needed to bring a project to a commercially operable status..

Capital Cost = (Lang Factor)(Sum of Purchased Costs of All Major Equipment)	
Type of Chemical Plant	Lang Factor = F_{Lang}
Fluid processing plant	4.74
Solid-fluid processing plant	3.63
Solid processing plant	3.10

7.10 Direct Cost

Installation Cost (40% of equipment cost) = $223,880

Instrumentation and Control Cost (20% of equipment cost) = $111,940

Piping Cost (25% of equipment cost) = $139,925

Electric Cost (20% of equipment cost) = $111,940

Building Cost (30% of equipment cost) = $167910

Service Facilities Cost (40% of equipment cost) = $223,880

Land Cost (4% of equipment cost) = $22,388

Total Direct Cost = $901,143

7.10.1 Indirect Cost

Engineering and Supervision Cost (15% of equipment cost) = $83,955

Construction Expenses (10% of equipment cost) = $55,970

Contractor fee (10% of equipment cost) = $55,970

Contingency fee (30% of equipment cost) = $167910

Total Indirect Cost = $363,805

7.10 Total Capital Investment Required

Fixed Capital Investment = Direct cost + Indirect Cost

Fixed Capital Investment = **$1,292,907**

Working Capital (18% of Fixed Capital Investment) = **$232,723**

Total Capital Investment = Working Capital + Fixed Capital Investment

Total Capital Investment = **$1,264,948**

7.11 Working Capital Cost (WCI)

WCI   =   15% of Fixed Capital Cost

WCI   =   15% (1,264,948)

       =   **$189,742**

Total Capital Cost   =   FCI + WCI

       =   **$1,454,690**

7.12 Total Product Cost (TPC)

(a) Direct Production Cost

1. Operating Labor cost   =   10% x TPC
2. Utility Cost   =   20% x TPC
3. Maintenance & Repair Cost =   4% x FCI   = $51,716
4. Operating & supply cost   =   15% (M&RC) = $7757.4
5. Laboratory & Testing cost   =   15% (10% x TPC)
6. Patent & Royalties   =   3% x TPC

Direct Product Cost (DPC)   =   1 + 2 + 3 + 4 + 5 + 6

$$= 0.1 \text{xTPC} + 0.2 \text{xTPC} + 51{,}716 + 7757 + .015 \text{xTPC} +$$

$$0.03 \text{xTPC}$$

$$\text{TPC} = 0.345 \text{xTPC} + \$59{,}473$$

(b) Plant Overhead Cost

Plant overhead cost $=$ 60% x (Labor + maintenance Cost)

$$= 0.06 \text{ x } (0.10 \text{ x TPC} + \$51{,}716)$$

(c) Administrative Expenses

Administrative Expenses $=$ 10% x labor cost

$$= 0.10 \text{ x } (0.10 \text{xTCP})$$

(d) Distribution & Market Expenses

Distribution & market expenses $=$ 10% x TCP

Research & development cost $=$ 5% x FCI

$$= \$64{,}645$$

(e) Interest

Interest $=$ 8% x FCI

$$= \$103{,}432$$

Total Product Cost $= a + b + c + d + e$

TPC $= 0.345 \text{xTPC} + \$59{,}473 + 250{,}918 + 0.10 \text{ x } (0.10 \text{xTCP}) + 10\% \text{TCP} +$

$64,645 + $ $\qquad 0.06 \text{ x } (0.10 \text{ x TPC} + \$51{,}716) + \$103{,}432$

TPC $= 0.515(\text{TPC}) + \$481{,}571$

TPC $= \$100{,}000{,}0$

7.13 Profitability Anaylsis

Wholesale selling price per bag (50 kg) in Pakistan = 1950Rs

$$= \$9.8$$

Per 1kg price of Urea in Pakistan   = $0.196

Total Income   =   Selling Price x Quantity of Product

Production in 1 day   =   1900 metric ton   =   1900000

Production in 365 days = 1900000 x 365   =   693,500,000

Then, total income   =   693,500,000 kg x $0.196

= $135,926,000 kg

Gross Profit

Gross profit =   total income - Total Product Cost

=   135,926,000 – 100,000,0

= $134,926,000

Let tax be 40%,

Then, taxes   =   40% x gross profit

=   53,970,400

Net Profit   =   total income − taxes

=   135,926,000 − 53,970,00

= $81,955,600.

7. Chapter 8 – Conclusion

Most of the energy required for industrial urea production is consumed in the following steps: compression and pumping of raw materials to achieve the required pressure levels for urea synthesis. Separation of unconverted ammonium carbamate and excess ammonia from reaction product.

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