

MANUFACTURE OF UREA

A THESIS SUBMITTED IN PARTIAL FULFILLMENT
OF THE REQUIREMENTS FOR THE DEGREE OF

Bachelor of technology
in
Chemical Engineering

By

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Department of Chemical Engineering

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Under the Guidance of
Prof. S K Agarwal



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National Institute of Technology

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2007



**National Institute of Technology
Rourkela**

CERTIFICATE

This is to certify that the thesis entitled, "MANUFACTURE OF UREA" submitted by Sri KUMAR BHASKAR & PRATAP CHANDRA DAS in partial fulfillments for the requirements for the award of Bachelor of Technology Degree in Chemical Engineering at National Institute of Technology, Rourkela (Deemed University) is an authentic work carried out by them under my supervision and guidance.

To the best of my knowledge, the matter embodied in the thesis has not been submitted to any other University / Institute for the award of any Degree or Diploma.

Date:

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3rd May 2007

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ABSTRACT

Urea is in many ways the most convenient form for fixed nitrogen. It has the highest nitrogen content available in a solid fertilizer (46 %). It is easy to produce as prills or granules and easily transported in bulk or bags with no explosive hazard. It leaves no salt residue after use on crops. Its specific gravity is 1.335, decomposes on boiling and is fairly soluble in water.

The principal raw materials required for this purpose are NH_3 & CO_2 . Two reactions are involved in the manufacture of urea. First, ammonium carbamate is formed under pressure by reaction between CO_2 & NH_3 .



This highly exothermic reaction is followed by an endothermic decomposition of the ammonium carbamate.



Various processes for the manufacture of urea are:

- 1) Snamprogetti ammonia stripping process
- 2) Stamicarbon CO_2 stripping process
- 3) Once through urea process
- 4) Mitsui Toatsu total recycle urea process

We selected the Snamprogetti ammonia stripping process for the manufacture of urea.

In this process ammonia & CO_2 are compressed & fed to the reactor. The unconverted carbamate is 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 & recycled back to the reactor by gravity. Part of the liquid NH_3 reactor feed, vapourized in a steam heated exchanger, is used as inert gas to decompose & strip ammonium carbamate in the steam heated high pressure stripper.

Energy balance & material balance of the plant is done. The selected capacity of the plant is 4,50,000 tons/year of urea producing 62,500 kg/hr of urea with 98 % purity. Urea reactor & vacuum evaporator are designed. The volume of reactor is calculated & found to be 195 m³. The length & diameter of the reactor are 40 m & 2.5 m respectively. The evaporator used is of climbing-film long- tube type.

Snamprogetti ammonia-stripping urea process is selected because it involves a high NH₃ to CO₂ 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.

Uses of Urea:

- About 56 % of Urea manufactured is used in solid fertilizer.
- About 31 % of Urea manufactured is used in liquid fertilizer.
- Urea-formaldehyde resins have large use as a plywood adhesive.
- Melamine-formaldehyde resins are used as dinnerware & for making extra hard surfaces.

Chapter 1

INTRODUCTION

1.1 MANUFACTURE OF UREA [1]

Urea is an important nitrogenous fertilizer. Its utilization is increasing steadily, it being the preferred nitrogen fertilizer worldwide. It is used in solid fertilizer, liquid fertilizer, formaldehyde resins and adhesives.

Rouelle first discovered urea in urine in 1773. His discovery was followed by the synthesis of urea from ammonia and cyanic acid by Woehler in 1828. This is considered to be the first synthesis of an organic compound from an inorganic compound. In 1870, Bassarow produced urea by heating ammonium carbamate in a sealed tube in what was the first synthesis of urea by dehydration. The chemical formula of, NH_2CONH_2 , indicates that urea can be considered to be the amide of carbamic acid NH_2COOH , or the diamide of carbonic acid $\text{CO}(\text{OH})_2$.

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 aim of the fertilizer industry is to provide the primary and secondary nutrients which are required in macro quantities. Primary nutrients are normally supplied through chemical fertilizers. They are chemical compounds containing one or more of the primary nutrients and are generally produced by chemical reactions. Whatever may be the chemical compounds, its most important ingredient for plant growth is the nutrient content. The primary 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. Thus ammonium sulphate is represented by 20.6-0-0.

1.2 PROPERTIES OF UREA :

PHYSICAL PROPERTIES OF UREA

Urea is a white, odorless, hygroscopic solid. It is non-corrosive.

CHEMICAL PROPERTIES OF UREA

Molecular weight	60.05
Relative humidity	60 %
Maximum Nitrogen content	46.6 %
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

Table-1.1 SPECIFIC HEAT OF UREA

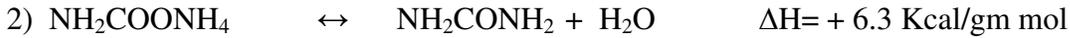
Temperature, °C	Specific heat, Kj/kg °C
0	1.398
50	1.66
100	1.89
150	2.11

1.3 USES OF UREA

- About 56 % of Urea manufactured is used in solid fertilizer.
- About 31 % of Urea manufactured is used in liquid fertilizer.
- Urea-formaldehyde resins have large use as a plywood adhesive.
- Melamine-formaldehyde resins are used as dinnerware & for making extra hard surfaces.

1.4 PROCESS TECHNOLOGY

Although there are several processes currently used for the manufacture of urea, the underlying principle for all the processes is same. The two main reactions involved are:



Undesirable side reaction taking place is:



(Biuret)

Both 1st & 2nd reactions are equilibrium reactions. The 1st reaction almost goes to completion at 185-190 °C & 180-200 atms. The 2nd reaction (decomposition reaction) is slow and determines the rate of the reaction. Unconverted CO₂ & NH₃, along with undecomposed carbamate, must be recovered and re-used. This is a troublesome step. The synthesis is further complicated by the formation of a dimer called biuret, NH₂CONHCONH₂, which must be kept low because it adversely affects the growth of some plants.

1.5 PROCESS IN GENERAL. [2]

Ammonia & CO₂ are compressed separately and fed to the high pressure (180 atms) autoclave as shown in fig-1.1 which must be water cooled due to the highly exothermic nature of the reaction. A mixture of urea, ammonium carbamate, H₂O and unreacted (NH₃+CO₂) is produced.

This liquid effluent is let down to 27 atms and fed to a special flash-evaporator containing a gas-liquid separator and condenser. Unreacted NH₃, CO₂ & H₂O are thus removed & recycled. An aqueous solution of carbamate-urea is passed to the atmospheric flash drum where further decomposition of carbamate takes place. The off gases from this step can either be recycled or sent to NH₃ processes for making chemical fertilizer.

The 80 % aqueous urea solution can be used as it is, or sent to a vacuum evaporator to obtain molten urea containing less than 1 % water. The molten mass is then sprayed into a prilling tower. To avoid formation of biuret and keep it less than 1 %, the temperature must be kept just above the melting point for processing times of 1-2 seconds in this phase of the operation.

THE VARIABLES THAT AFFECT THE AUTOCLAVE REACTIONS ARE [2]:

1) Temperature [Fig-1.2]:

Process temperature (185 °C) favours equilibrium yield at a given pressure (180 atm). The conversion of ammonium carbamate to urea gradually increases as the temperature increases. However, after a particular temperature, depending upon the pressure, the conversion suddenly drops with further increase in temperature. The pressure corresponding to this temperature which is usually in the range of 175-185°C, is known as the decomposition pressure which is about 180 atm.

2) Pressure [Fig-1.3]:

The main reaction is sufficiently slow at atmospheric pressure. However, it starts almost instantaneously at pressure of the order of 100 atm and temperature of 150 °C. There is reduction in volume in the overall reaction and so high pressure favors the forward reaction. This pressure is selected according to the temperature to be maintained & NH₃:CO₂ ratio.

3) Concentration:

Higher the concentration of the reactants, higher will be the forward reaction according to the law of mass action. CO₂ being the limiting reagent higher NH₃:CO₂ ratio favors conversion. Since, dehydration of carbamate results in urea production, lesser H₂O:CO₂ ratio favors conversion. Water intake to the reactor should therefore be minimum.

4) Residence time:

Since, urea reaction is slow and takes about 20 mins to attain equilibrium, 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 and concentration.

5) Biuret formation:

A problem faced during manufacture of urea is the formation of biuret during the production of urea. It is not a desirable substance because it adversely affects the growth of some plants. Its content in urea should not be more than 1.5 % by weight.



(Biuret)

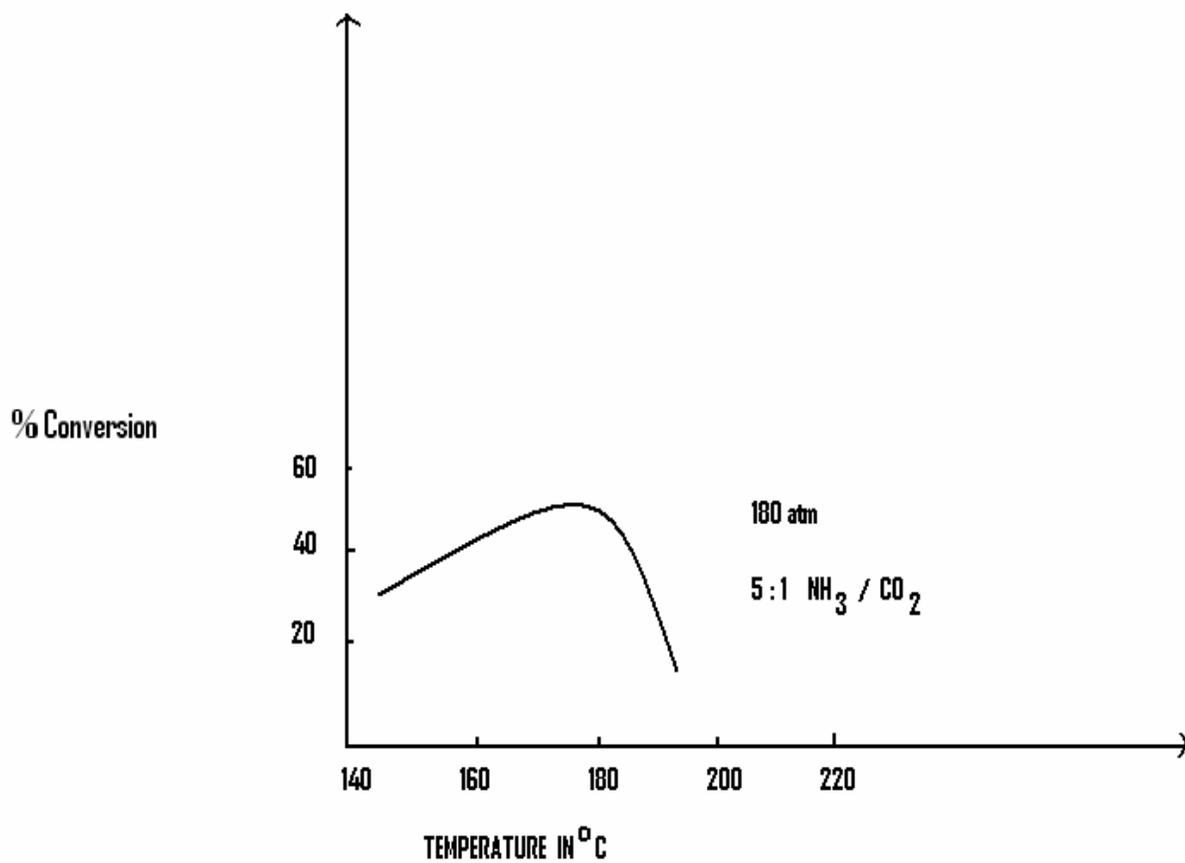


Fig 1.1 Percentage conversion Vs temperature graph for the yield of urea

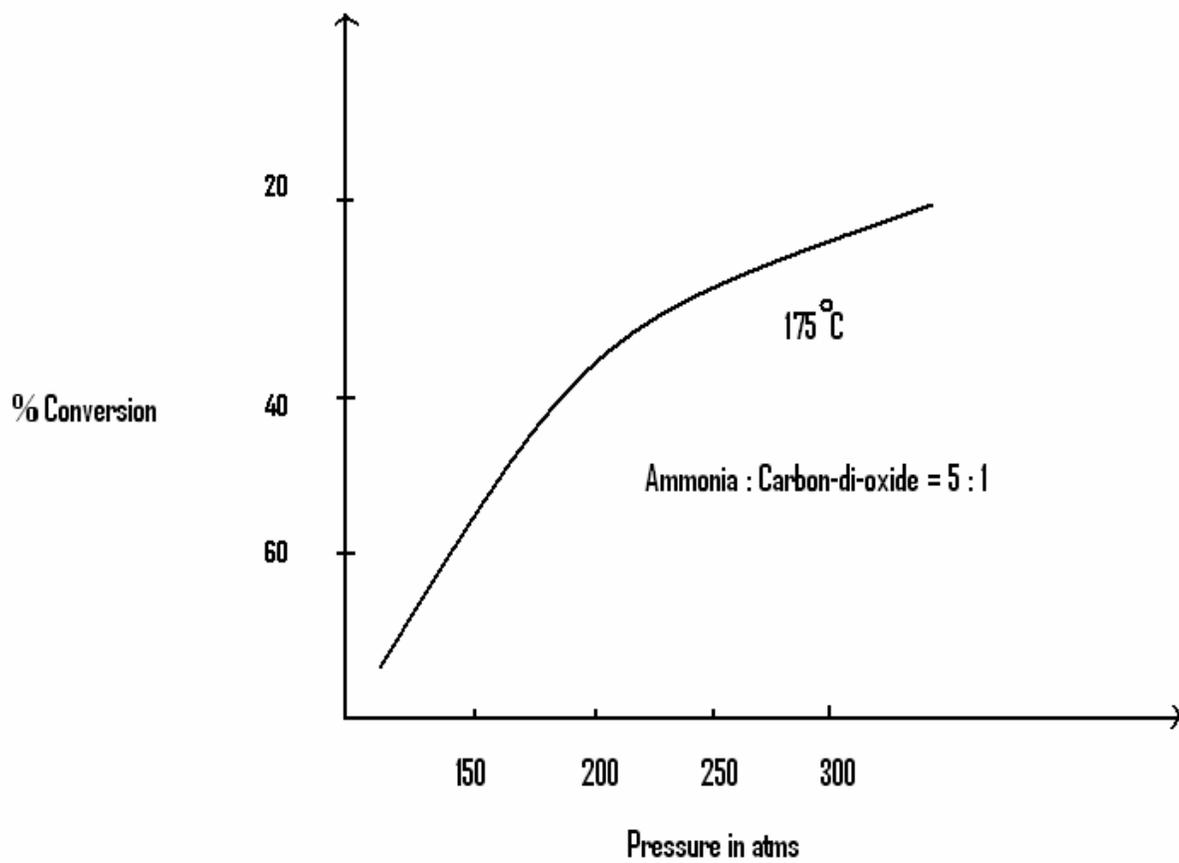


Fig 1.2 Percentage conversion Vs pressure graph for the yield of urea

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.

1.6 MAJOR ENGINEERING PROBLEMS

1) Carbamate decomposition and 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 $\text{NH}_3\text{-CO}_2$ process was slow to develop. High cost silver or tantalum liners are used in the autoclaves with hastelloy C, titanium, stainless steel (321 SS), and aluminium 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.

Chapter 2

Various methods of manufacture
of urea along with flow sheets.

2.1 VARIOUS PROCESSES FOR THE MANUFACTURE OF UREA [3]

The urea synthesis reactor always contains unreacted carbamate & more or less excess ammonia, depending upon the composition of the feeds. This poses the practical problem of separating the unreacted material from the urea solution & of reutilizing this unreacted material. Depending upon the method of reutilization of the unreacted material, the commercial urea synthesis processes are divided into the following main categories:

1) **Once-through urea process:**

The unconverted carbamate is decomposed to NH_3 & CO_2 gas by heating the urea synthesis reactor effluent mixture at low pressure. The NH_3 & CO_2 gas is separated from the urea solution and utilized to produce ammonium salts by absorbing NH_3 , either in sulfuric or nitric acid.

Once through urea process [Fig-2.1]:

In this process liquid NH_3 is pumped through a high pressure plunger pump and gaseous CO_2 is compressed through a compressor up to the urea synthesis reactor pressure at an NH_3 to CO_2 feed mole ratio of 2/1 or 3/1. The reactor usually operates in a temperature range from 175 to 190 °C. The reactor effluent is let down in pressure to about 2 atm and the carbamate decomposed and stripped from the urea-product solution in a steam heated shell & tube heat exchanger. The moist gas, separated from the 85-90 % urea product solution, & containing about 0.6 tons of gaseous NH_3 per ton of urea produced is usually sent to an adjacent ammonium nitrate or ammonium sulfate producing plant for recovery. An average conversion of carbamate to urea of about 60 % is attained. Excess heat is removed from the reactor by means of a low-pressure steam-producing coil in an amount of about 280,000 cal/Kg urea produced.

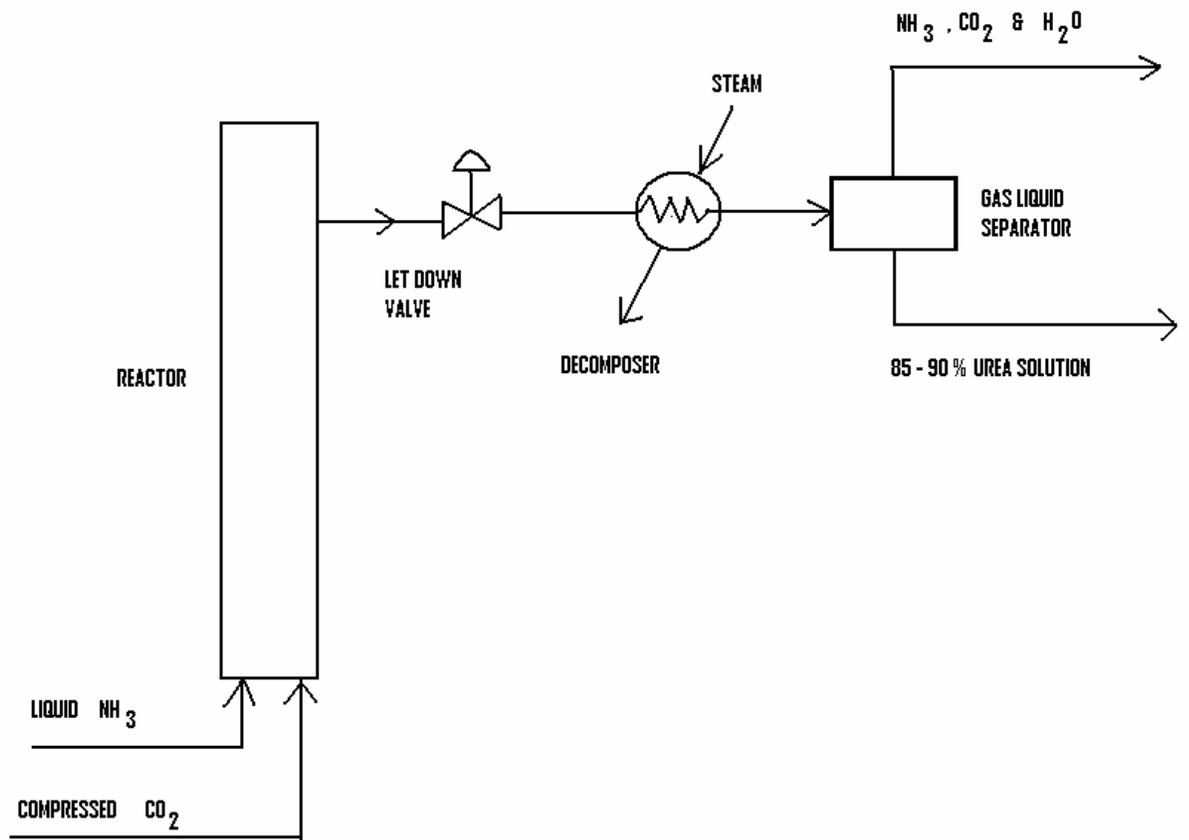


Fig 2.1 Once through urea process

2) Solution recycle urea process:

The NH_3 & CO_2 gas recovered from the reactor effluent mixture either in one or in several pressure staged decomposition sections is absorbed in water and recycled back to the reactor in the form of an ammoniacal aqueous solution of ammonium carbamate.

3) Internal carbamate recycle urea process:

The unreacted carbamate & the excess ammonia are stripped from the urea synthesis reactor effluent by means of gaseous hot CO_2 or NH_3 at the reactor pressure, instead of letting the reactor effluent down to a much lower pressure. The NH_3 & CO_2 gas, thus recovered at reactor pressure, is condensed and returned to the reactor by gravity flow for recovery.

a) Snam Progetti (Italy) [Fig-2.2]

This 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 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.

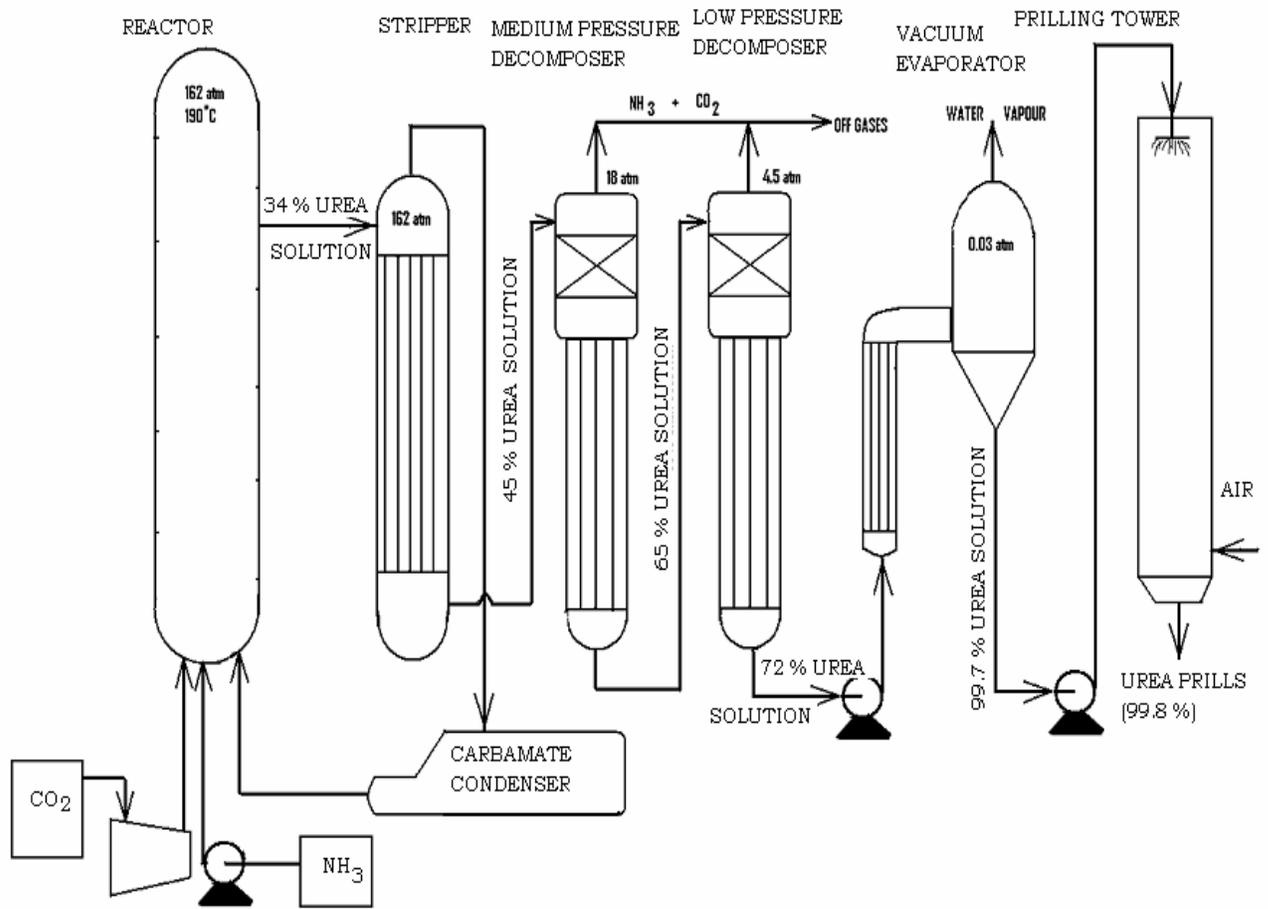


Fig 2.2 Snamprogetti process

b) Stamicarbon (Holland) [Fig-2.3]

The novelty of the CO₂ stripping process consists of the fact that the reactor effluent is not let down to a lower pressure as in the conventional liquid recycle urea process, but is stripped at synthesis pressure by the gaseous CO₂ reactor feed stream in a steam heated vertical heat exchanger.

The high pressure stripper operates at about 140 atm & 190 °C. The stripped urea solution still contains about 15 % of the unconverted carbamate, & it is let down to about 3 atm for further degassing in the steam heated low pressure decomposer at about 120 °C. The off gas recovered is condensed with cooling water in the low pressure condenser, operating at about 65 °C & 3 atm. The solution thus obtained is pumped to the high pressure condenser by means of high pressure carbamate pump. The off gas recovered from the high pressure stripper is condensed in the high pressure condenser, which operates at about 170 °C & 140 atm. The heat of condensation is removed on the shell side of the condenser by vaporizing the equivalent amount of condensate. The 3.4 atm steam thus produced can be reused in another section of the plant.

The mixture of gas & liquid issuing from the high pressure condenser is fed to the reactor for total CO₂ condensation to carbamate & subsequent conversion to urea. The inerts are vented from the reactor through a water cooled vent condenser. The reactor effluent, at about 185 °C & 140 atm, is fed to the high pressure stripper as described above.

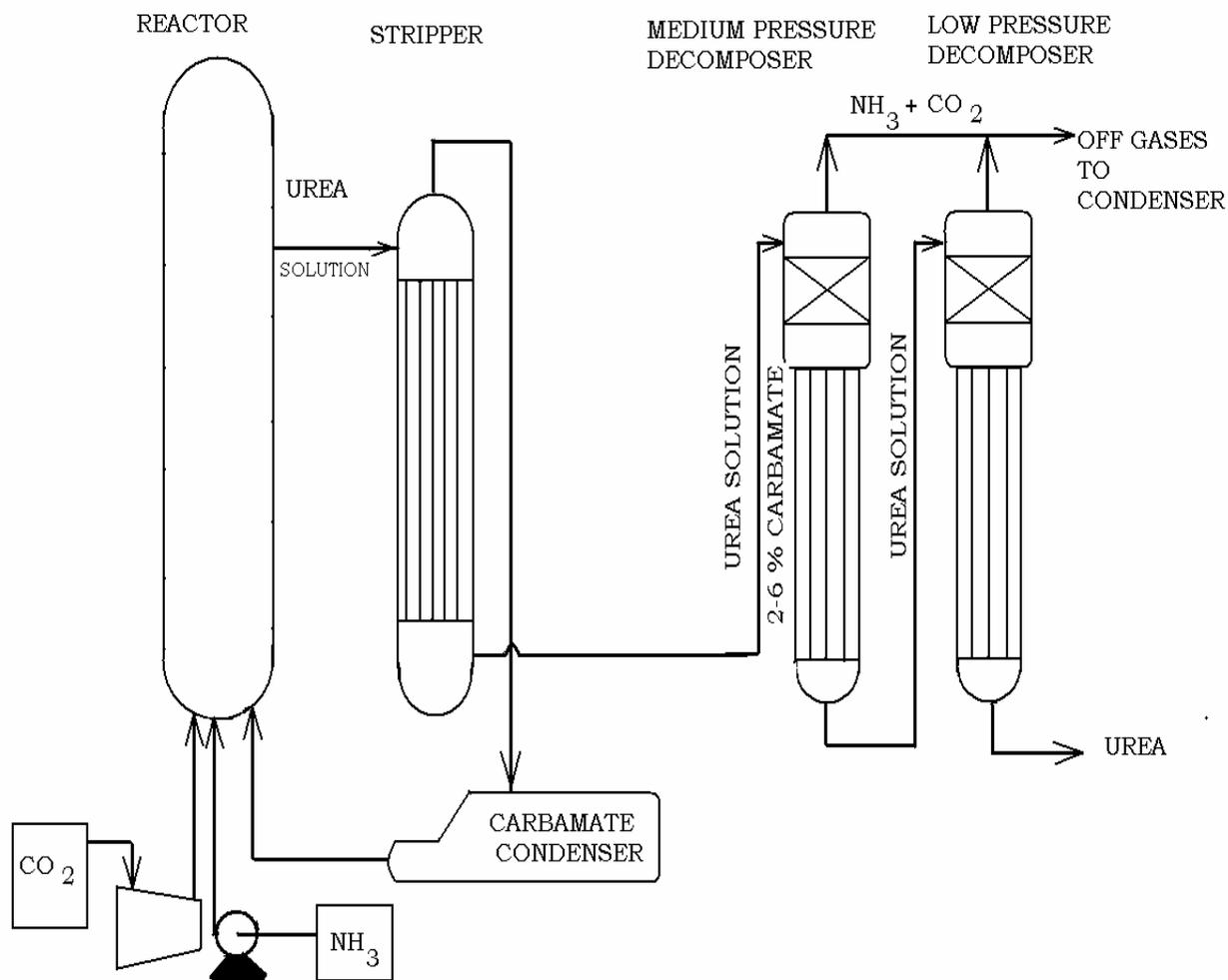


Fig 2.3 Stamicarbon process

c) Mitsui Toatsu or Toyo Koatsu process (Japan) [Fig-2.4]

This total recycle urea process is a conventional carbamate solution recycle process with three pressure-staged carbamate decomposition & recovery systems. The reactor is operated at about 195 °C, 240 atm, & with a NH₃ to CO₂ molar ratio of about 4.3. About 67 % of the total ammonium carbamate present in the reactor is converted to urea.

The unconverted carbamate is decomposed and stripped from the urea solution together with excess NH₃ in a series of three pressure-stage decomposers, operating respectively at about 18 atm & 150 °C, 3.06 atm & 130 °C, & atmospheric pressure & 120 °C. The main feature of the Mitsui Toatsu process is the fact that the gaseous phase in each decomposition stage is contacted in counter current flow with the urea product solution issuing from the preceding decomposition stage. Either a packed section or a sieve tray section is used for this purpose. The effect is that the NH₃ & CO₂ gaseous mixture obtained from the decomposition of carbamate is considerably reduced in vapor content. Thus the amount of water recycled to the reactor is maintained at a relatively low level and a relatively high conversion in the reactor is attained. The off-gas from each decomposition stage is condensed to solution in its respective water cooled condenser and the solution thus obtained is pumped to the next high pressure staged condenser.

Excess NH₃ is separated from the aqueous solution of carbamate & scrubbed from the last traces of CO₂ in counter current flow with reflux liquid, NH₃, fed to the top of the high pressure absorber. The pure excess NH₃ thus obtained is condensed to liquid with cooling water & recycled to the reactor. The carbamate solution is recycled back to the reactor for total recovery.

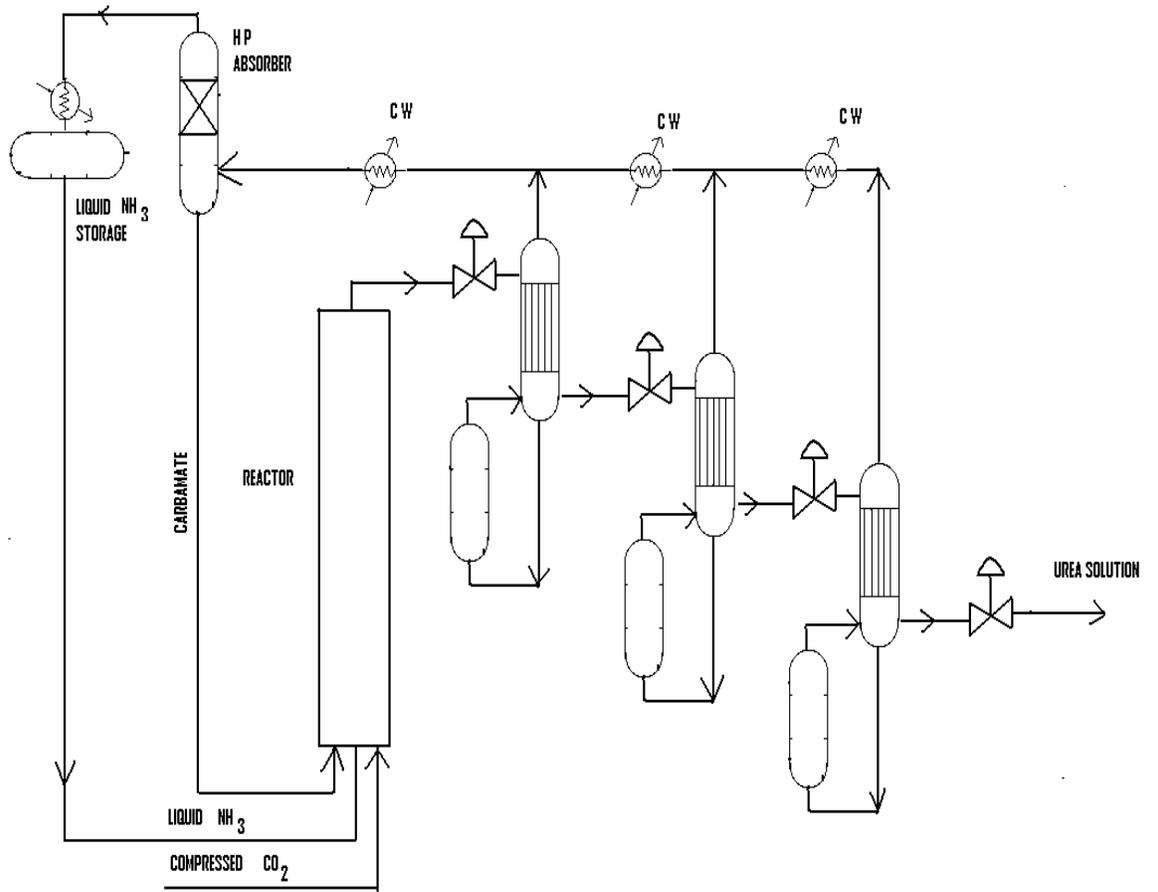
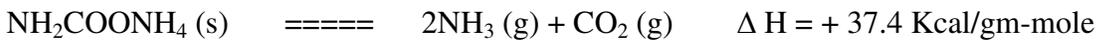


Fig 2.4 Mitsui toatsu total recycle urea process

2.2 SELECTION OF THE PROCESS [Fig 2.5]

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.

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 mfg process depends on how economically we can recycle carbamate to the reactor. Snamprogetti process of urea manufacturing accomplishes the above task by stripping process.



This reaction involves increase in volume & absorption of heat. Thus this reaction will be favored by decrease in pressure & increase in temp. 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 :

$$K = (\text{pNH}_3)^2 * (\text{pCO}_2) \quad [\text{where, } K = \text{equilibrium constant}]$$

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 increases pCO_2 . So pNH_3 will be reduced to maintain P constant as $P = \text{pCO}_2 + \text{pNH}_3$.

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 alongwith the

carbamate to reactor again & if water enters reactor production will be affected adversely due to hydrolysis of urea . So , stagewise decomposition of carbamate is done . Second consideration in favor of isobaric stripping is that higher carbamate recycle pressure results in condensation at higher temp & that recovery in the form of low pressure steam. This is why stagewise reduction in pressure is practised

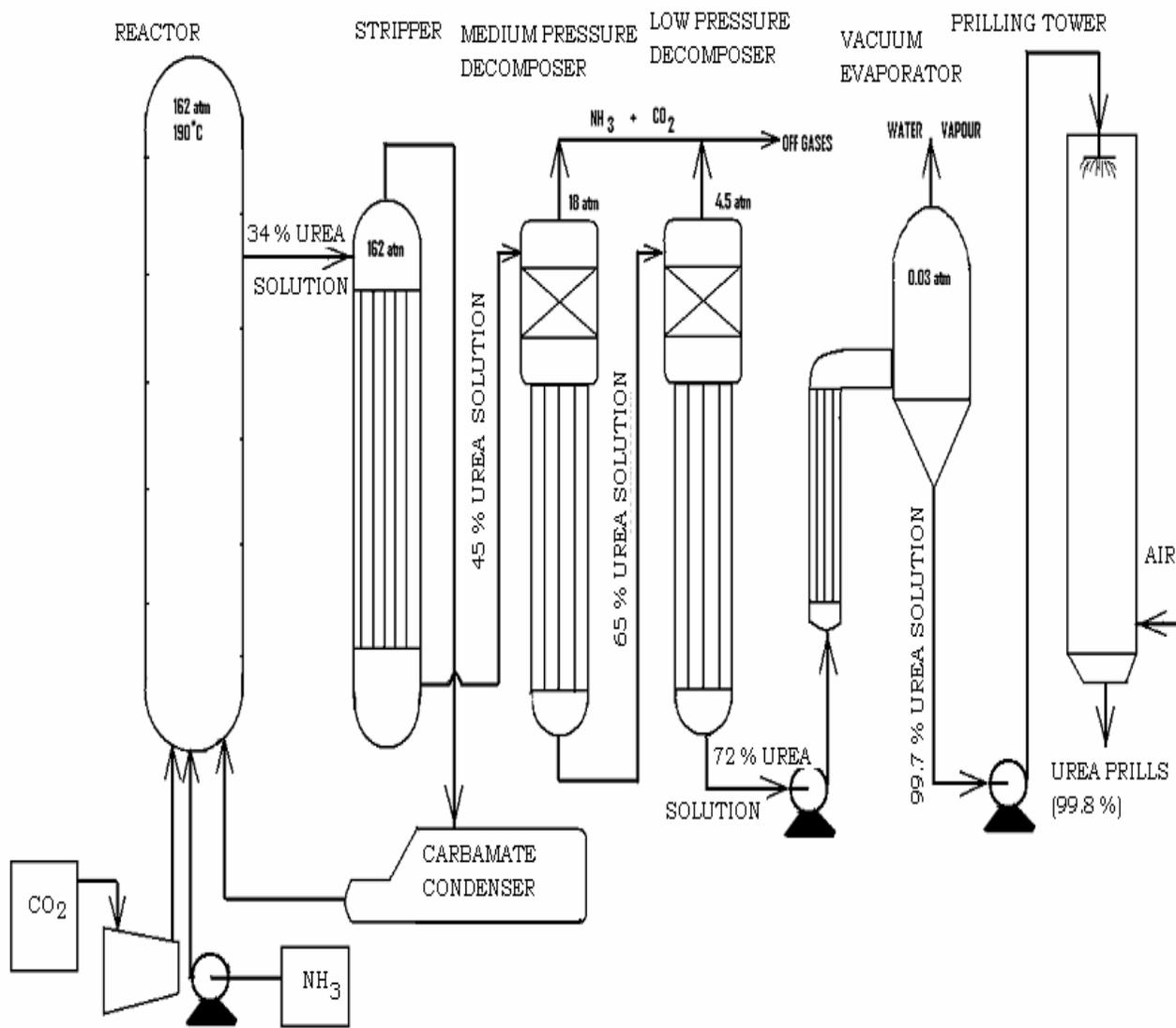


Fig 2.5 Snamprogetti process for manufacture of urea

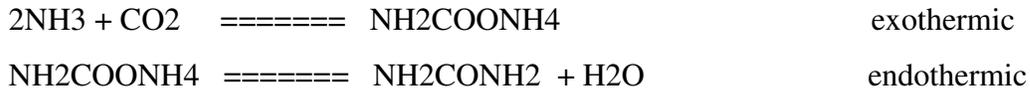
Chapter 3

Selected process : Snamprogetti
ammonia stripping process

SNAMPROGETTI AMMONIA STRIPPING PROCESS

3.1 MANUFACTURING PROCESS [4]

In the reactor, when liquid ammonia reacts with compressed CO₂ (at 162 ata) at high temperature & pressure gives urea according to the following reactions:



As the reactions are reversible in nature only partial conversion occurs in the reactor. Urea solution consisting of Urea ,Carbamate,Water & unconverted CO₂ & NH₃ are fed into the stripper where stripping action of NH₃ favours decomposition of carbamate,and hence 80% of carbamate is decomposed here. Pressure in the stripper is same as that of the reactor.

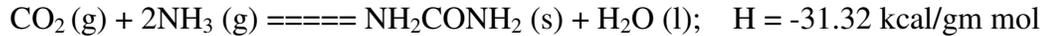
Urea solution from the stripper is sent to Medium pressure decomposer where Urea purification takes place by the dehydration of the Carbamate.Urea solution is further purified in Low pressure decomposer. Off gases from the M P decomposer & L P decomposer are sent to the Medium pressure condenser & Medium pressure absorber for the recovery of unconverted Ammonia. In this way 71.12% of Urea solution resulting from L P decomposer is sent to Vacuum concentrators 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

NH₃ & CO₂ react under specific concentration , temperature & pressure conditions to form Urea as per the following reactions:



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 an urea synthesis reaction will be only 5.74 kcal/gm mol because of the heat lost in evaporation of liquid NH₃, evaporation of water & melting of urea. This is based on the actual plant data.

Further energy is consumed in feeding CO₂ & NH₃ 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.

3.2 EFFECT OF VARIOUS PARAMETERS

TEMPERATURE [Fig-1.2]

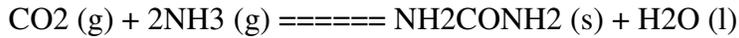
In above reactions 1st reaction is exothermic & 2nd one is endothermic. So, according to Le chatelier's Principle 1st reaction is favoured at low temp & 2nd one at high temp. Further , reaction no. 1 is fast & reaches to completion but reaction no. 2 is slow & determines the overall rate of urea production. 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.



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 $K_2 > K_1$.

PRESSURE [Fig-1.3]:

Overall urea synthesis reaction is given below:



There is reduction in volume in the overall reaction & so high pressure favours the forward reaction. This pressure is selected according to the temp to be maintained & $\text{NH}_3:\text{CO}_2$ ratio.

CONCENTRATION:

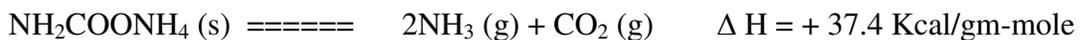
Higher the concentration of the reactants , higher will be the forward reaction according to the law of mass action. CO_2 being limiting reagent higher $\text{NH}_3:\text{CO}_2$ ratio favors conversion . Since , dehydration of carbamate results in urea production , lesser $\text{H}_2\text{O}:\text{CO}_2$ 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.

3.3 SNAMPROGETTI STRIPPING PROCESS [4]

Formation of urea from ammonia & carbon-di-oxide 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.



This reaction involves increase in volume & absorption of heat . Thus this reaction will be favored by decrease in pressure & increase in temp . 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 NH₃ or CO₂ is called stripping process. According to above equation we have :

$$K = (p\text{NH}_3)^2 \cdot (p\text{CO}_2) \quad [\text{where, } K = \text{equilibrium constant}]$$

The stripping is effected at synthesis pressure itself using CO₂ or NH₃ as stripping agent . If CO₂ is selected , it is to be supplied to the decomposers/stripper as in Stamicarbon CO₂ stripping process. While if NH₃ is selected , it is to be obtained from the system itself because excess NH₃ is present in the reactor as in Snam's process. CO₂ stripping is advantageous because introducing CO₂ increase pCO₂. So pNH₃ will be reduced to maintain P constant as $P = p\text{CO}_2 + p\text{NH}_3$.

At a particular temp K is constant so when pNH₃ is reduced to keep K constant , carbamate will be reduced much faster by decomposition as pNH₃ 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 alongwith the carbamate to reactor again & if water enters reactor, production will be affected adversely due to hydrolysis of urea . So , stagewise decomposition of carbamate is done . Second consideration in favor of isobaric stripping is that higher carbamate recycle pressure results in condensation at higher temp & that recovery in the form of low pressure steam. This is why stagewise reduction in pressure is practised.

3.4 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 di oxide drawn from battery limits at about 1.6 ata pressure & about 40°C temp is compressed in a centrifugal compressor upto 162 ata . A small quantity of air is added to the CO₂ 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 23 ata 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 NH₃ of this synthesis loop is compressed to a pressure of about 240 ata . 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 CO₂ .

In the reactor the NH₃ & gaseous CO₂ react to form amm. 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 NH₃ / CO₂ is around 2:1 , the mole ratio of water to CO₂ is around 0.67 : 1 .



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 no.s of sieve trays to avoid back mixing & to avoid escape of gaseous CO₂ which must react in the lower part of the reactor .

Stagewise decomposition is carried out to reduce water carry over to the reactor which could adversely affect conversion .

Urea solution containing urea , carbamate , H₂O & unconverted CO₂ & NH₃ 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 CO₂ content of the solution is reduced by the stripping action of NH₃ as it boils out of the solution . The carbamate decomposition heat is supplied by 24 ata 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 4.5 ata 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 MP decomposer.

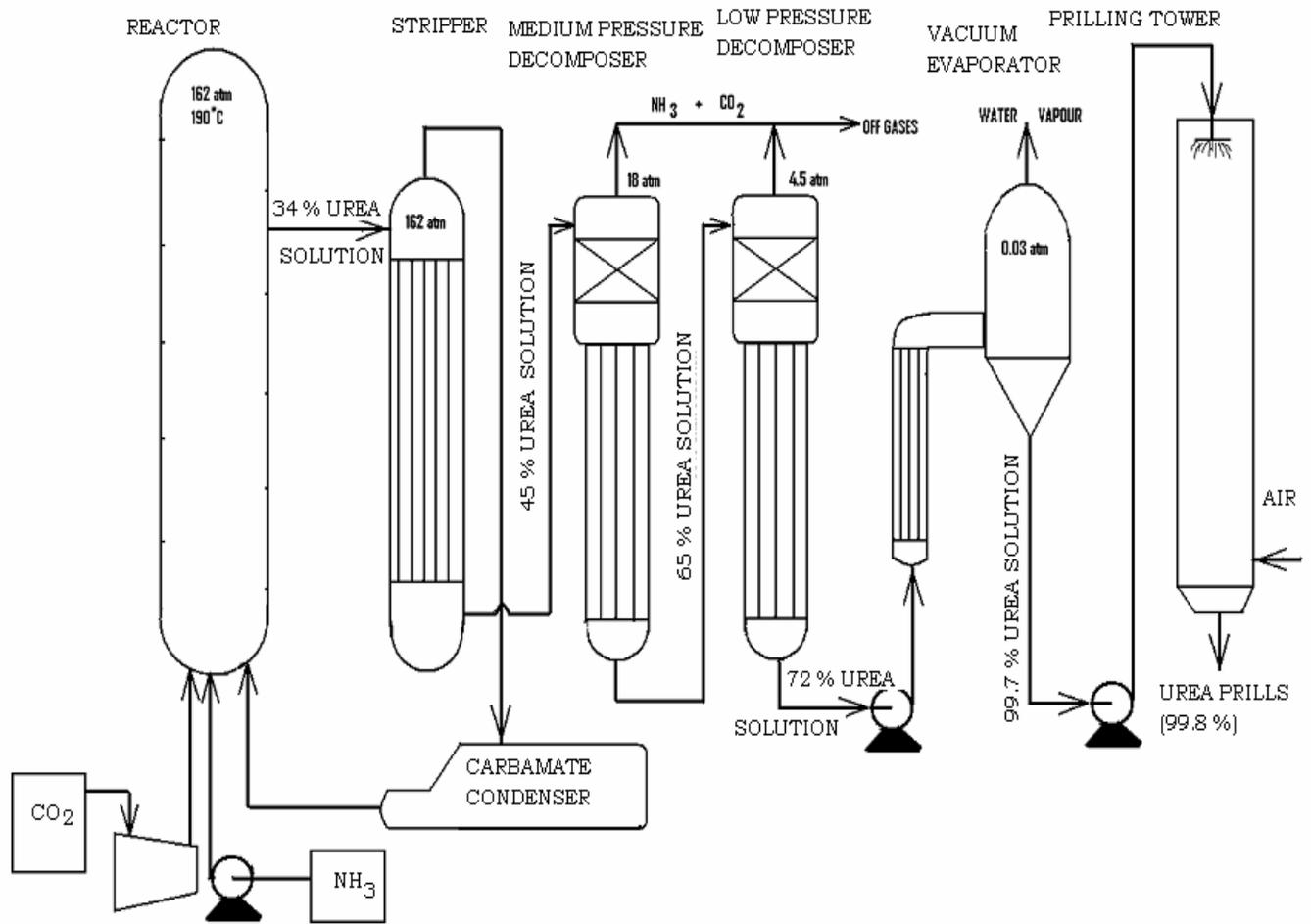


Fig 3.1 Snamprogetti urea process

UREA PURIFICATION & LOW PRESSURE RECOVERY

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

1st stage at 18 ata pressure, i.e, MP decomposer

2nd stage at 4.5 ata pressure ,i.e, LP decomposer

1st stage purification & recovery stage at 18 ata:

It is falling film type MP 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 24 ata steam condensate flowing out of the stripper .

2nd purification & recovery stage at 4.5 ata:

The solution leaving the bottom of MP decomposer is expanded at 4.5 ata pressure& enters the LP 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 4.5 ata.

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 LP decomposer bottom with about 72% urea is sent to the 1st vacuum concentrator operating at a pressure of 0.23 ata .The mixed phase coming out enters the gas liquid separator, wherefrom the vapours are extracted by the 1st vacuum system, while the solution enters the 2nd vacuum concentrator operating at a pressure of 0.03 ata . The two concentrators are fed by saturated steam at 4.5 ata . The mixed phase coming out enters the gas liquid separator , wherefrom the vapours 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 .

Chapter 4

Material balance

MATERIAL BALANCE

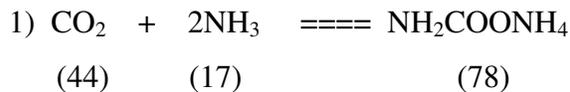
Selected capacity :	4,50,000 tons/year
No. of working days:	300
Daily production :	$4,50,000/300 = 1500$ tons/day
Urea	62,500 Kg/hr of 98 % purity

Composition of the final product :

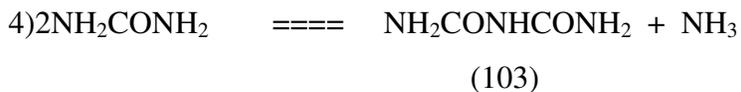
Urea	:	98 %	(61,250 Kg/hr)
Biuret	:	1 %	(625 Kg/hr)
Water	:	1 %	(625 Kg/hr)

Assumption_: Overall conversion to urea is assumed to be 95 %.

MAIN REACTIONS:



Side reaction:



625 Kg/hr of Biuret produced by = $(120/103)*625 = 728$ Kg/hr of urea (reaction 4)

So, urea produced by reaction (2) = $61250 + 728 = 61978$ Kg/hr

61978 Kg/hr of urea produced by = $(34/60)*61978 = 35,120$ Kg/hr NH_3

Similarly, CO_2 reacted in reaction (1) = $(44/60)*61978 = 45,450$ Kg/hr

Assuming 95 % conversion we get

NH_3 actually required = $35120/0.95 = 36,968 \text{ Kg/hr}$

CO_2 actually required = $45450/0.95 = 47,842 \text{ Kg/hr}$

Now, considering reaction (4) :

If reaction (3) is 100 % complete then,

Urea produced = $(60/44)*47842 = 65,239 \text{ Kg/hr}$

But, for 95 % conversion

Urea produced = $0.95*65,239 = 61,977 \text{ Kg/hr}$

Therefore, Urea converted to Biuret & $\text{NH}_3 = 61,977 - 61,250 = 727 \text{ Kg/hr}$

So, from reaction (4)

Biuret produced = $(103/120)*727 = 624 \text{ Kg/hr}$

Water produced in reaction (2) = $(18/60)*61978 = 18,593 \text{ Kg/hr}$

At reactor's exit (Urea = 34 %)

Flow rate of stream = $61,250/0.34 = 1,80,147 \text{ Kg/hr}$

NH_3 reacted in reaction (1) = $(34/60)*61977 = 35,120 \text{ Kg/hr}$

NH_3 produced in reaction (4) = $(17/120)*727 = 103 \text{ Kg/hr}$

So, NH_3 unreacted = $36968 - 35120 + 103 = 1951 \text{ Kg/hr}$

CO_2 reacted in reaction (1) = $(44/60)*61977 = 45450 \text{ Kg/hr}$

Therefore, CO_2 unreacted = $47842 - 45450 = 2392 \text{ Kg/hr}$

Now,

Flow rate of stream at reactor's exit – (flow rate of urea+ CO_2 + NH_3 +water+biuret) =

Flow rate of carbamate

$1,80,147 - (61,250 + 2392 + 1951 + 18,593 + 624) = 95,337 \text{ Kg/hr}$

4.1 REACTOR

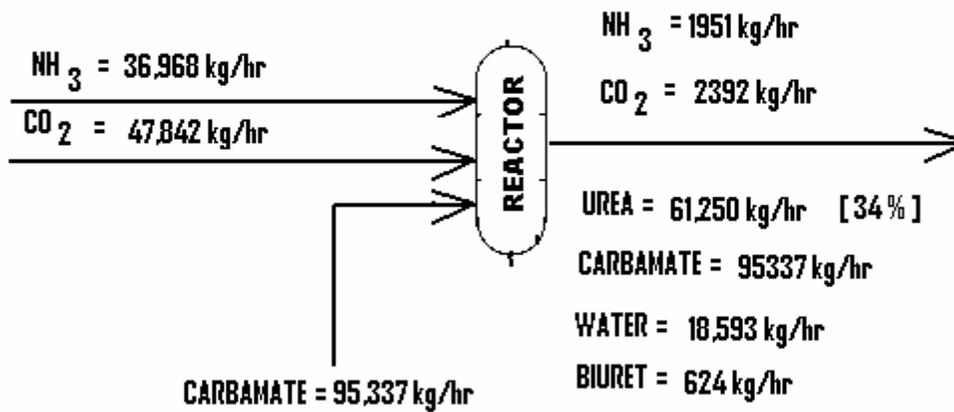


Fig 4.1 Flow of material across reactor

Table 4.1 Flow of material across reactor

INPUT			OUTPUT		
MATERIAL	FLOW RATE (Kg/hr)	%	MATERIAL	FLOW RATE (Kg/hr)	%
<u>Feed</u>			<u>Unreacted</u>		
NH ₃ (liq)	36968	43.59	NH ₃ (liq)	1951	1.08
CO ₂ (gas)	47842	56.41	CO ₂ (gas)	2392	1.32
Total	84810	100			
<u>Recycle</u>			<u>Products</u>		
CARBAMATE	95337	100	UREA	61250	34
			WATER	18593	10.32
			BIURET	624	0.36
			CARBAMATE	95,337	52.92
TOTAL	180147	100	TOTAL	180147	100

4.2 STRIPPER

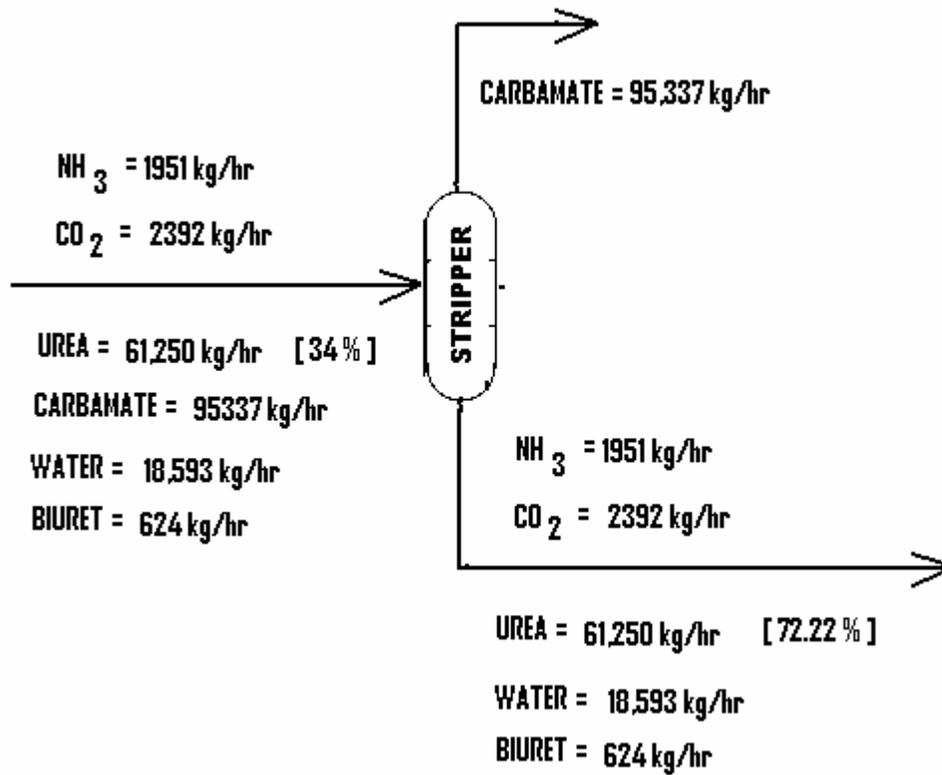


Fig 4.2 Flow of material across stripper

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.

Table 4.2 Flow of material across stripper

INPUT			OUTPUT		
MATERIAL	FLOW RATE (Kg/hr)	%	MATERIAL	FLOW RATE (Kg/hr)	%
NH3	1951	1.08	<u>Bottom product</u>		
CO2	2392	1.32	NH3	1951	2.30
CARBAMATE	95337	52.92	CO2	2392	2.82
UREA	61250	34	UREA	61250	72.22
WATER	18593	10.32	WATER	18593	21.9
BIURET	624	0.36	BIURET	624	0.76
			Total	84810	100
			<u>Top product</u>		
			Ammonium carbamate	95337	100
TOTAL	180147	100			

4.3 MEDIUM PRESSURE SEPARATOR

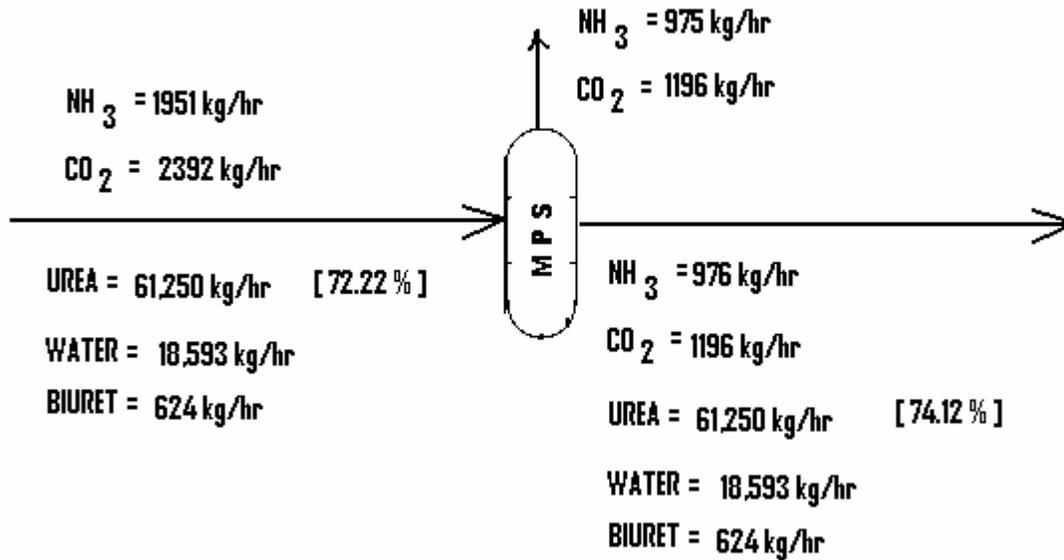


Fig 4.3 Flow of material across medium pressure separator

The amount of ammonia, carbon-di-oxide, water & biuret will remain constant as no reaction is taking place.

50 % of ammonia & carbon-di-oxide 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.

Table 4.3 Flow of material across medium pressure separator

INPUT			OUTPUT		
MATERIAL	FLOW RATE (Kg/hr)	%	MATERIAL	FLOW RATE (Kg/hr)	%
NH3	1951	2.3	NH3	976	1.18
CO2	2392	2.82	CO2	1196	1.44
UREA	61250	72.22	UREA	61250	74.11
WATER	18593	21.9	WATER	18593	22.49
BIURET	624	0.76	BIURET	624	0.78
			Total	82,639	100
			<u>Losses</u>		
			NH3	975	44.91
			CO2	1196	55.09
TOTAL	84812	100		2171	100

4.4 LOW PRESSURE SEPARATOR

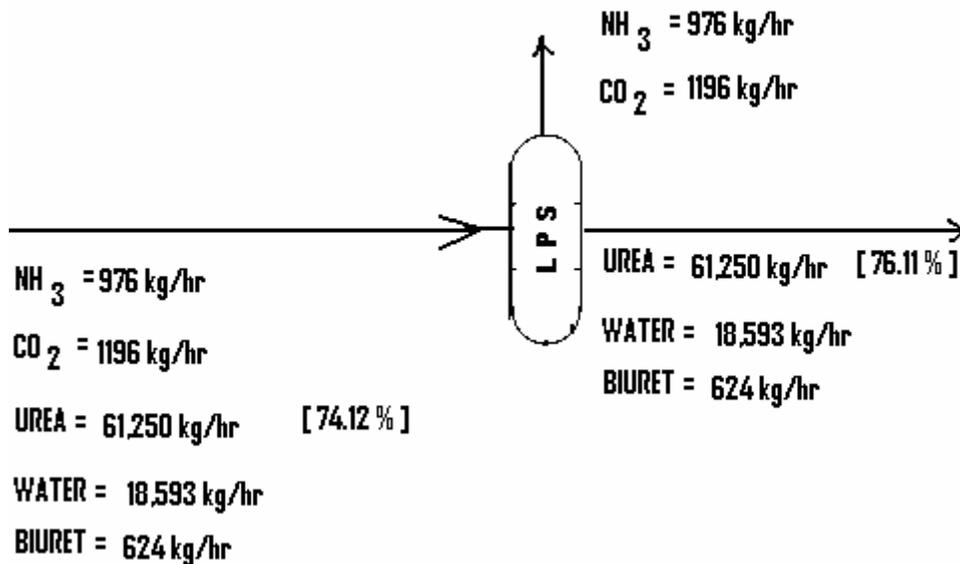


Fig 4.4 Flow of material across low pressure separator

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

Table 4.4 Flow of material across low pressure separator

INPUT			OUTPUT		
MATERIAL	FLOW RATE (Kg/hr)	%	MATERIAL	FLOW RATE (Kg/hr)	%
NH3	976	1.18			
CO2	1196	1.44			
UREA	61250	74.11	UREA	61250	76.11
WATER	18593	22.49	WATER	18593	23
BIURET	624	0.78	BIURET	624	0.79
			Total	80467	100
			<u>Losses</u>		
			NH3	975	44.91
			CO2	1196	55.09
TOTAL	82639	100		2171	100

4.5 VACUUM EVAPORATOR

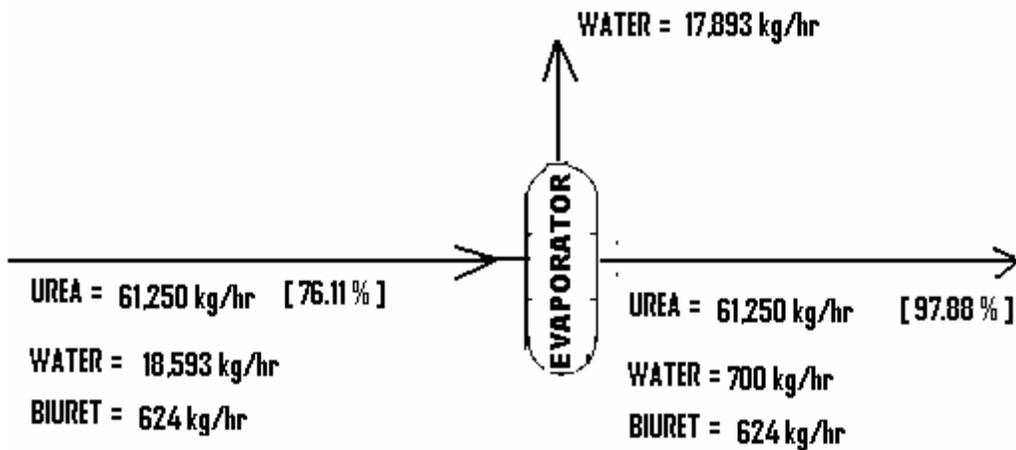


Fig 4.5 Flow of material across vacuum evaporator

Let x & y be the mass fractions of Urea in feed (F) & product (P) resp.

$$x = 0.7611 \quad (76.11 \%)$$

$$y = 0.9788 \quad (97.88 \%)$$

Making urea balance:

$$F \cdot x = P \cdot y$$

$$80467 \cdot 0.7611 = P \cdot 0.9788$$

$$P = 62574 \text{ Kg/hr}$$

Overall material balance gives:

$$F = P + E$$

$$80467 = 62574 + E$$

$$E = 17893 \text{ Kg/hr}$$

Table 4.5 Flow of material across vacuum evaporator

INPUT			OUTPUT		
MATERIAL	FLOW RATE (Kg/hr)	%	MATERIAL	FLOW RATE (Kg/hr)	%
UREA	61250	76.11	UREA	61250	97.88
WATER	18593	23.10	WATER	700	1.11
BIURET	624	0.79	BIURET	624	1.01
			Total	62574	100
			<u>Losses</u>		
			WATER	17893	100
TOTAL	80467	100			

4.6 PRILLING TOWER

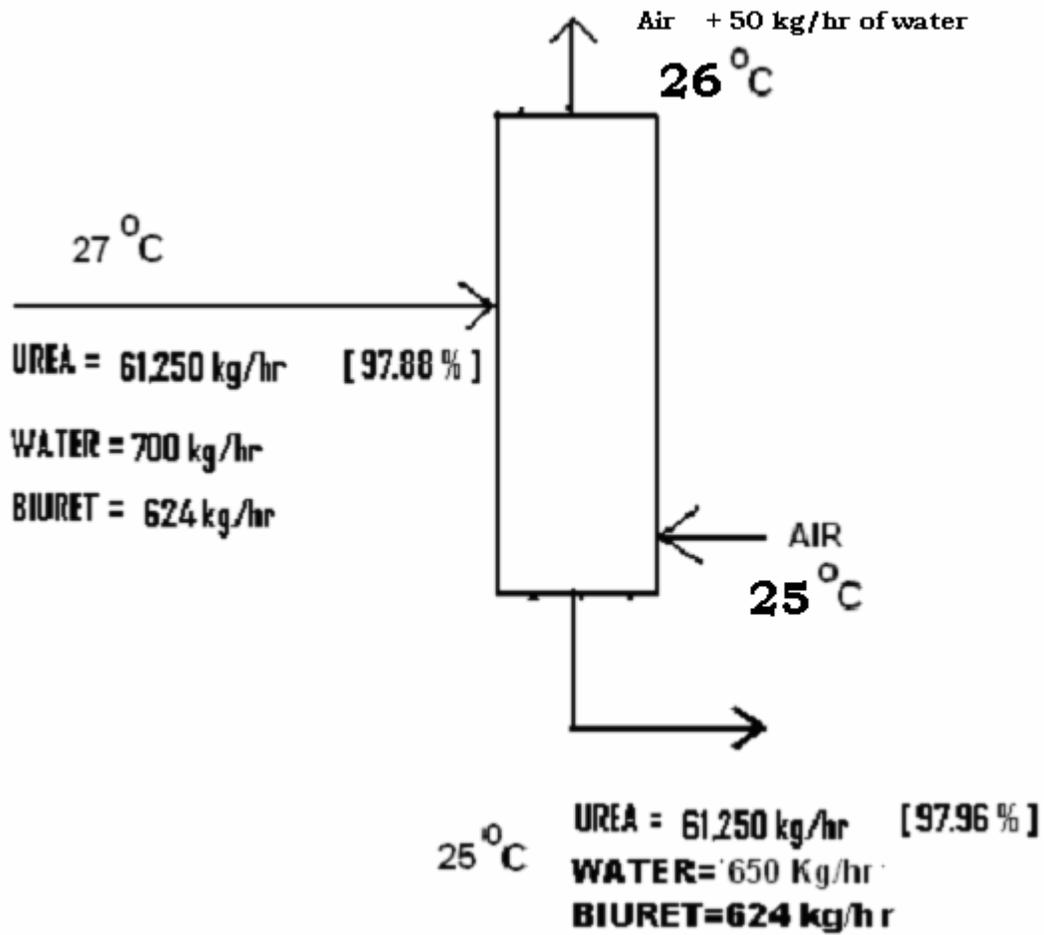


Fig 4.6 Flow of material across prilling tower

Let x & y be the mass fractions of Urea in feed (F) & product (P) resp.

$$x = 0.9788 \quad (97.88 \%)$$

$$y = 0.9796 \quad (97.96 \%)$$

Making urea balance:

$$F \cdot x = P \cdot y$$

$$62574 \cdot 0.9788 = P \cdot 0.9796$$

$$P = 62524 \text{ Kg/hr}$$

Table 4.6 Flow of material across prilling tower

INPUT			OUTPUT		
MATERIAL	FLOW RATE (Kg/hr)	%	MATERIAL	FLOW RATE (Kg/hr)	%
UREA	61250	97.88	UREA	61250	97.96
WATER	700	1.11	WATER	650	1.11
BIURET	624	1.01	BIURET	624	0.93
			Total	62,524	100
			<u>Losses</u>		
			WATER	50	100
TOTAL	62574	100			

Chapter 5

Energy balance

ENERGY BALANCE

Assumption : Datum temperature = 0°C

5.1 REACTOR [5]

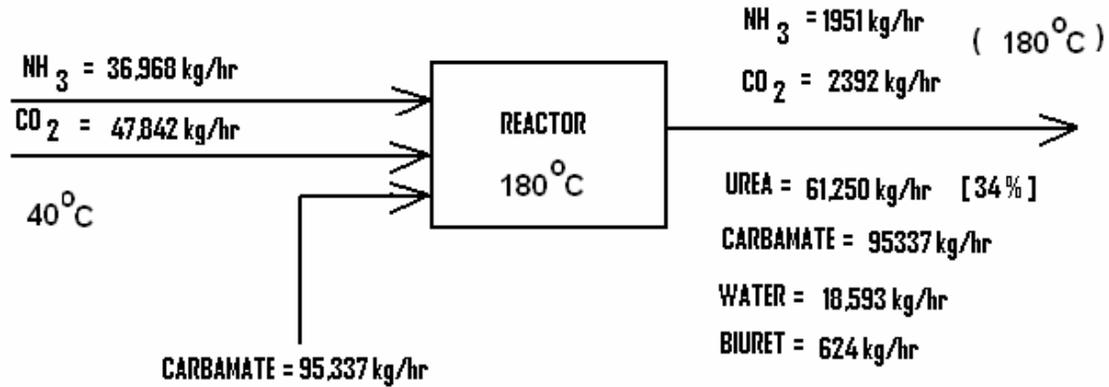


Fig 5.1 Energy flow across reactor

INLET STREAM

Material

specific heat at 40°C

NH₃

0.53 cal/gm °C = 2.219 Kj/Kg °C

CO₂

0.22 cal/gm °C = 0.9211 Kj/Kg °C

specific heat at 180°C

Carbamate

0.62 cal/gm °C = 2.596 Kj/Kg °C

Heat input

mCpΔt

NH₃ :

$3.6968 \times 10^4 \times 2.219 \times 40 = 0.328 \times 10^7$ Kj/hr

CO₂ :

$4.7843 \times 10^4 \times 0.9211 \times 40 = 0.176 \times 10^7$ Kj/hr

Carbamate:

$9.5336 \times 10^4 \times 2.24 \times 180 = 4.455 \times 10^7$ Kj/hr

Heat input = 4.959×10^7 Kj/hr

$$\Delta H_R = -31.32 \text{ Kcal/gm mol}$$

$$= -0.013 \times 10^7 \text{ Kj/Kmol of Urea formed.}$$

Amount of urea formed during the reaction = 1020.83 Kmol/hr

$$\Delta H_R = 1020.83 \times 0.013 \times 10^7 \text{ Kj/hr}$$

$$= 13.27 \times 10^7 \text{ Kj/hr}$$

OUTLET STREAM

Material	specific heat at 180°C	mol fractions (x)	Flow rate (Kmol/hr)
NH ₃	0.55 cal/gm °C = 39.15 Kj/Kmol°C	0.033	114.76
CO ₂	0.23 cal/gm °C = 42.37 Kj/Kmol °C	0.0158	54.36
Carbamate	0.62 cal/gm °C = 202.49 Kj/Kmol°C	0.354	1222.3
Urea	0.4828 cal/gm °C = 121.32 Kj/Kmol°C	0.296	1020.83
Water	1 cal/gm °C = 75.37 Kj/Kmol°C	0.299	1032.94
Biuret	183.8 Kj/Kmol°C	0.002	<u>6.07</u>
			Total = 3,451.3

$$C_p \text{ of mixture} = \sum x_i C_{p_i}$$

$$\text{So, } C_p = 0.033 \times 39.15 + 0.0158 \times 42.37 + 0.296 \times 121.32 + 0.354 \times 202.49 + 0.002 \times 183.8 + 0.299 \times 75.37 = 132.46 \text{ Kj/Kmol°C}$$

$$\text{So, heat carried by outlet stream} = m C_p \Delta t$$

$$= 3,451.3 \times 132.46 \times 180$$

$$= 8.229 \times 10^7 \text{ Kj/hr}$$

Heat input + ΔH_R - Heat output = rate of accumulation

$$4.959 \times 10^7 + 13.27 \times 10^7 - 8.229 \times 10^7 = \text{rate of accumulation}$$

$$\text{rate of accumulation} = 10 \times 10^7 \text{ Kj/hr}$$

Assumption : Cooling water at 25°C is used to remove heat from the reactor. The outlet is steam at an absolute pressure of 4.5 bar ($T_s = 147.9^\circ\text{C}$).

$$\text{So, heat gained by cooling water} = 10 \times 10^7 \text{ Kj/hr}$$

$$mC_p\Delta t + m\lambda = 10 \times 10^7 \text{ Kj/hr}$$

$$\text{or, } m(C_p\Delta t + \lambda) = 10 \times 10^7$$

$$m [4.187 \times (147.9-25) + 2120.6] = 10 \times 10^7 \text{ Kj/hr}$$

(Here $\lambda = 2120.6 \text{ kJ/kg}$ & $C_p = 4.187 \text{ KJ/kg } ^\circ\text{C}$)

$$m = 10^8 / 2635.18$$

$$m = 37,948 \text{ Kg/hr}$$

5.2 STRIPPER [6]

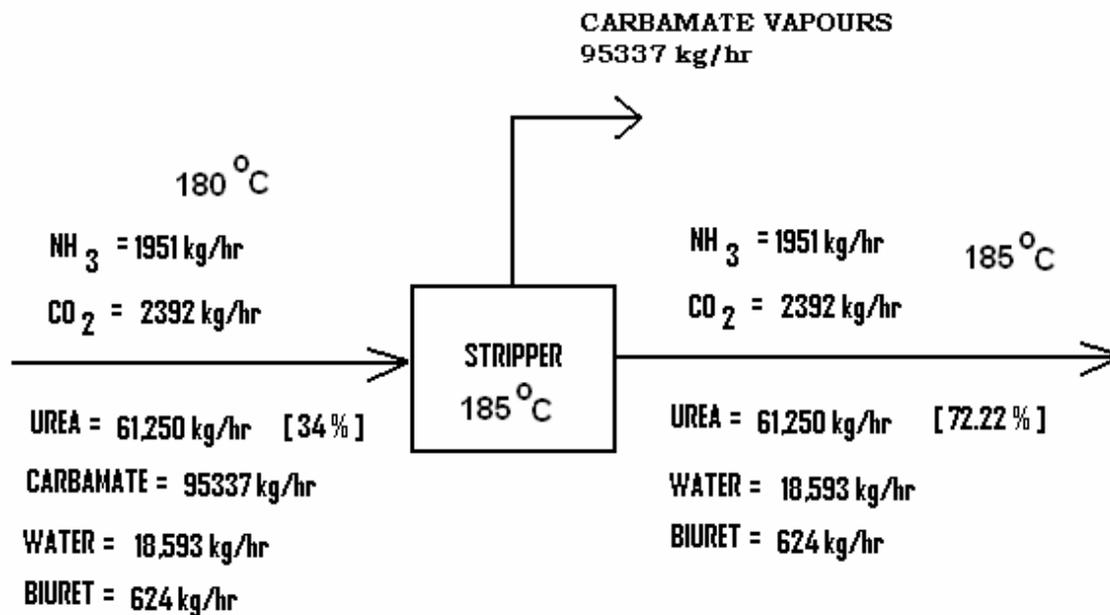


Fig 5.2 Energy flow across stripper

Total heat input = 8.229×10^7 Kj/hr

OUTLET STREAM

1) Liquid

Material	specific heat at [185°C] [5]	mol fractions (x)	Flow rate (Kmol/hr)
NH ₃	0.58 cal/gm °C = 41.31 Kj/Kmol°C	0.05	114.76
CO ₂	0.24 cal/gm °C = 44.22 Kj/Kmol °C	0.024	54.36
Urea	0.5385 cal/gm °C=135.3 Kj/Kmol°C	0.46	1020.83
Water	1 cal/gm °C = 75.37 Kj/Kmol°C	0.463	1032.94
Biuret	183.8 Kj/Kmol°C	0.003	<u>6.07</u>
			Total = 2,228.96

$$C_p \text{ of mixture} = \sum x_i C_{p_i}$$

$$\text{So, } C_p = 0.05 \times 41.31 + 0.024 \times 44.22 + 0.46 \times 135.3 + 0.003 \times 183.8 + 0.463 \times 75.37 = 100.81 \text{ Kj/Kmol}^\circ\text{C}$$

$$\begin{aligned} \text{So, heat carried by outlet stream} &= m C_p \Delta t \\ &= 2228.96 \times 100.81 \times 185 \\ &= 4.157 \times 10^7 \text{ Kj/hr} \end{aligned}$$

2) Vapour stream : Ammonium carbamate

Material	specific heat at [185°C]	Flow rate (Kmol/hr)
Carbamate	0.62 cal/gm °C = 202.49 Kj/Kmol°C	1222.3

For carbamate $\lambda = 210$ Kj/Kg

$$\text{So, heat carried by carbamate} = m C_p \Delta t + m \lambda$$

$$= 1222.3 \times 202.49 \times 185 + 95336 \times 210$$

$$= 6.581 \times 10^7 \text{ Kj/hr}$$

Here, steam at 24 atm is used ($T_s = 221.8^\circ\text{C}$).

$$\lambda \text{ of steam} = 1855.3 \text{ Kj/kg}$$

Heat supplied by steam = Heat output – Heat input

$$= (6.581 + 4.157 - 8.229) \times 10^7 \text{ Kj/hr}$$

$$m \lambda = 2.509 \times 10^7 \text{ Kj/hr}$$

$$m = 2.509 \times 10^7 / 1855.3$$

$$m = 13,523 \text{ Kg/hr}$$

5.3 CARBAMATE CONDENSER [6]

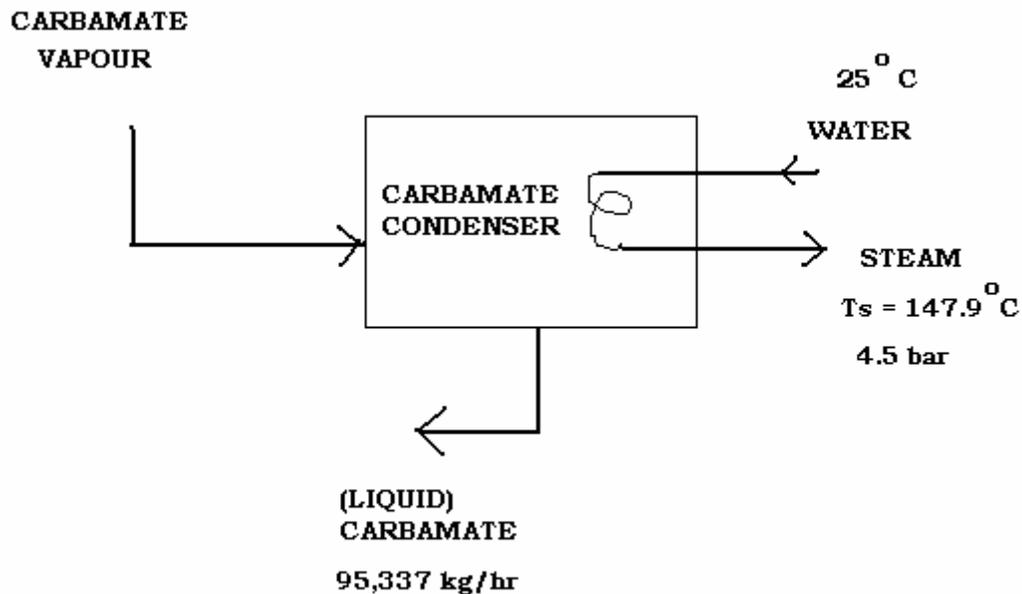


Fig 5.3 Energy flow across carbamate condenser

Energy balance

$$m_v \lambda_v = m_s C_p (T_s - 25) + m_s \lambda_s$$

Putting the values we get :

$$95337 \times 210 = m_s [4.187 \times (147.9 - 25) + 2120.6] \quad [\text{where } \lambda_s = 2120.6 \text{ kj/kg}]$$

So, $m_s = 7,597.5 \text{ kg/h}$

5.4 MEDIUM PRESSURE SEPARATOR [5]

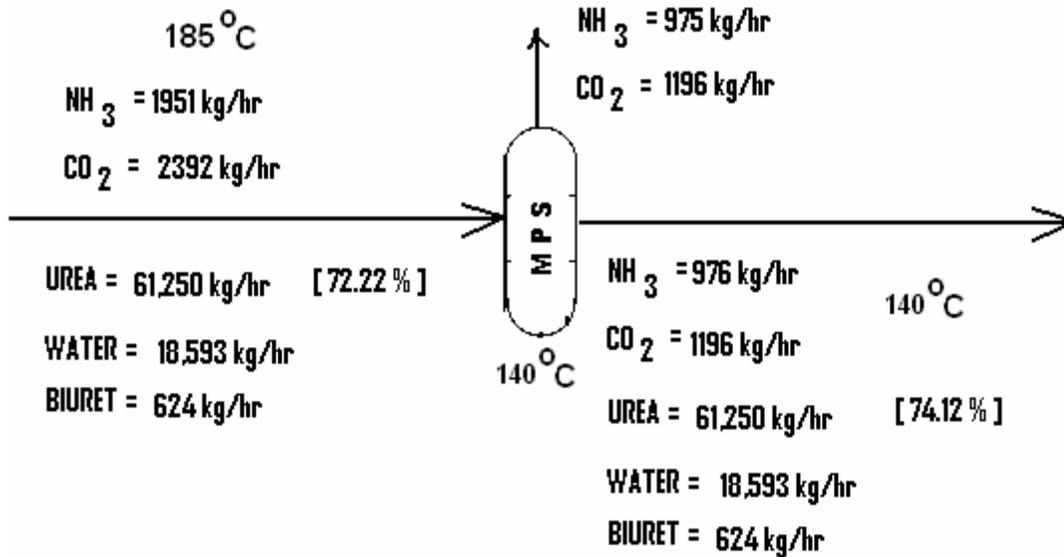


Fig 5.4 Energy flow across medium pressure separator

Heat input = $4.157 \times 10^7 \text{ KJ/hr}$

1) Liquid

OUTLET STREAM

Material	specific heat at [140°C]	mol fractions (x)	Flow rate (Kmol/hr)
NH3	0.54 cal/gm °C = 38.4 KJ/Kmol°C	0.027	57.4
CO2	0.23 cal/gm °C = 42.37 KJ/Kmol °C	0.0127	27.182
Urea	0.493 cal/gm °C=123.84 KJ/Kmol°C	0.476	1020.83
Water	1 cal/gm °C = 75.37 KJ/Kmol°C	0.4815	1032.94
Biuret	170.92 KJ/Kmol°C	0.0028	<u>6.07</u>
			Total = 2144.42

$$C_p \text{ of mixture} = \sum x_i C_{p_i}$$

So, $C_p = 0.027 \times 38.4 + 0.0127 \times 42.37 + 0.476 \times 123.84 + 0.0028 \times 170.92 + 0.4815 \times 75.37 \text{ KJ/Kmol}^\circ\text{C}$

$$= 97.29 \text{ Kj/Kmol}^\circ\text{C}$$

$$\begin{aligned} \text{heat output} &= 2144.42 \times 97.29 \times 140 \text{ Kj/hr} \\ &= 2.921 \times 10^7 \text{ Kj/hr} \end{aligned}$$

2) For gases escaping from the top

Material	λ at 140°C	mol fractions (x)	Flow rate (Kmol/hr)
NH ₃	$320 \text{ cal/gm }^\circ\text{C} = 22.777 \times 10^3 \text{ Kj/Kmol}^\circ\text{C}$	0.6785	57.35
CO ₂	$110 \text{ cal/gm }^\circ\text{C} = 20.265 \times 10^3 \text{ Kj/Kmol }^\circ\text{C}$	0.3215	<u>27.182</u>
Total =			84.53

$$\lambda \text{ of mixture} = \sum x_i \lambda_i$$

$$\begin{aligned} \text{So, } \lambda &= (0.6785 \times 22.777 + 0.321 \times 20.265) \times 10^3 \text{ Kj/Kmol}^\circ\text{C} \\ &= 21.969 \times 10^3 \text{ Kj/Kmol} \end{aligned}$$

Material	specific heat at [140°C]	mol fractions (x)	Flow rate (Kmol/hr)
NH ₃	$0.54 \text{ cal/gm }^\circ\text{C} = 38.4 \text{ Kj/Kmol}^\circ\text{C}$	0.6785	57.35
CO ₂	$0.23 \text{ cal/gm }^\circ\text{C} = 42.37 \text{ Kj/Kmol }^\circ\text{C}$	0.3215	<u>27.182</u>
Total =			84.53

$$C_p \text{ of mixture} = \sum x_i C_{p_i}$$

$$\begin{aligned} \text{So, } C_p &= 0.6785 \times 38.4 + 0.321 \times 42.37 \text{ Kj/Kmol}^\circ\text{C} \\ &= 39.676 \text{ Kj/Kmol}^\circ\text{C} \end{aligned}$$

$$\begin{aligned} \text{Heat escaping from the top} &= m (C_p \Delta t + \lambda) \\ &= 84.53 (39.676 \times 140 + 21.969 \times 10^3) \\ &= 0.2327 \times 10^7 \text{ Kj/hr} \end{aligned}$$

Assumption : Cooling water enters at 25°C & leaves at 50°C.

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

$$= (4.157 - 2.921 - 0.2327) \times 10^7 \text{ Kj/hr}$$

$$mC_p\Delta t = 1.00 \times 10^7 \text{ Kj/hr}$$

$$m = 1.00 \times 10^7 / (4.187 \times 25)$$

$$m = 95,533.8 \text{ Kg/hr}$$

5.5 LOW PRESSURE SEPARATOR [5]

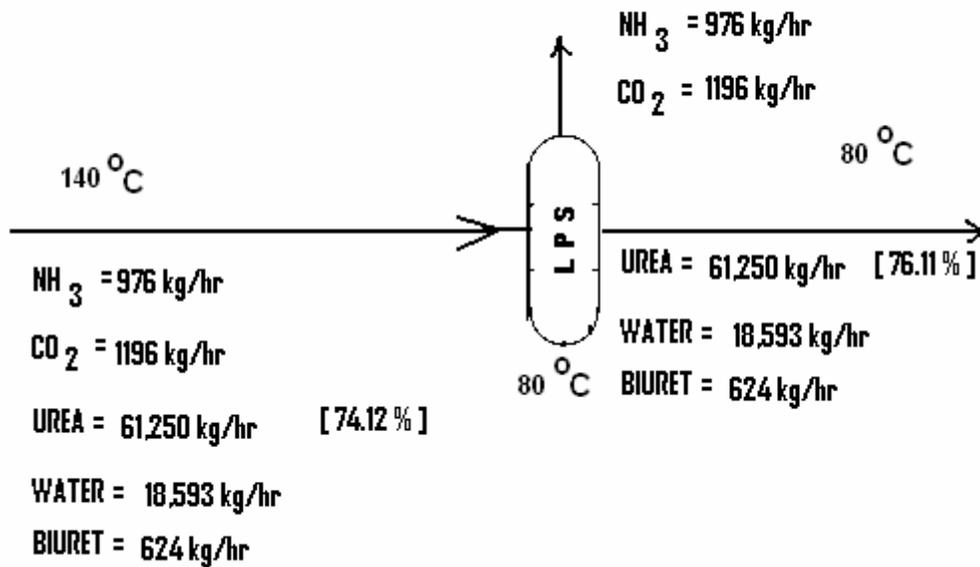


Fig 5.5 Energy flow across low pressure separator

Heat input = 2.921×10^7 Kj/hr

1) Liquid

OUTLET STREAM

Material	specific heat at [80°C]	mol fractions (x)	Flow rate (Kmol/hr)
Urea	0.429 cal/gm °C = 107.76 Kj/Kmol°C	0.496	1020.83
Water	1 cal/gm °C = 75.37 Kj/Kmol°C	0.5	1032.94
Biuret	149 Kj/Kmol°C	0.004	<u>6.07</u>
			Total = 2059.8

$$C_p \text{ of mixture} = \sum x_i C_{p_i}$$

$$\text{So, } C_p = 0.496 \times 107.76 + 0.5 \times 75.37 + 0.004 \times 149 \text{ Kj/Kmol}^\circ\text{C}$$

$$= 91.73 \text{ Kj/Kmol}^\circ\text{C}$$

$$\text{heat output} = 2059.8 \times 91.73 \times 80$$

$$= 1.51 \times 10^7 \text{ Kj/hr}$$

2) For gases escaping from the top

Material	λ at 140°C	mol fractions (x)	Flow rate (Kmol/hr)
NH ₃	260 cal/gm °C = 18.51 x 10 ³ Kj/Kmol°C	0.679	57.41
CO ₂	85 cal/gm °C = 15.66 x 10 ³ Kj/Kmol °C	0.321	<u>27.182</u>
			Total = 84.59

$$\lambda \text{ of mixture} = \sum x_i \lambda_i$$

$$\text{So, } \lambda = (0.679 \times 18.51 + 0.321 \times 15.66) \times 10^3 \text{ Kj/Kmol}^\circ\text{C}$$

$$= 17.6 \times 10^3 \text{ Kj/Kmol}$$

Material	specific heat at [140°C]	mol fractions (x)	Flow rate (Kmol/hr)
NH ₃	0.52 cal/gm °C = 37.013 Kj/Kmol°C	0.027	57.41
CO ₂	0.21 cal/gm °C = 38.69 Kj/Kmol °C	0.0127	<u>27.182</u>
			Total = 84.59

$$C_p \text{ of mixture} = \sum x_i C_{p_i}$$

$$\text{So, } C_p = 0.679 \times 37.013 + 0.321 \times 38.69 \text{ Kj/Kmol}^\circ\text{C}$$

$$= 37.55 \text{ Kj/Kmol}^\circ\text{C}$$

$$\text{Heat escaping from the top} = m (C_p \Delta t + \lambda)$$

$$= 84.59 (37.55 \times 80 + 17.7 \times 10^3)$$

$$= 0.1743 \times 10^7 \text{ Kj/hr}$$

Assumption : Cooling water enters at 25°C & leaves at 50°C.

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

$$= (2.921 - 1.51 - 0.1743) \times 10^7 \text{ Kj/hr}$$

$$m C_p \Delta t = 1.2367 \times 10^7 \text{ Kj/hr}$$

$$m = 1.2367 \times 10^7 / (4.187 \times 25)$$

$$m = 1,18,146 \text{ Kg/hr}$$

5.6 EVAPORATOR [5]

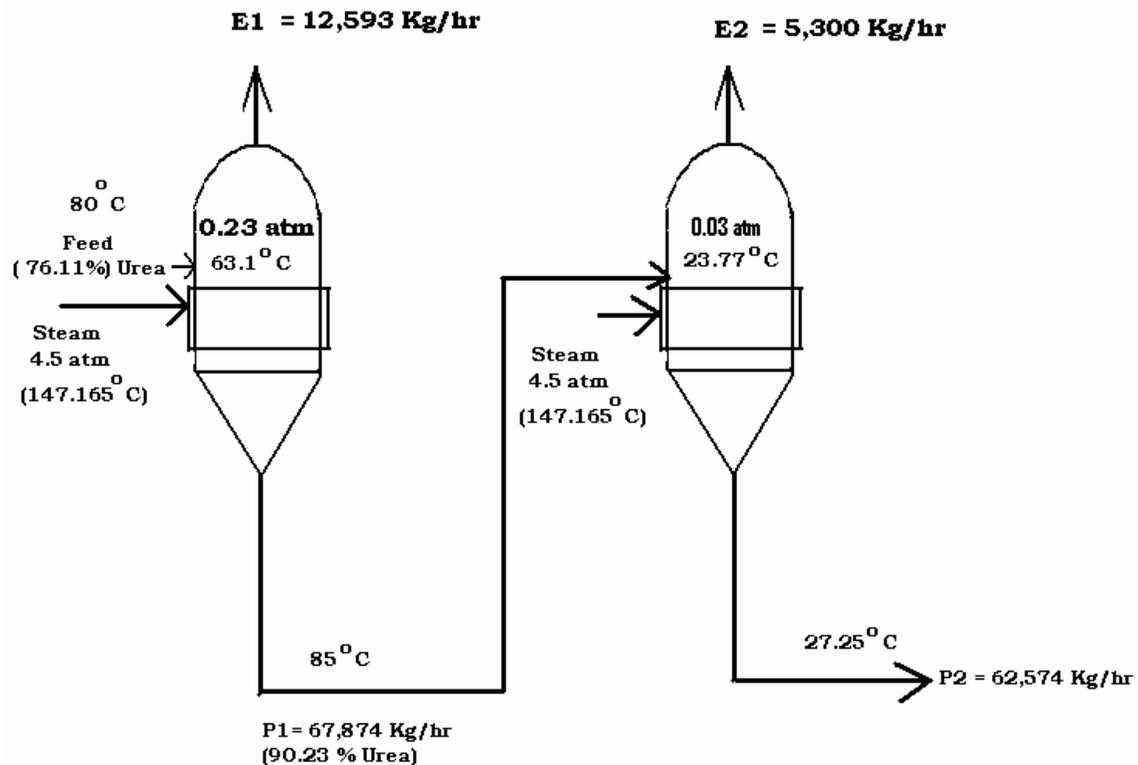


Fig 5.6 Energy flow across evaporator

For product stream coming out of 1st evaporator:

Material	specific heat at [85°C]	mol fractions (x)	Flow rate (Kmol/hr)
Urea	0.435cal/gm °C=109.28 Kj/Kmol°C	0.75	1020.83
Water	1 cal/gm °C = 75.37 Kj/Kmol°C	0.245	333.33
Biuret	149 Kj/Kmol°C	0.005	<u>6.06</u>
			Total = 1360.223

$$C_p \text{ of mixture} = \sum x_i C_{p_i}$$

$$\text{So, } C_p = 0.75 \times 109.28 + 0.245 \times 75.37 + 0.005 \times 149 \text{ Kj/Kmol}^\circ\text{C}$$

$$= 101.17 \text{ Kj/Kmol}^\circ\text{C}$$

$$mC_p\Delta t = 1360.223 \times 101.17 \times 85$$

$$= 1.17 \times 10^7 \text{ Kj/hr}$$

Heat balance

1st evaporator :

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

$$\text{Heat input (feed)} + S_1 \lambda_{s1} = E_1 H_{E1} + \text{energy of the bottom product}$$

$$1.537 \times 10^7 + S_1 \times 2123.2 = 12,593 \times 2614.97 + 1.17 \times 10^7$$

$$S_1 = 13,781 \text{ Kg/hr}$$

2nd evaporator :

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

$$\text{Heat input (feed)} + S_2 \lambda_{s2} = E_2 H_{E2} + \text{energy of the bottom product}$$

$$1.17 \times 10^7 + S_2 \times 2123.2 = 5,300 \times 2545.7 + 1065.8 \times 96.84 \times 27$$

$$S_2 = 2,157 \text{ Kg/hr}$$

5.7 PRILLING TOWER [5]

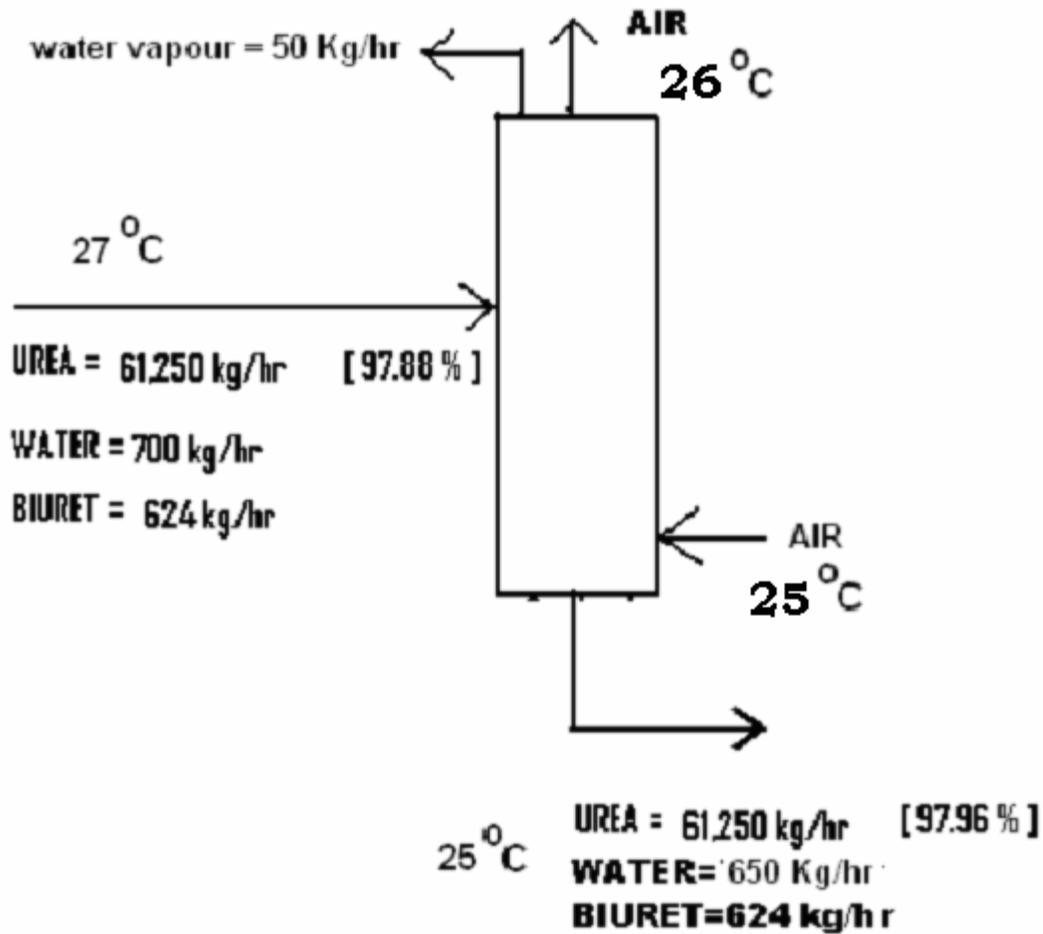


Fig 5.7 Energy balance across prilling tower

$$\text{Heat input} = 1065.8 \times 96.84 \times 27 = 0.279 \times 10^7 \text{ Kj/hr}$$

OUTLET STREAM

Material	specific heat at [30°C]	mol fractions (x)	Flow rate (Kmol/hr)
Urea	0.3758 cal/gm °C = 94.41 Kj/Kmol°C	0.96	1020.83
Water	1 cal/gm °C = 75.37 Kj/Kmol°C	0.034	36.11
Biuret	133.02 Kj/Kmol°C	0.006	<u>6.07</u>
			Total = 1063.01

$$C_p \text{ of mixture} = \sum x_i C_{p_i}$$

$$\text{So, } C_p = 0.96 \times 94.41 + 0.034 \times 75.37 + 0.006 \times 133.02 \text{ Kj/Kmol°C}$$

$$= 93.99 \text{ Kj/Kmol°C}$$

$$\begin{aligned}\text{Heat output} &= 1063.01 \times 93.99 \times 25 \\ &= 0.250 \times 10^7 \text{ Kj/hr}\end{aligned}$$

Assuming, humidity of air at 25°C = 0.01

Heat carried away by air = heat input – heat output

$$(mC_p\Delta t)_{\text{dry air}} = (0.279 - 0.250) \times 10^7$$

$$m = .029 \times 10^7 / (1.009 \times 1)$$

$$m = 28,74,133 \text{ Kg/hr}$$

So, flow rate of air = 28,74,133 Kg/hr

Chapter 6

Equipment design:
Design of reactor & evaporator

EQUIPMENT DESIGN

6.1 REACTOR DESIGN [7]

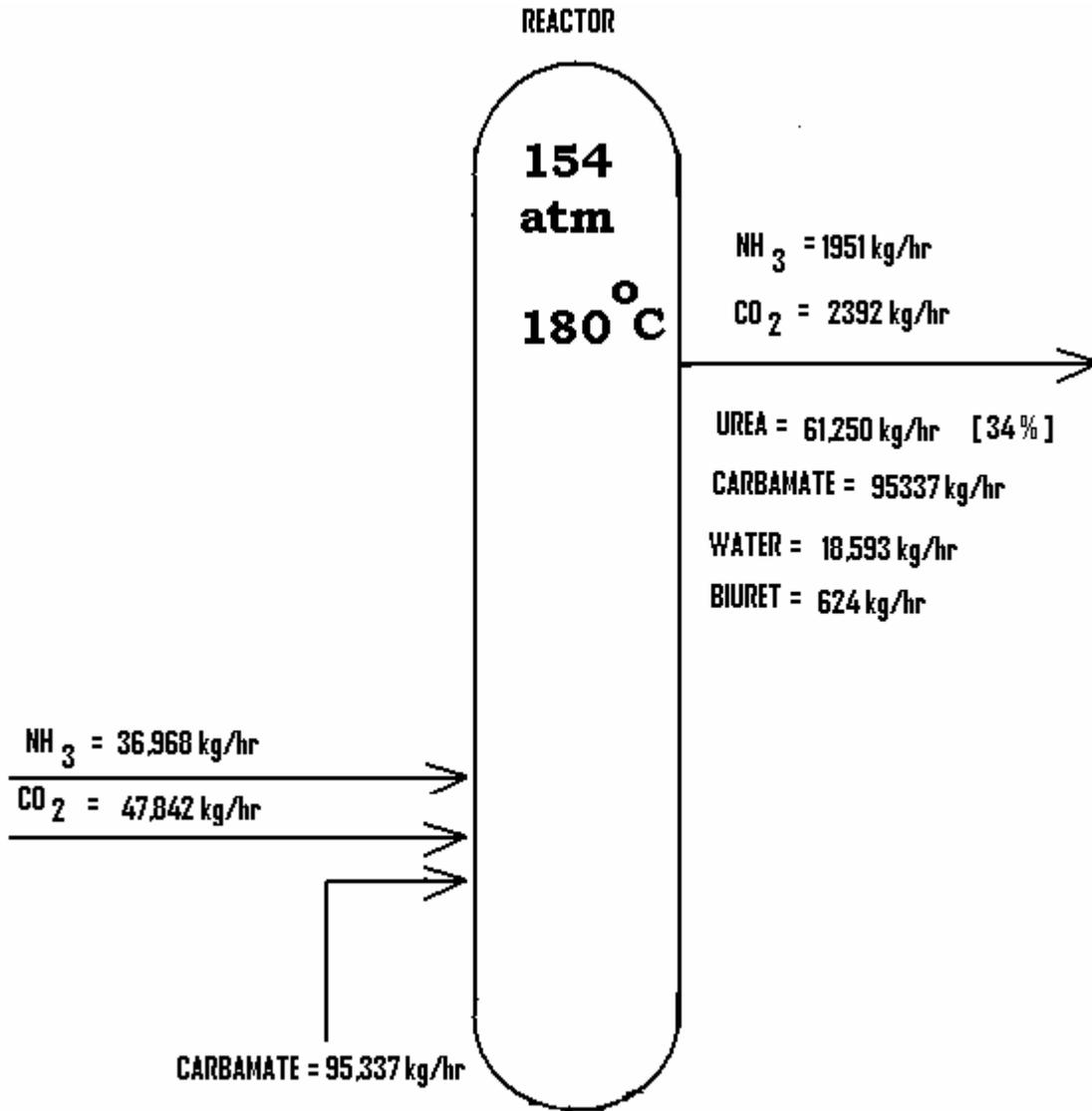


Fig 6.1.1 Urea reactor

From fig-6.1.2, for NH_3 to CO_2 ratio of 2 corresponding yield of urea is 50%. From fig-6.1.3, for 50% yield of urea the residence time is 40 min.

$$t = 40 \text{ min}$$

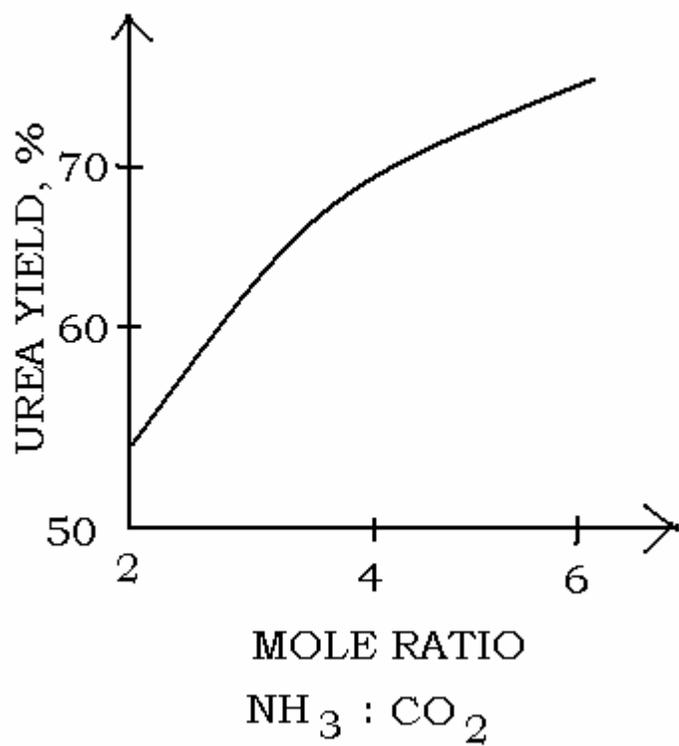


Fig 6.1.2 Graph of % urea yield Vs molar ratio of NH_3 Vs CO_2

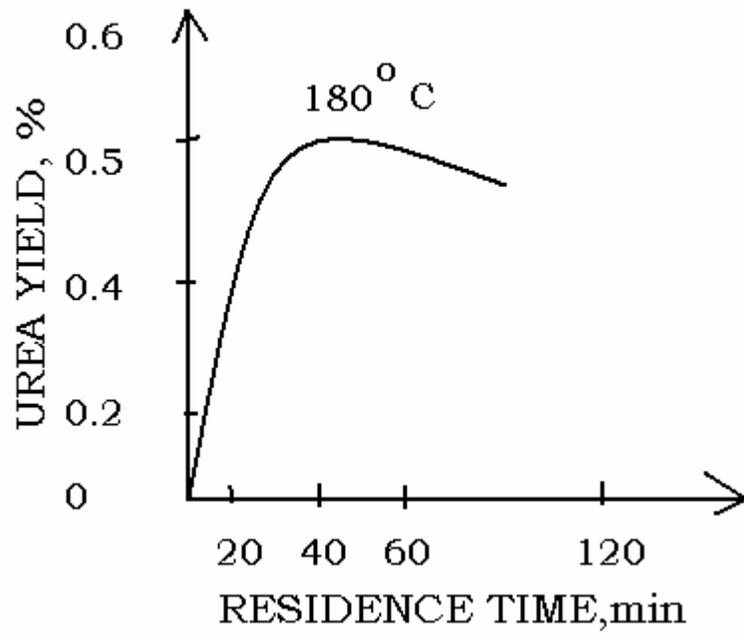


Fig 6.1.3 Graph of % urea yield Vs residence time.

Now, $t = V/F$ (Ref : Chemical reactor design-Peter Harriott, Pg-90)

Where, $t =$ residence time

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

$V =$ Volume of the reactor in m^3 .

Now,

Density of liquid $\text{NH}_3 = 618 \text{ Kg/ m}^3$ (Ref: J H Perry)

Density of CO_2 gas at $40^\circ\text{C} = 277.38 \text{ Kg/ m}^3$ (density= PM/RT ; $P=162 \text{ atm}$, $T=313 \text{ K}$)

Density of Carbamate = 1600 Kg/ m^3 (Ref: <http://www.inorganics.basf.com>)

So, NH_3 flowing into the reactor = $36,968/618 = 59.82 \text{ m}^3/\text{hr}$

CO_2 flowing into the reactor = $47,842/277.38 = 172.478 \text{ m}^3/\text{hr}$

Carbamate flowing into the reactor = $95,337/1600 = 59.59 \text{ m}^3/\text{hr}$

Total flow rate into the reactor = $59.82 + 172.478 + 59.59$
 $= 291.89 \text{ m}^3/\text{hr}$

Since, $t = V/F$

Therefore, $V = t \times F$

$$= (40 \times 291.89)/60$$

$$V = 194.59 \text{ m}^3$$

For design purpose $V = 195 \text{ m}^3$

Now, volume of the reactor = $(\pi D^2/4)L = 195$ [D = 2.5 m (given)]

(Ref : Equipment Design-Brownell & Young;Pg-80)

or, $L = 195 \times 4/(3.14 \times 2.5^2)$

$$= 39.75 \text{ m}$$

or, $L = 40 \text{ m}$

6.1.1 THICKNESS OF SHELL [8]

Data available :

Temperature inside the reactor = 180 °C

Pressure inside the reactor = 154 atm

Material of construction :

Low alloy carbon steel (Ref : Fertilizer manufacture- M E Pozin)

Material specification :

IS : 2002-1962 Grade 2B (Ref: B C Bhattacharya, Table-A1,Pg-261)

Allowable stress = $1.18 \times 10^8 \text{ N/m}^2$

Diameter of the reactor = 2.5 m

(Ref : Fertilizer manufacture- M E Pozin,Pg-263;for plants having capacity of 4,50,000 tons/yr)

Now, volume of the reactor = $(\pi D^2/4)L = 195$

(Ref : Equipment Design-Brownell & Young;Pg-80)

or, $L = 195 \times 4 / (3.14 \times 2.5^2)$

= 39.75 m

or, L = 40 m

Also, $L/D = 40/2.5 = 16$ which is consistent with the actual ratio which is between 14 to 20.

Now,

$$t = pD_i / (2fj - p) \quad [\text{Ref : Equipment design- M V Joshi,Pg-96}]$$

where, t = thickness of the shell

D_i = internal diameter

J = joint efficiency

p = design pressure

f = permissible stress

internal pressure = 154 atm = $1.56 \times 10^7 \text{ N/m}^2$

Design pressure $p = (10 \% \text{ extra})$

$$= 1.1 \times 1.56 \times 10^7 \text{ N/m}^2$$

$$= 1.716 \times 10^7 \text{ N/m}^2$$

$J = 1$ [For class 1 pressure vessels , BIS-2825]

$f = 1.18$

$D_i = 2.5 \text{ m}$

So, $t = 1.716 \times 10^7 \times 2.5 / (2 \times 1.18 \times 10^8 \times 1 - 1.716 \times 10^7)$

$$t = 0.196 \text{ m}$$

$$= 196 \text{ mm}$$

or, $t = 200 \text{ mm}$

6.1.2 HEAD DESIGN [10]

For 2 : 1 ellipsoidal dished head

$t_h = pDV/2fJ$ [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, $t_h = 1.716 \times 10^7 \times 2.5 \times 1.5 / (2 \times 1.18 \times 10^8 \times 1)$

$$t = 0.273 \text{ m}$$

or, $t = 273 \text{ mm}$

or, $t = 300 \text{ mm}$

6.1.3 DIAMETER OF PIPES [11]

We know that,

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

For inlet pipes:

$$\begin{aligned}(D_i)_{NH_3} &= 0.0144 \times (36968)^{0.45}/(618)^{0.32} \\ &= 0.2093 \text{ m} \\ &= 8.24 \text{ inch}\end{aligned}$$

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

$$\begin{aligned}\text{NPS} &= 10 \\ \text{Schedule no.} &= 60 \\ \text{OD} &= 10.75 \text{ inch} \\ \text{ID} &= 9.75 \text{ inch}\end{aligned}$$

$$\begin{aligned}(D_i)_{CO_2} &= 0.0144 \times (47842)^{0.45}/(277.38)^{0.32} \\ &= 0.3037 \text{ m} \\ &= 11.95 \text{ inch}\end{aligned}$$

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

$$\begin{aligned}\text{NPS} &= 12 \\ \text{Schedule no.} &= 30 \\ \text{OD} &= 12.75 \text{ inch} \\ \text{ID} &= 12.09 \text{ inch}\end{aligned}$$

$$\begin{aligned}(D_i)_{\text{carbamate}} &= 0.0144 \times (95337)^{0.45}/(1600)^{0.32} \\ &= 0.2364 \text{ m} \\ &= 9.307 \text{ inch}\end{aligned}$$

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

$$\begin{aligned}\text{NPS} &= 10 \\ \text{Schedule no.} &= 60 \\ \text{OD} &= 10.75 \text{ inch} \\ \text{ID} &= 9.75 \text{ inch}\end{aligned}$$

$$\begin{aligned}
 (D_i)_{\text{outlet stream}} &= 0.0144 \times (1,80,147)^{0.45} / (1283.97)^{0.32} \\
 &= 0.3378 \text{ m} \\
 &= 13.29 \text{ inch}
 \end{aligned}$$

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

$$\begin{aligned}
 \text{NPS} &= 16 \\
 \text{Schedule no.} &= 30 \\
 \text{OD} &= 16 \text{ inch} \\
 \text{ID} &= 15.25 \text{ inch}
 \end{aligned}$$

6.1.4 SKIRT SUPPORT FOR REACTOR [10]

$$\begin{aligned}
 \text{Wt. of the reactor} &= \text{wt. of material of construction} + \text{weight of the contents of the reactor} \\
 &= \Pi D t L \rho + \text{weight of the contents of the reactor} \\
 &= \Pi \times 2.5 \times 0.2 \times 44 \times 7857 + 1,80,147 \quad [\rho = 7857 \text{ kg/m}^3]
 \end{aligned}$$

$$\begin{aligned}
 W &= 722 \text{ tons} \\
 &= 7.085 \times 10^7 \text{ N}
 \end{aligned}$$

Material of construction :

IS : 2002-1962 Grade 2B

(Ref: B C Bhattacharya, Table-A1,Pg-261)

$$\text{Allowable tensile stress} = 1.18 \times 10^8 \text{ N/m}^2$$

$$\text{Yield stress} = 2.55 \times 10^8 \text{ N/m}^2$$

$$\text{Wind pressure upto} = 1300 \text{ N/m}^2$$

Stress due to dead weight:

$$f_d = \frac{\sum W}{(\Pi D_{ok} t_{sk})}$$

where,

$$f_d = \text{Stress}$$

$$\sum W = \text{Dead wt of vessel}$$

$$D_{ok} = \text{Outside diameter of the skirt}$$

$$t_{sk} = \text{thickness of skirt}$$

$$f_d = 7.085 \times 10^7 / (3.14 \times 2.5 \times t_{sk})$$

$$= 9.025 \times 10^6 / t_{sk} \text{ N/m}^2$$

Assuming height of skirt = 5 m

$$f_{wb} = M_w / Z = 4M_w / (\pi D_{ok}^2 t_{sk}) \quad [\text{Ref: equation 13.22, M V Joshi}]$$

$$M_w = P_{lw} (h_1/2) + P_{uw} (h_1 + h_2/2)$$

$$P_{lw} = kP_1 h_1 D_o \quad [\text{where } k = 0.7]$$

$$P_{uw} = kP_2 h_2 D_o$$

$$f_{wb} = 0.7 \times 1300 \times 20 \times 2.5 \times (20/2) \times 4 / (3.14 \times 2.5^2 \times t_{sk}) +$$

$$0.7 \times 1300 \times 20 \times 2.5 \times 30 \times 4 / (3.14 \times 2.5^2 \times t_{sk})$$

$$f_{wb} = 3.709 \times 10^5 / t_{sk} \text{ N/m}^2$$

Stress due to seismic load :

$$f_{sb} = (2/3) \times CWH / (\pi R_{ok} t_{sk}) \quad [\text{here } C = 0.08]$$

$$f_{sb} = (2/3) \times (0.08 \times 7.085 \times 10^7 \times 40) / (3.14 \times (2.5/2) \times t_{sk})$$

$$f_{sb} = 3.85 \times 10^7 / t_{sk} \text{ N/m}^2$$

$$\text{Maximum tensile stress} = f_{sb} - f_d$$

$$= 3.85 \times 10^7 / t_{sk} - 0.9025 \times 10^7 / t_{sk}$$

$$(f_t)_{\text{Max}} = 2.9475 \times 10^7 / t_{sk}$$

$$\text{Now, permissible tensile stress} = 1.18 \times 10^8 \text{ N/m}^2$$

$$t_{sk} = 2.9475 \times 10^7 / (1.18 \times 10^8)$$

$$= 0.2498 \text{ mm}$$

Maximum compressive stress : [Ref : Equation 13.29, Pg-326, M V Joshi]

$$(f_c)_{\text{Max}} = 3.85 \times 10^7 / t_{sk} + 0.9025 \times 10^7 / t_{sk}$$

$$(f_c)_{\text{Max}} = 4.7525 \times 10^7 / t_{sk}$$

$$(f_c)_{\text{Permissible}} \leq \frac{1}{3} \text{ Y.P}$$

$$\leq \frac{1}{3} \times 2.55 \times 10^8 \text{ N/m}^2$$

or, $t_{sk} = 4.7525 \times 10^7 / (0.85 \times 10^8)$

$$t_{sk} = 559.1 \text{ mm}$$

So, thickness to be used = 600 mm

6.2 EVAPORATOR DESIGN [5]

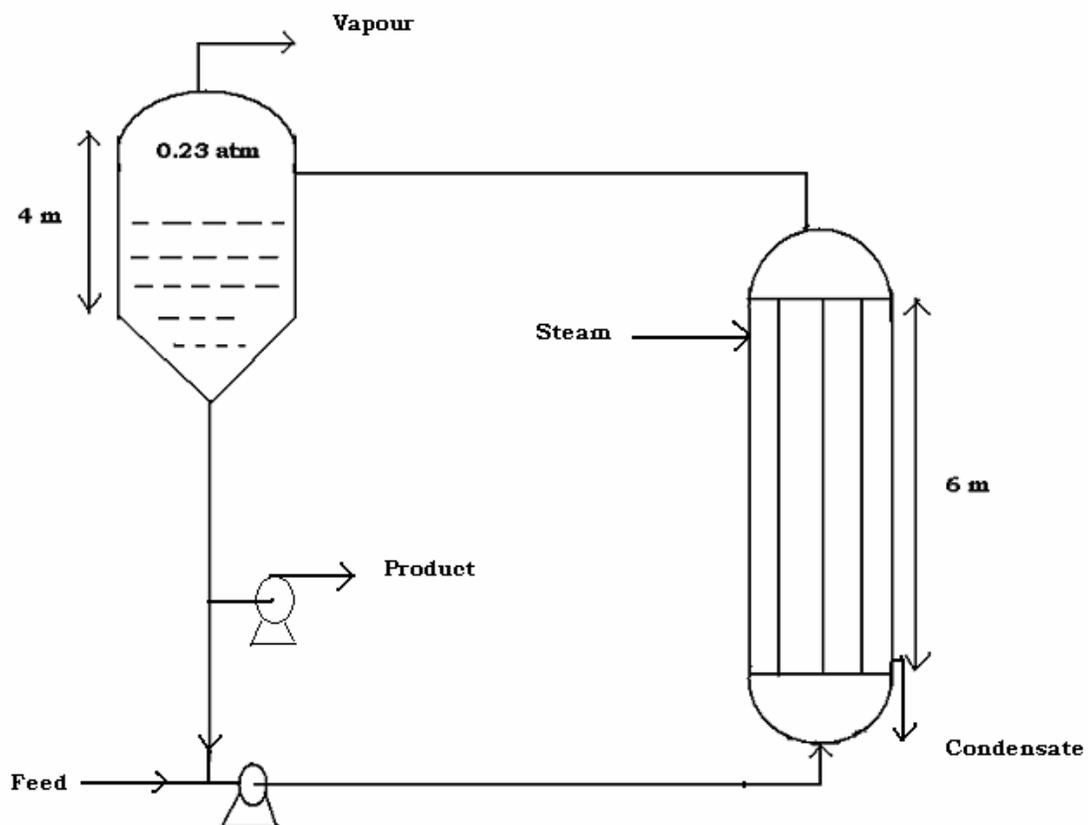


Fig 6.2.1 Urea evaporator (climbing film long tube vertical evaporator)

Vapour space pressure = 0.23 atm

Vapour space temperature = 63.1 °C

BPR = 21.9 °C

[Ref : Kirk Othmer, Encyclopedia of chemical technology, Vol-21]

Boiling point of liquid = 85 °C

For product stream coming out of 1st evaporator:

Material	specific heat at [85°C]	mol fractions (x)	Flow rate (Kmol/hr)
Urea	0.435cal/gm °C=109.28 Kj/Kmol°C	0.75	1020.83
Water	1 cal/gm °C = 75.37 Kj/Kmol°C	0.245	333.33
Biuret	149 Kj/Kmol°C	0.005	<u>6.06</u>
			Total = 1360.223

$$C_p \text{ of mixture} = \sum x_i C_{p_i}$$

$$\text{So, } C_p = 0.75 \times 109.28 + 0.245 \times 75.37 + 0.005 \times 149 \text{ Kj/Kmol}^\circ\text{C}$$
$$= 101.17 \text{ Kj/Kmol}^\circ\text{C}$$

$$mC_p\Delta t = 1360.223 \times 101.17 \times 85$$
$$= 1.17 \times 10^7 \text{ Kj/hr}$$

Heat balance

1st evaporator :

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

$$\text{Heat input (feed)} + S_1 \lambda_{s1} = E_1 H_{E1} + \text{energy of the bottom product}$$

For steam at 147.165 °C, $\lambda_{s1} = 2123.2 \text{ kj/kg}$

Putting the values we get

$$1.537 \times 10^7 + S_1 \times 2123.2 = 12,593 \times 2614.97 + 1.17 \times 10^7$$
$$S_1 = 13,781 \text{ Kg/hr}$$

$$\text{Economy} = 12,593/13781 = 0.914$$

Now,

U_1 value is obtained from fig-6.2.2. At 63.1°C (145.58°F) the value of U_1 is 270 Btu/hr.sq.ft. $^\circ\text{F}$. Multiplying this value by 5.6783 gives the value of U_1 in $\text{W}/\text{m}^2\text{K}$.

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

$$\begin{aligned} \Delta T_1 &= (\Delta T)_{\text{app}} - \text{BPR}_1 \\ &= 147.165 - 63.1 - 21.9 \\ &= 62.165^\circ\text{C} \end{aligned}$$

$$\begin{aligned} \text{So, } A_1 &= 13,781 \times 2123.2 / 1533 \times 62.165 \\ &= 307.03 \text{ m}^2 \end{aligned}$$

similarly,

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

$$\begin{aligned} \Delta T_2 &= (\Delta T)_{\text{app}} - \text{BPR}_2 \\ &= 147.165 - 23.77 - 3.48 \\ &= 119.915^\circ\text{C} \end{aligned}$$

$$\begin{aligned} \text{So, } A_2 &= 2157 \times 2123.2 / 738 \times 119.915 \\ &= 51.75 \text{ m}^2 \end{aligned}$$

(Ref: values of U_1 & U_2 from Perry's handbook,10-35)

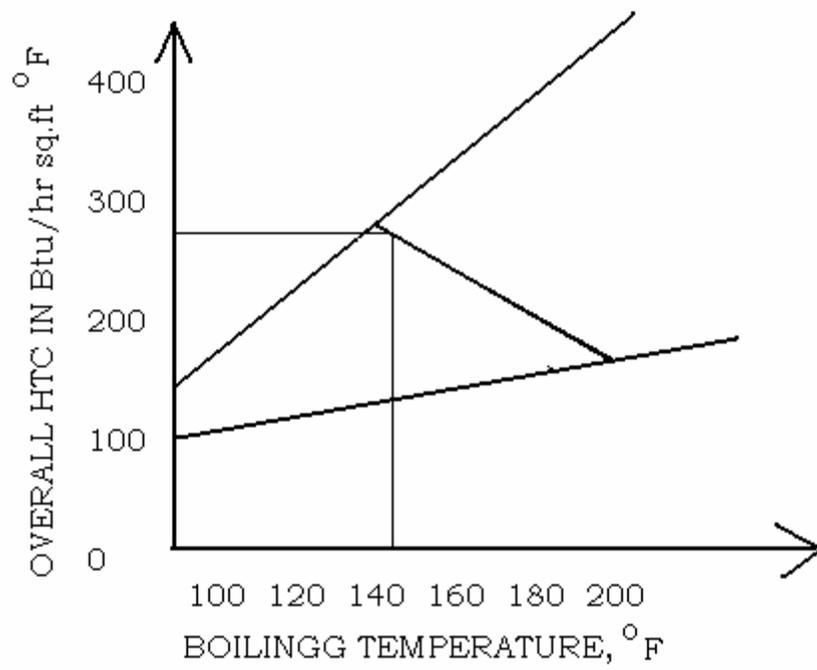


Fig 6.2.2 Graph to find out heat transfer co-efficient

6.2.1 DESIGN :

Assuming :

Length = 6 m [Ref: M V Joshi,Pg-220]

Tube OD = 1 inch [Table-10,PHT,D Q Kern]

Tube ID = 0.834 inch [14 BWG]

Minimum pitch = 1.25 x OD

$$= 1.25 \times 25 = 31.25 \text{ mm} \quad [\text{Ref: M V Joshi,Pg-220}]$$

Let pitch = 32 mm square pitch

$$\text{Area} = 307.03 \text{ m}^2$$

No. of tubes (N) :

$$307.03 = \Pi \times 0.025 \times 6 \times N$$

$$N = 651.8$$

or, N = 652

Let OTL = D

$$\text{So, } (\Pi/4) \times D^2 = 652 \times (0.032)^2$$

$$D = 922 \text{ cm}$$

Now, $D_{di} = \text{OTL} + 2C$

$$= 0.922 + 2 \times 0.075$$

$$D_{di} = 1.072 \text{ m}$$

Standardizing D using Table-B4,Pg271,B C Bhattacharya

$$D_{di} = 1100 \text{ mm}$$

6.2.2 WALL THICKNESS CALCULATION [10]

Material of construction : Mild steel

Specification : IS 2002-1962 Grade-1 [Ref : B C bhattacharya,Pg-261,Appendix-A]

$$f_{all} = 0.93 \times 10^8 \text{ N/m}^2$$

$$C = 0 \text{ mm}$$

$$J = 0.85$$

$$t = p_d D_i / (2fj - p) + C \quad [\text{Ref : Equipment design- M V Joshi, Pg-96}]$$

where, t = thickness of the shell

D_i = internal diameter

J = joint efficiency

p_d = design pressure

f = permissible stress

C = Corrosion allowance

$$p_d = 1.1 \times p_s$$

$$p_s = 4.5 \text{ ata} = 4.413 \text{ bar}$$

$$p_d = 4.854 \times 10^5 \text{ N/m}^2$$

$$t = (4.854 \times 10^5 \times 1.1) / (2 \times 0.93 \times 10^8 \times 0.85 - 4.854 \times 10^5)$$

$$= 3.3876 \text{ mm}$$

Assuming, $t_{\text{std}} = 5 \text{ mm}$ [Ref: Pg-269, B C Bhattacharya]

Checking this thickness for critical buckling pressure:

P_c = critical external buckling pressure

$$P_c = [2.42E / (1 - \mu^2)^{3/4}] \times [(t/D_o)^{5/2} / \{ L/D_o - 0.45 \times (t/D_o)^{1/2} \}]$$

[Ref : M V Joshi ,eqn- 5.14, Pg-100]

$$E = 2 \times 10^6 \text{ N/cm}^2 \quad [\text{For mild steel}]$$

$$\mu = 0.30 \quad [\text{Ref: B C bhattacharya , Pg-269, table-A-8}]$$

$$L = 6 \text{ m}$$

$$D_o = D_i + 2 \times t$$

$$= 1.5 + 2 \times 0.006$$

$$= 1.512 \text{ m}$$

Putting values in the above equation of P_c we get

$$P_c = 1.316 \text{ Kg/cm}^2$$

Taking factor of safety = 4

$$P_{\text{all}} = P_c / 4 = 1.316 / 4 = 0.329 \text{ Kg/cm}^2$$

Which is less than 1 Kg/cm^2

Hence, this thickness is not acceptable.

Again, taking $t_{\text{std}} = 6 \text{ mm}$

[Ref: B C Bhattacharya, Pg-269]

We get,

$$P_c = 2.072 \text{ Kg/cm}^2$$

$$P_{\text{all}} = 2.072 / 4 = 0.518 \text{ Kg/cm}^2$$

Which is less than 1 Kg/cm^2

Hence, this thickness is also not acceptable.

Again, taking $t_{\text{std}} = 7 \text{ mm}$

[Ref: B C Bhattacharya, Pg-269]

We get,

$$P_c = 3.0387 \text{ Kg/cm}^2$$

$$P_{\text{all}} = 3.0387 / 4 = 0.75 \text{ Kg/cm}^2$$

Which is less than 1 Kg/cm^2

Hence, this thickness is also not acceptable.

Again, taking $t_{\text{std}} = 8 \text{ mm}$

[Ref: B C Bhattacharya, Pg-269]

We get,

$$P_c = 4.234 \text{ Kg/cm}^2$$

$$P_{\text{all}} = 4.234 / 4 = 1.05 \text{ Kg/cm}^2$$

Which is greater than 1 Kg/cm^2

Hence, this thickness is acceptable.

So, $t_{\text{min}} = 8 \text{ mm}$

6.2.3 SEPARATOR [10]

Top head (Elliptical head)

For 2 : 1 ellipsoidal dished head

$$D_i = 1 \text{ m} \quad [\text{Ref: Table B-4, Pg-271, B C Bhattacharya}]$$

$$L = 4 \text{ m} \quad [\text{Ref: Table B-2, Pg-269, B C Bhattacharya}]$$

$$t_h = pDV/2fJ \quad [\text{ref : Equipment design- M V Joshi, Pg-106, Eq-5.24}]$$

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

$$p = 0.23 \text{ ata} = 0.226 \times 10^5 \text{ N/m}^2$$

$$j = 0.85$$

$$D_i = 1.6 \text{ m}$$

$$k = 2$$

$$V = 1.5$$

For internal pressure :

$$\begin{aligned} t_h &= (0.226 \times 10^5 \times 1.5 \times 1) / (2 \times 0.93 \times 10^8 \times 0.85) \\ &= 2.144 \times 10^{-4} \text{ m} \\ &= 0.214 \text{ mm} \end{aligned}$$

For external pressure :

$$P_{\text{ext}} = 1 \text{ Kg/cm}^2$$

Corresponding internal pressure to be used to calculate $t_h = 1.67 \times P_{\text{ext}}$

$$\text{So, } P_{\text{int}} = 1.67 \text{ Kg/cm}^2$$

So,

$$t_h = 4.4 \times Rc [3 \times (1 - \mu^2)]^{1/2} \times (p/2E)^{1/2} \quad [\text{Ref : M V Joshi ,eqn- 5.26, Pg-107}]$$

where, p = Design external pressure

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

E = modulus of elasticity

μ = Poisson's ratio

Putting the values we get

$$t_h = 0.0061 \text{ mm}$$

So, $t_{std} = 5 \text{ mm}$ [Ref: B C Bhattacharya,Pg-269]

6.2.4 BOTTOM HEAD DESIGN [10]

Assuming an apex angle of 90°

For , conical head

$$D = 1 \text{ m}$$

$$t_h = pDV/2fJC\cos \alpha \quad [\text{Ref : M V Joshi,Pg-106}]$$

Here, α = Half the apex angle

For, internal pressure :

$$p = 0.226 \times 10^5$$

$$t_h = (0.226 \times 10^5 \times 1.5 \times 1) / (2 \times 0.93 \times 10^8 \times 0.85 \times 0.707)$$

$$= 0.3032 \text{ mm}$$

For, external pressure :

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

$$= 1.67 \times 1 \text{ kg/cm}^2$$

So,

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

$$= 2.241 \text{ mm}$$

$t_{std} = 5 \text{ mm}$ [Ref: B C Bhattacharya,Pg-269]

Checking this thickness for critical buckling pressure :

$$P_c = [2.42E / (1-\mu^2)^{3/4}] \times [(t/D_o)^{5/2} / \{ L/D_o - 0.45 \times (t/D_o)^{1/2} \}]$$

[Ref : M V Joshi ,eqn- 5.14,Pg-100]

Here, $L = D/2$

Putting the values we get,

$$P_c = 19.11 \text{ kg/cm}^2$$

So, $P_{all} = P_c/4$

$$= 19.11 / 4$$

$$= 4.778 \text{ kg/cm}^2$$

Which is greater than 1 kg/cm^2

So, this thickness is acceptable

Hence, $t_{min} = 5 \text{ mm}$

Chapter 7

Result & discussion

RESULT & DISCUSSION

The selected capacity of the plant is 4,50,000 tons/year based on 300 working days. The product from the prilling tower contains 98 % urea. Critical review of all the manufacturing processes has been presented. Snamprogetti process has been selected for the project. The Snamprogetti ammonia-stripping urea process 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.

Material & energy balance for each of the equipment has been done. The reactor is designed & its volume is found to be 195 m^3 . The length & diameter of the reactor has been found to be 40 m & 2.5 m respectively. The L/D ratio of the reactor is found to be 16 which is consistent with the actual plant data. The L/D ratio of the urea reactor according to the actual plant data lies between 14 to 20.

Climbing-film, long-tube vertical evaporator is used for the concentration of urea. The length of the heat exchanger is found to be 6 m. The diameter & height of the separator is 1 m & 4 m respectively. 652 number of tubes of 1 inch OD, 14 BWG & 6 m long (Area = 307 m^2) have been used.

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