“Urea Manufacturing Process”

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Abstract - 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 [1]. The chemical formula of, NH2CONH2, indicates that urea can be considered to be the amide of carbamic acid NH2COOH, or the diamide of carbonic acid CO(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 (P2O5) and soluble K2O. The grade of a fertilizer is expressed as a set of three numbers in the order of percent N, P2O5 and K2O. If a nutrient is missing in a fertilizer, it is represented by a zero. Thus ammonium sulphate is represented by 20.6-0-0.

INTRODUCTION

Urea is an organic compound with the chemical formula (NH2)2CO. Urea is also known by the International Nonproprietary Name (INN) carbamide, as established by the World Health Organization. Other names include carbamide resin, ISO urea, carbonyl diamide, and carbonyl diamine [1].

1.1 History

Rouelle first discovered urea in urine in 1773. His discovery is 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, Bass row produced urea by heating ammonium carbamate in a sealed tube in what is the first synthesis of urea by dehydration.

1.2 Synthetic urea

It is the first organic compound to be artificially synthesized from inorganic starting materials, in 1828 by Friedrich Wohler, who prepared it by the reaction of potassium cyanate with ammonium sulfate. This artificial urea synthesis is mainly relevant to human health because of urea cycle in human beings. Within the 20th century it is found to be the best nitrogenic fertilizer for the plants and became widely used as a fertilizer. Urea is the leading nitrogen fertilizer worldwide. Apart from that urea is being utilized in many other industries. Urea' is produced on a scale of some 100,000,000 tons per year worldwide. For use in industry, urea is produced from ammonia and carbon dioxide. Urea can be produced as prills, granules, flakes, pellets, crystals, and solutions. More than 90% of world production is destined for use as a fertilizer. Urea has the highest nitrogen content of all solid nitrogenous fertilizers in common use (46.7%). Therefore, it has the lowest transportation costs per unit of nitrogen nutrient. Urea is highly soluble in water and is, therefore, also very suitable for use in fertilizer solutions (in combination with ammonium nitrate) [1].

1.3 Commercial production of urea

Urea is commercially produced from two raw materials, ammonia, and carbon dioxide. Large quantities of carbon dioxide are produced during the manufacture of ammonia from coal or from hydrocarbons such as natural gas and petroleum-derived raw materials. This allows direct synthesis of urea from these raw materials. The production of urea from ammonia and carbon dioxide takes place in an equilibrium reaction, with incomplete conversion of the reactants. The various urea processes are characterized by the conditions under which urea formation takes place and the way in which unconverted reactants are further processed. Unconverted reactants can be used for the manufacture of other products, for example ammonium nitrate or sulfate, or they can be recycled for complete conversion to urea in a total recycle process. Two principal reactions take place in the formation of urea from ammonia and carbon dioxide. The first reaction is exothermic:

$2\ NH_3 + CO2 \longleftrightarrow H_2N\text{-}COONH_4$ (ammonium carbamate) H= -37.4 Δ Kca1/gm mol

Whereas the second reaction is endothermic:

$H_2N\text{-}COONH4 \longleftrightarrow (NH_2)_2\ CO + H_2\ O$ H= +6.3 Kcal/gm mol Δ

Both reactions combined are exothermic [1].

1.4 Chemical characteristics of urea

The urea molecule is planar and retains its full molecular point symmetry, due to conjugation of one of each nitrogen's P orbital to the carbonyl double bond. Each carbonyl oxygen atom accepts four N-H-O hydrogen bonds, a very unusual feature for such a bond type. This dense (and energetically favorable) hydrogen bond network is probably established at the cost of efficient molecular packing: The structure is quite open, the ribbons forming tunnels with square cross-section. Urea is stable under normal conditions [2].

Table 1.1:Chemica1 characteristics of urea

IUPAC name	Diaminomethanal
Chemical formula	$(NH_2)_2CO$
Molecular Mass 60.07 g/mol (approximate) Ph	(100g.L-1 in water, 20°C)~9
Dipole moment	4.56

1.5 Physical characteristics of urea

Urea is a white odourless solid. Due to extensive hydrogen bonding with water (up to six hydrogen bonds may form-two from the oxygen atom and one from each hydrogen) urea is very soluble [2].

Table 1.2: Physical Characteristics of Urea

Density	1330 kg/m^3, solid
Melting point,	132.7 °C (406 K) decompose
Boiling point	NA
Solubility in water	108 g/100 ml (20 °C)
	167 g/100 ml (40 °C)
	251 g/ 100 ml (60 °C)

	400 g/l00 m1(80 °C)
	733 g/100 ml (100 °C)
Specific heat	0.320 cal/g.^{0}C
Molecular Volume	48.16^3/kmol

1.6 Raw materials

1.6.1 Ammonia

Ammonia (NH_3) is a comparatively stable, colourless gas at ordinary temperatures, with a boiling point of -33 °C. Ammonia gas is lighter than air, with a density of approximately 0.6 times that of air at the same temperature. The characteristic pungent odors of ammonia can be detected as low as l-Sppm. Ammonia can be highly toxic to a wide range of organisms. When dissolved in water, elevated levels of ammonia are also toxic to a wide range of aquatic organisms. Ammonia is highly soluble in water, although solubility decreases rapidly with increased temperature. Ammonia reacts with water in a reversible reaction to produce ammonium $(NH_4)^+$ and hydroxide $(OH)^-$ ions, as shown in reaction. Ammonia is a weak base, and at room temperature only about 1 in 200 molecules are present in the ammonium form $(NH_4)+$. The formation of hydroxide ions in this reaction increases the pH of the water, forming an alkaline solution. If the hydroxide or ammonium ions react further with other compounds in the water, more ammonia will react to reestablish the equilibrium.

$$NH_3 + H_2O \qquad (NH_4)^+ + OH^- \longrightarrow$$

While ammonia-air mixtures are flammable when the ammonia content is 16-25% by volume, these mixtures are quite difficult to ignite. About 85% of the ammonia produced worldwide is used for nitrogen fertilizers. The remainder is used in various industrial products including fibers, animal feed, and explosives [3].

1.6.2 Carbon Dioxide

CO_2 is an odorless and colorless gas which contains 0.03% in the atmosphere. It is emitted as a pollutant from number of industries. CO_2 can be obtained from ammonia production process as a byproduct [3].

1.7 Applications of urea

1.7.1 Agricultural use

More than 90% of world production is destined for use as a fertilizer. Urea is used as a nitrogen-release fertilizer, as it hydrolyses back to ammonia and carbon dioxide, but its most common impurity, biuret, must be present at less than 2%, as it impairs plant growth. Urea has the highest nitrogen content of all solid nitrogenous fertilizers in common use (46.4%N.). Urea is highly soluble in water and is therefore also very suitable for use in fertilizer solutions, e.g. in foliar feed fertilizers.

Commercially, fertilizer urea can be purchased as prills or as a granulated material. In the past, it is usually produced by dropping liquid urea from a "prilling tower" while drying the product. The prills formed a smaller and softer substance than other materials commonly used in fertilizer blends. Today, though, considerable urea is manufactured as granules. Granules are larger, harder, and more resistant to moisture. As a result, granulated urea has become a more suitable material for fertilizer blends [3].

1.7.2 Spreading of Urea

Urea can be bulk-spread, either alone or blended with most other fertilizers. Urea often has a lower density than other fertilizers with which it is blended. Urea and fertilizers containing urea can be blended quite readily with monoammonium phosphate (11-52-0) or Diammonium phosphate (18-46-0). Urea should not be blended with superphosphates unless applied shortly after mixing. Urea will react with superphosphates, releasing water molecules and resulting in a damp material which is difficult to store and apply.

Urea fertilizer can be coated with certain materials, such as sulphur, to reduce the rate at which the nitrogen becomes available to plants. Under certain conditions these slow-release materials result in more efficient use by growing plants. Urea in a slow release form is popular for use on golf courses, parks, and other special lawn situations.

1.7.3 Industrial use

I Urea have the ability to form 'loose compounds', called clathrates, with many organic compounds. The organic compounds are held in channels formed by interpenetrating helices comprising of hydrogen-bonded urea molecules. This behaviour can be used to separate mixtures, and has been used in the production of aviation fuel and lubricating oils. As the helices are interconnected, all helices in a crystal must have the same 'handedness'. This is determined when the crystal is nucleated and can thus be forced by seeding. This property has been used to separate racemic mixtures [3].

1.7.4 Further commercial uses,

- A stabilizer in nitrocellulose explosives.
- A component of fertilizer, providing a relatively cheap source of nitrogen to promote growth. A raw material for the manufacture of plastics, to be specific, urea-formaldehyde resin.
- A raw material for the manufacture of various glues (urea-formaldehyde or urea melamine-formaldehyde); the latter is waterproof and is used for marine plywood.
- An alternative to rock salt in the de-icing of roadways and runways; it does not promote metal corrosion to the extent that salt does.
- An additive ingredient in cigarettes, designed to enhance flavour.
- A browning agent in factory-produced pretzels. An ingredient in some hair conditioners, facial cleansers, bath oils, and lotions.
- A reactant in some ready-to-use cold compresses for first-aid use, due to the endothermic reaction it creates when mixed with water.
- A cloud seeding agent, along with salts, to expedite the condensation of water in clouds, producing precipitation.

- An ingredient used in the past to separate paraffin’s, due to the ability of urea to form clathrates (also called host-guest complexes, inclusion compounds, and adducts). '

- A Flame proofing agent (commonly used in dry chemical fire extinguishers as Urea potassium bicarbonate).

- An ingredient in many tooth whitening products.

- A cream to soften the skin, especially cracked skin on the bottom of one's feet.

- Feed for hydrolyzation into ammonia which in turn is used to reduce emissions from

- Power plants and combustion engines. Other, miscellaneous products such as de-icing material for airport runways.

- Although on a smaller scale than as a fertilizer or as raw material for synthetic resins, urea is also used as a raw material or auxiliary material in the pharmaceutical industry, the fermenting and rewing industries and in the petroleum industry [3].

1.7.5 Laboratory use

Urea is a powerful protein denaturant. This property can be exploited to increase the solubility of some proteins. For this application, it is used in concentrations up to 10M. Urea is used to effectively disrupt the covalent bonds in proteins. Urea is an ingredient in the synthesis of urea nitrate. Urea nitrate is also a high explosive very similar to ammonium nitrate; however it may even be more powerful because of its complexity.

1.7.6 Global uses of urea

Urea can be added directly to feed, such as in urea-treated wheat or rice straw, or mixed with molasses (urea-molasses licks" or urea multi-nutrient blocks") for sheep, cattle, buffalo and horses. Urea may also be used as a fertilizer of the grasslands on which cattle or sheep may graze.

Another direct application of urea to land is as urea-based herbicides or pesticides (sulfonyl urea pesticides). In this case, urea is chemically synthesized with a poison or inhibitor. Sulfonyl urea is one of the preferred herbicides for broadleaf and grassy weeds. Urea may also be spread on agricultural crops to prevent frost when temperatures drop «to a level that may cause crop damage and commercial formulations of urea are available for this purpose. It is used in the growing world aquaculture industry. In intensive shrimp culture, for example, ponds may be fertilized with urea and superphosphate to initiate an algal bloom that eventually serves as food for the commercial resource. A significant proportion of such nutrients are subsequently discharged to local waters with pond effluent, as only a small fraction of added nutrients ultimately winds up in marketable product.

May also be spread on coastal oil spills, to stimulate the growth of natural bacteria populations which break down the oil. In addition to the direct applications of urea to land and sea, urea is used in many other applications, including manufacture of a wide range of common materials such as urea formaldehyde and plastics. This use represents about 50% of the non-fertilizer urea. There are numerous uses of urea in holistic medicine therapies. One application currently being considered which would greatly expand the global use of urea is as a reductant in catalytic and non catalytic reduction of combustion products in vehicles [3].

2. Body of Paper

The body of the paper consists of numbered sections that present the main findings. These sections should be organized to best present the material.

It is often important to refer back (or forward) to specific sections. Such references are made by indicating the section number, for example, "In Sec. 2 we showed…" or "Section 2.1 contained a description…." If the word Section, Reference, Equation, or Figure starts a sentence, it is spelled out. When occurring in the middle of a sentence, these words are abbreviated Sec., Ref., Eq., and Fig.

At the first occurrence of an acronym, spell it out followed by the acronym in parentheses, e.g., charge-coupled diode (CCD).

LITERATURE REVIEW

Several processes are used for urea manufacturing. Some of them used are conventional technologies and others use modern technologies to achieve high efficiency. These processes have several comparable advantages and disadvantages based on capital cost, maintenance cost, energy cost, efficiency and product quality.

Some of the widely used urea production processes are [4].

i. Conventional processes

ii. Stamicarbon CO2 stripping process

iii. Snarnprogetti Ammonia and self-stripping process

iv. Isobaric double recycle process

2.1 Conventional Processes

2.1.1 Once through Process

In this process as shown in figure 2.1 non converted ammonia is neutralized with acid such as nitric acid to produce ammonium salt such as ammonium nitrate as co products of urea production. In this way, a relatively simple urea process scheme is realized. The main disadvantages of this process are the large quantity of ammonia salt formed as co product and the limited amount of overall carbon dioxide conversion that can be achieved [4].

2.1.2 Conventional Recycle Process

Here all of the non-converted ammonia and carbon dioxide are recycled to the urea reactor. In first generation of this process the recirculation of non-converted NH_3 and CO_2 is performed in two stages. The first recirculation is operated at medium pressure (18-25 bars); the second at low pressure (2-5 bars). The first recirculation comprises at least a decomposition heater, in which carbamate decompose into gaseous NH_3 and CO_2, and while excess NH_3 evaporate simultaneously. The off gas from this first decomposition step is subjected to rectification, from which relatively pure ammonia at the top and a bottom product consisting of an aqueous ammonium carbamate solution are obtained. Both products are recycled separately to the urea reactor. In these processes, all non-converted CO_2 is recycled as associated water recycle. All conventional processes therefore typically operate at high $NH_3 2CO_2$ ratios (4-5 mol/mol) to maximize CO_2 conversion per pass. Although some of these conventional processes partly equipped with ingenious heat exchanging networks have survived until now.

Their importance decreased rapidly as the so-called stripping process is developed [4].

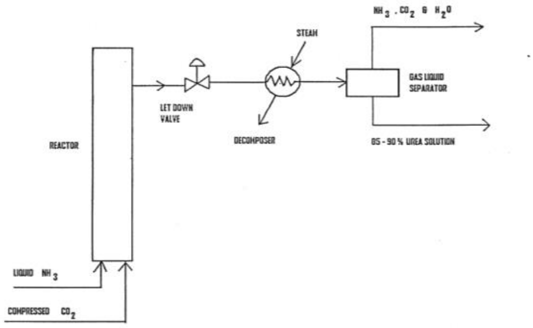


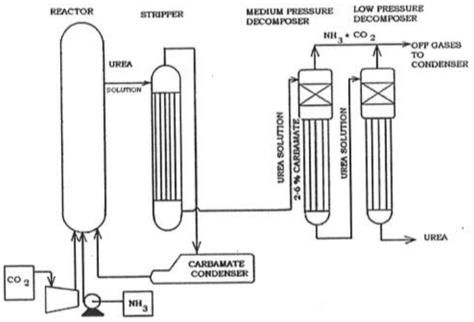
Fig. 2.1: Once through Process Flow Diagram [4]

2.2 Stamicarbon CO_2 -stripping process

In this process to achieve maximum urea yield per pass through the reactor at the stipulated optimum pressure of 140 bars, an NH_3:CO_2 molar ratio of 3:1 is applied. The greater part of the unconverted carbamate is decomposed in the stripper, where ammonia and carbon dioxide are stripped off. This stripping action is effected by counter current contact between the urea solution and fresh carbon dioxide at synthesis pressure. Low ammonia and carbon dioxide concentration in the stripped urea solution are obtained. Such that the recycle from the low pressure recirculation stage is minimized. This low concentration of both ammonia and carbon dioxide in the stripper effluent can be obtained at relatively low temperatures of the urea solution because carbon dioxide is only sparingly soluble under such conditions. Condensation of ammonia and carbon dioxide gases, leaving the stripper, occurs in the high pressure carbamate condenser as synthesis pressure. As a result, the heat liberated from ammonium carbamate formation is at a high temperature. This heat is used for the production of 4.5bar steam for use in the urea plant itself. The condensation in the high pressure carbamate condenser is not effected completely. Remaining gases are condensed in the reactor and provide the heat required for the dehydration of carbamate, as well as for heating the mixture to its equilibrium temperature. In recent improvement to this process, the condensation of off gas from the stripper is carried out in a pre-reactor, where sufficient residence time for the liquid phase is provided.

As a result of urea and water formation in condensing zone, the condensation temperature is increased, thus enabling the production of steam at higher pressure level. The feed carbon dioxide, invariably originating from an associated ammonia plant, always contains hydrogen. To avoid the formation of explosive hydrogen oxygen mixture in the tail gas of the plant, hydrogen is catalytically removed from the CO_2 feed. Apart from the air required for this purpose, additional air is supplied to the fresh CO_2 input stream. This extra potion of oxygen is needed to maintain a corrosion-resistance layer on the stainless steel in the synthesis section. Before the inert gases, mainly oxygen and nitrogen, are purged from the synthesis section, they are washed with carbamate solution from the low pressure recirculation stage in the high pressure scrubber to obtain a low ammonia concentration in the subsequently purged gas. Further washing of the off gas is performed in a low pressure absorber to obtain a purge gas that is practically ammonia free. Only one low pressure recirculation stage is required due to the low ammonia and carbon dioxide in the stripped urea solution. Because of the ideal ratio between ammonia and carbon dioxide in the recovered gases in this section, water dilution of the resultant ammonium carbamate is at a minimum despite the low pressure (about 4 bar). As a result of efficiency of the stripper, the quantities of ammonium carbamate for recycle to the synthesis section are also minimized, and no separate ammonia recycle is required. The urea solution coming from the recirculation stage contains about 75 wt. % urea. This solution is concentrated in the evaporation section. If the process is combined with a prilling tower for final product shaping, the final moisture content of urea from the evaporation section is 0.25 wt. %. If the process is combined with a granular unit, the final moisture content may wary from 1 to 5 wt. %, depending on granulation requirements. Higher moisturenent can be realized in a single stage vaporator; whereas low moisture content is economically achieved in a two stage evaporation section.

When urea with extremely low biuret content is required (at maximum of 0.3 wt. %) pure urea crystals are produced in a crystallization section. These crystals are separated from the mother liquor by combination of sieve bends and centrifuges and are melted prior to final shaping in a prilling tower or granulation unit.



The process condensate emanating from water evaporation from the evaporation or crystallization sections contains ammonia and urea. Before this process condensate is purged, urea is hydrolysed into ammonia and carbon dioxide, which are stripped off with steam and return to urea synthesis via the recirculation section. This process condensate treatment section can produce water with high purity, thus transforming this “waste water” treatment into the production unit of a valuable process condensate, suitable for, e.g., cooling tower or boiler feed water makeup. Since the introduction of the Stamicarbon CO_2 stripping process, some 125 units have been built according to this process all over the world [5].

Fig. 2.2 CO_2 stepping process flow diagram

2.3 Snamprogetti Ammonia and self-stripping process N

In the first generation of NH_3 and self-stripping processes, ammonia is used as stripping agent. Because of the extreme solubility of ammonia in the urea containing synthesis fluid, the stripper effluent contained rather large amounts 'of dissolved ammonia, causing ammonia overload in downstream section of the plant. Later versions of the process abandoned the idea of using ammonia as stripping agent; stripping is achieved only by supply of heat. Even without using ammonia as a stripping agent, the NH_3:CO_2 ratio in the stripper effluent is relatively high. So the recirculation section of the plant requires an ammonia-carbamate separation section the process uses a vertical layout in the synthesis section. Recycle within the synthesis section, from the stripper via the high pressure carbamate condenser, through the carbamate separator back to the reactor, is maintained by using an ammonia-driven liquid-liquid ejector. In the reactor, this is operated at 150 bars, NH_3:CO_2 molar feed ratio of 2:1 is applied. The stripper is of the falling film type. Since stripping is achieved thermally, relatively high temperatures (200-210 ^{0}C) are required to obtain a reasonable stripping efficiency. Because of this high temperature, stainless steel is not suitable as a construction material for the stripper from a corrosion point of view; titanium and bimetallic zirconium stainless steel tubes have been used off gas from the stripper is condensed in a kettle type boiler. At the tube side of this condenser the off gas is absorbed in recycled liquid carbamate from the medium pressure recovery section. The heat of absorption is removed through the tubes, which are cooled by the production of low pressure steam at the shell side. The steam produced is used effectively in the back end of the process. In the medium pressure decomposition and recirculation section, typically operated at 18 bar, the urea solution from the high pressure stripper is subjected to the decomposition of carbamate and evaporation of ammonia. The off gas from this medium pressure decomposer is rectified. Liquid ammonia reflux is applied to the top of this rectifier; in this way a top product consisting of pure gaseous ammonia and a bottom product of liquid ammonium carbamate are obtained. The pure ammonia off gas is condensed and recycled to the synthesis section.

To prevent solidification of ammonium carbamate in the rectifier, some water is added to the bottom section of the column to dilute the ammonium carbamate below its crystallization point. The liquid ammonium carbamate-water mixture obtained in this way is also recycled to the synthesis section. The purge gas of the ammonia condenser is treated in a scrubber prior to being purged to the atmosphere. The urea solution from the medium pressure decomposer is subjected to a second low pressure decomposition step. Here further decomposition of ammonium carbamate is achieved, so that a substantially carbamate -free aqueous urea solution is obtained. Off gas from this low pressure decomposer is condensed and recycled as an aqueous ammonium carbamate solution to the synthesis section via the medium pressure recovery section. Concentrating the urea water mixture obtained from the low pressure decomposer is performed in a single or double evaporator depending on the requirement of the finishing section. Typically, if prilling is chosen as the final shaping procedure, a two stage evaporator is required, whereas in the case of a fluidized bed granulator a single evaporation step is sufficient to achieve the required final moisture content of the urea melt. In some versions of the process, heat exchange is applied between the off gas from the medium pressure decomposer and the aqueous urea solution to the evaporation section. In this way, the consumption of low pressure steam by the process is reduced. The process condensate obtained from the evaporation section is subjected to desorption hydrolysis operation to recover the urea and ammonia contained in the process condensate [6].

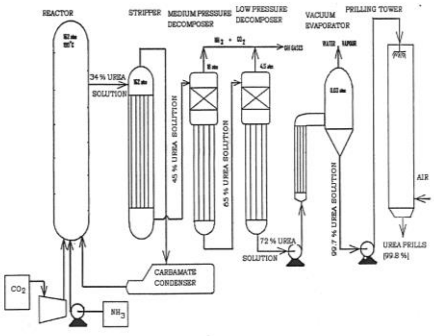


Fig 2.3: SW diagram [6]

2.4 Isobaric double recycle process

This process is developed by Mont Edison, is characterized by recycle of most of the unreacted ammonia and ammonium carbamate in two decomposer in series, both operating at the synthesis pressure. A high molar NH_3:CO_2 ratio (4:1 to 5:1) in the reactor is applied. As a result of this choice ratio, the reactor effluent contains a relatively high amount of non-converted ammonia. In the first, steam heated, high pressure decomposer, this large quantity of free ammonia is mainly removed from the urea solution. Most of the residual solution, as well as some ammonium carbamate, is removed in the second high pressure decomposer where steam heating and CO_2 stripping are applied. The high pressure synthesis section is followed by two low pressure decomposing stages of traditional design, where heat exchange between the condensing off gas of the medium pressure decomposition stage and the aqueous urea solution to the final concentration section improves the overall energy consumption of the process. Probably because of the complexity of this process, it has not achieved great popularity so far. This process or parts of the process are used in four revamps of older conventional plant [6].

PROCESS SELECTION

3.1 Selection of the Process

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. Some comparison based on various Advantages & disadvantage among the processes are given below [7].

Table 3.1: Comparison of processes

Process	Advantages	Disadvantages
Conventional Process-Once through Process.	• Simple Process	• High production cost. • High energy cost. • High environment pollution • Large quantity of ammonia salt is formed as a coproduct
Conventional Process-Conventional recycle process	• High CO_2 conversion	• High production cost • High energy cost

Stamicarbon CO_2 – Stripping Process	• High urea yield per pass	• High production cost • High energy cost • High environment pollution
Snamprogetti ammonia and self- Stripping Process	• Low environment pollution • High efficiency • High energy recovery	• High initial capital investment

Among above urea manufacturing processes, Snamprogetti process is selected because of it has following advantages compared to other processes

3.1.1 Advantages of Snamprogetti Ammonia and self-stripping process

- Less HP piping and construction materials owing to lower elevation layout, fewer HP vessels and simplified synthesis loop.
- Easier erection using commonly available construction equipment and techniques owing to low elevation layout and fewer and smaller HP vessels.
- Easier operation supported by forced circulation by HP ejector.
- Easier maintenance owing to low elevation layout and fewer HP equipment.
- Less energy consumption owing to optimized synthesis conditions and proprietarily designed reactor and stripper.
- Even though initial capital investment is higher than the other processes, it will overcome by lower production cost per metric ton of urea [7].

3.2 Manufacturing Process

In the reactor, when liquid ammonia reacts with compressed CO_2 (at 162 atm) at high temperature & pressure gives urea according to the following reactions:

$$2NH_3 + CO_2 \longrightarrow NH_2COONH_4$$

exothermic

$$NH_2COONH_4 \longrightarrow NH_2CONH_2 + H_2O$$

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 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 [8].

Urea Synthesis

NH_3& CO_2 react under specific concentration, temperature & pressure 00nditions to form Urea as per the following reactions:

1) $CO_2\ (g) + 2NH_3\ (g) \longrightarrow NH_2COONH_4(S)$

$H = -37.64$ kcal/gm mol

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

Overall Reaction:

$CO_2\ (g) + 2NH_3\ (g) \longrightarrow NH_2CONH_2\ (s) + H20$
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.

3.3 Effect of Various Parameters Temperature

In above reactions 1St reaction is exothermic & 2"done is endothermic. So, according to Le chatelier's Principle 1St reaction is favoured at low temp & 2"done at high temp. Further, reaction no. 1 is fast & teaches 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.

$$NH_2COONH_4\ (s) \longrightarrow 2NH_3\ (g) + CO_2\ (g)$$
$$H = -ve\ (K1)$$

$$NH_2COONH_4\ (s) \longrightarrow NH_2CONH_2\ (s) + H_2O$$
$$H = we\ (K2)$$

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:

$$CO_2\ (g) + 2NH_3\ (g) \qquad NH_2CONH_2\ (s) \longrightarrow + H_2O$$

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 slowing, 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 [9].

3.4 Snamprogetti Stripping Process

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.

$NH_2COONH_4\ (s) \longrightarrow 2NH_3\ (g) + CO_2\ (g)$
H = + 37-4 Kcal/gm-mole This reaction involves increase in volume & absorption of heat. Thus this reaction will be favoured by decrease in pressure & increase in temp. Moreover decreasing the partial pressure of either of the products will also favour 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:

$K = (P_{NH3})^2 * (P_{CO2})$ [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 NH3 is present in the reactor as in Snamprogetti process. CO_2 stripping is advantageous because introducing CO_2 increase P_{CO2}. So P_{NH3} will be reduced to maintain P constant as $P = P_{CO2} + P_{NH3}$. At a particular temp K is constant so when P_{NH3} is reduced to keep K constant, carbamate will be reduced much faster by decomposition as P_{NH3} 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 favour of isobaric stripping is those higher carbamates recycle 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 [9].

3.4.1 Process Description

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

i. Urea synthesis & high pressure recovery.
ii. Urea purification & low pressure recovery.
iii. Urea concentration.
iv. Urea prilling.

3.4.1.1 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 23 atm pressures 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 240 atm. Before entering the reactor it is' used as a driving fluid in the carbamate ejector, where the Carbamate coming from carbamate separator is compressed up to 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 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 NH_3 / CO_2 is around 2:1; the mole ratio of water to CO_2 is around 0.67: 1.

$$2NH_3\,(g) + CO_2\,(g) \rightarrow NH_2COONH_4\,(s) \text{ exothermic}$$

$$NH_2COONH_4\,(s) \rightarrow NH_2CONH_2\,(s) + H_2O \text{ endothermic}$$

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 no's 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 24 atm steam. The overhead gases from shipper 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 inert 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 atm 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 [9].

3.4.1.2 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., MP decomposer
2nd stage at 4.5 atm pressure, i.e., LP decomposer

1st stage purification & recovery stage at 18 atm:

It is falling film type 1va 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 atm steam condensate flowing out of the stripper.

2nd purification & recovery stage at 4.5 atm:

The solution leaving the bottom of MP decomposer is expanded at 4.5 atm 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 atm.

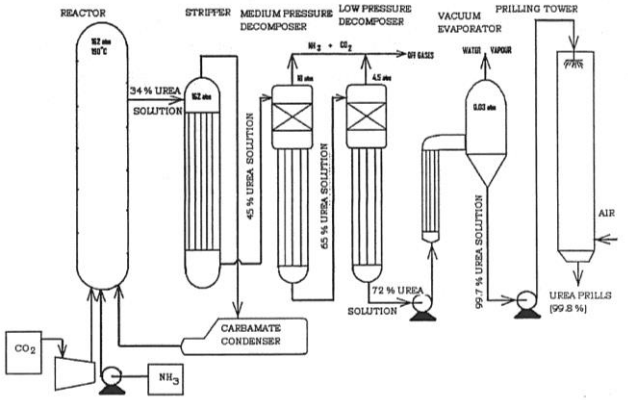


Fig. 3.1: Snarnprogetti urea process flow diagram [9].

3.4.1.3 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 atm .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 0.03 atm . The two concentrators are fed by saturated steam at 4.5 atm. The mixed phase coming out enters the gas liquid separator, wherefrom the vapors are extracted by the 2nd vacuum system.

3.4.1.4 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.

MATERIAL AND ENERGY BALANCE

4.1 Material Balance

Selected capacity : 100 tons/day

Urea : 4083.33 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:

1) $CO_2 + 2NH_3 \longrightarrow NH_2COONH_4$

(44) (17) (78)

2) $NH_2COONH_4 \longrightarrow NH_2CONH_2 + H_2O$

(60) (18)

3) $CO_2 + 2NH_3 \longrightarrow NH_2CONH_2 + H_2O$ (Overall reaction)

Side reaction:

4) 2NH2CONH2 ⟶ NH2CONHCONH2 + NH3 (103)

4.1.1 Reactor

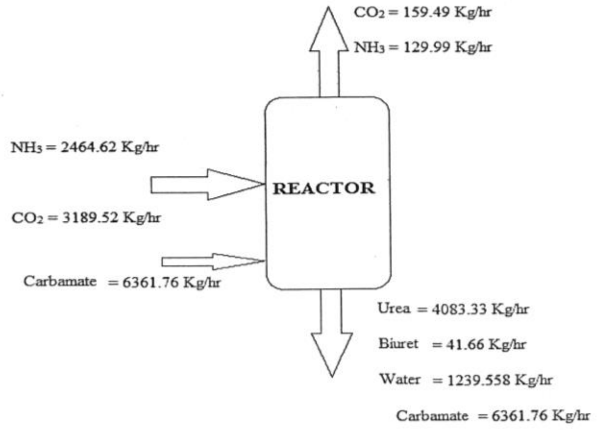


Fig. 4.1: Flow of material across reactor

From reaction 4

Concentration of biuret in product = 1 %

i.e. biuret = 41.66 kg/hr.

Biuret produce from Urea = 120/103 *41.66

= 48.536 kg/hr. Urea produced by reaction 2

= 4083.33+48.536

= 4131.866 kg/hr.

NH3 required from reaction 3

= 34/60*4131.87

= 2341.39 kg/hr.

& CO2 required = 44/60*4131.87

= 3030.04 kg/hr.

Our conversion is 95%

NH3 actually fed = 2464.62 kg/hr

CO2 actually fed = 3189.52 kg/hr.

If the reaction 3 is 100% complete

Urea produced = 60/44*3189.52

= 4349.35 kg/hr.

But actual conversion is 95% Urea produced

= 0.95*4349.35

= 4131.86 kg/hr.

Urea converted to biuret & NH_3

= 4131.86-4083.33

= 48.53 kg/hr.

From reaction 4

Biuret produced = 103/ 120*48.53

= 41.66 kg/hr.

H20 produced in reaction 2

= 18/60*4131.86

= 1239.56 kg/hr.

At reactor exit (urea = 34%)

Exit flow rate = 4083.33/0.34

= 12009.79 kg/hr.

Actual urea produced = 4083.33-48.53

= 4083.33 kg/hr.

NH3:

From reaction 3, NH3 reacted is

= (34/60)*4131.86

= 2341.39 kg/hr.

From reaction 4 NH3 produced

= 17/120*48.53

= 6.88 kg/hr.

NH3 unreacted = 2464.6 2341.49 + 6.88

=129.99 kg/hr.

CO_2:

From reaction 3 CO_2 reacted = 44/60*4l31.86

= 3030.03 kg/hr.

CO_2 unreacted = 31892-303003

= 159.49 kg/hr.

Carbamate flowrate = exit flow rate-[urea + biuret + water + NH_3 + CO_2]

= 6361.76 kg/hr.

4.1.2 Stripper

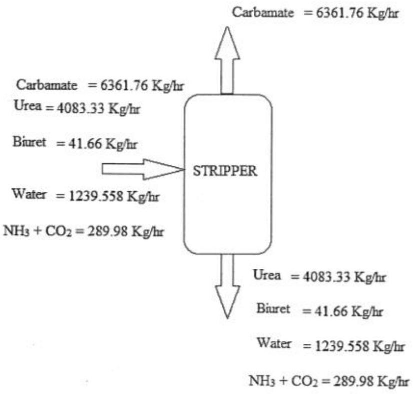


Fig. 4.2: Flow of material across stripper

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 is in the inlet stream.

4.1.3 Medium Pressure Separator

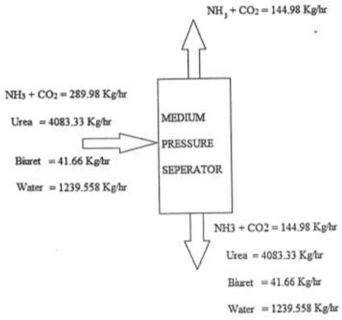


Fig. 4.3: Flow of material across medium pressure separator

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

x = 0.7221,

y = 0.7411

Fx = Py

F = (Py)/ x

= (5509.528 x 0.7411)/0.7221

= 5659.458 Kg/hr.

NH_3 + CO_2 Removed

= 5654.458-5509.528

= 144.98 Kg/hr.

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.

4.1.4 Low Pressure Separator

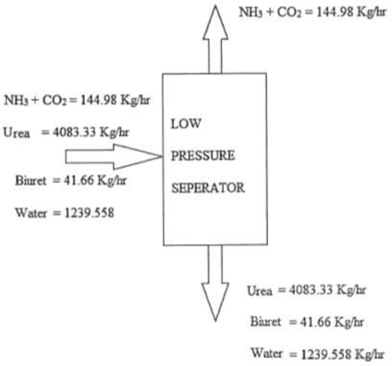


Fig. 4.4: Flow of material across low pressure separator

y=0.7611 x=0.7411

P = 5364.55 kg/hr

F = 5364.55*0.7611/0.7411

F = 5509.53 kg/hr

(NH_3+CO_2) separated from top = 144.98 kg/hr

4.1.5 Evaporator

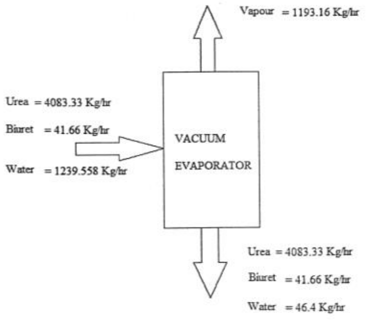


Fig. 4.5: Flow of material across evaporator

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

x = 0.7611

y = 0.9788

Taking urea balance:

F.x = P.y

F*O.7611 = 4171.39*0.9788

F = 5364.55 Kg/hr.

Overall material balance gives:

F = P + E

5364.55 = 4171.39 + E

E = 1193.16 Kg/hr.

4.1.6 Prilling Tower

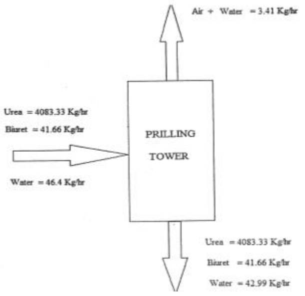


Fig. 4.6: Flow of material across prilling tower

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

x = 0.9788 y = 0.9796

Making urea balance:

F.x = P.y

F*O.9788 = 4167.98*0.979S

F = 4171.39Kg/hr.

Water removed = 4171.39 4167.98 = 3.41kg/hr.

4.2 Energy Balance

Assumption: Datum temperature = 0°C

4.2.1 Reactor

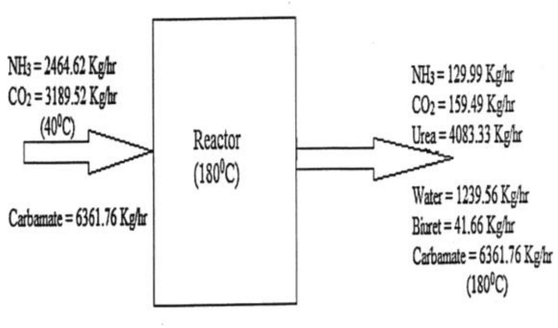


Fig. 4.7: Energy 110w across reactor

Inlet Stream

Material	**specific heat at 40°C**
NH_3	0.53 cal/gm°C = 2.219 KJ/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/Kg °C

Heat input

Material	mCp $\underline{\Delta t}$
NH_3	2464.62x 2.219 x 40 = 218759.67 KJ/hr.
CO_2	3189.52x 0.9211 x 40=117514.67 KJ/hr.
Carbamate	6361.76 X 2.596 x 180= 2972725 .213KJ/hr.

Heat input = 3.309x 10^6 KJ/hr.

ΔH_R = -31.32 Kcal/gm mol

= -0.013 x 10^7 KJ/Kmol of Urea formed.

Amount of urea formed during the reaction = 68.056 Kmol/hr.

ΔH_R = 68.056 x 0.013 x 10^7 KJ/hr.

= 8.85 x 10^6 KJ/hr.

Outlet Stream

Table 4.1: Outlet Stream Composition from reactor

Material	Specific heat at 180°C	Molfractions (x)	Flow rate (Kmol/hr.)
NH_3	0.55 cal/gm°C = 39.15 KJ/Kmol°C	0.0334	7.646
CO_2	0.23 cal/gm°C = 42.37 KJ/ Kmol°C	0.01586	3.6248
Carbamate	0.62 cal/ gm°C = 202.49 KJ/ Kmol°C	0.354	80.279
Urea	0.4828 cal/gm°C=121.32 KJ/Kmol°C	0.296	68.056
Water	1 cal/gm°C = 75.37 KJ/Kmol°C	0.299	68.364
Biuret	183.8 KJ/Kmol^{0}C	0.002	0.4045
TOTAL			**228.874**

Cpof mixture = $\Sigma x_i Cp_i$

SO, CD = 0.0334 x 39.15 + 0.01586 x 42.37 + 0.296 x 121.32 + 0.354 x 202.49 +
0.002 x 183.8
+ 0.299 x 75.37 5 132.46 KJ/Kmol °C

So, heat carried by outlet stream = mCpΔt
= 228.874 x 132.46 x 180
= 5.4569 x 10^6 KJ /hr.

Heat input + AHR Heat output = rate of accumulation

3.309x 10^6+8.85 x 10^6-54569 x 10^6 = rate of accumulation

Rate of accumulation = 6.697x 10^6 KJ/hr.

Brine at -17°C is used to remove heat from the reactor. The outlet is stream at temperature 35°C.

So, heat gained by brine = 6.697 x 10^6 KJ/hr.

mCpΔt = 6.697 x 10^6 KJ/hr.

m x 3.2408 x (-17-35) = 6.697 x 10^6 KJ/hr.

m = 39095.01 Kg/hr.

4.2.2 Stripper

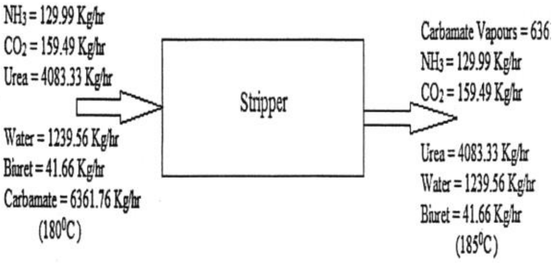


Fig. 4.8: Energy flow across stripper

Total heat input = 5.4569 x 10^6 KJ/hr.

Outlet Stream

Table 4.2: Outlet stream composition of liquid from stripper

Material	Specific heat at 185°C	Molfractions (x)	Flow rate (Kmol/hr.)
NH_3	0.58 cal/gm°C = 41.31KJ/Kmol°C	0.0514	7.696
CO_2	0.24 cal/gm°C = 44.22 KJ/ Kmol°C	0.24	3.6248
Urea	0.5385 cal/gm°C=135.3 KJ/Kmol°C	0.46	68.008
Water	1 cal/gm°C = 75.37 KJ/Kmol°C	0.463	68.864
Biuret	183.8 KJ/Kmol°C	0.003	0.4043
TOTAL			148.6

Cp of mixture = $\Sigma x_i Cp_i$

So, Cp = 0.05 x 41.31 + 0.024 x 44.22 + 0.46 x 135.3 + 0.003 x 183.8 + 0.463 x 75.37

= 100.81 KJ/ Kmol°C

So, heat carried by outlet stream = mCpΔt

= 148.6 x 100.81 x 185 = 2771367.7 KJ/hr.

Table 4.3: Outlet stream composition of Vapour (Ammonium carbamate) from stripper

Material	Specific heat at 185°C	Flow rate (Kmol/hr.)

Carbonate	0.62 cal/g In °C = 202.49 KJ/Km01°C	81.56

For carbamate λ = 210 KJ/Kg

So, heat carried by carbamate = $mCp\Delta t + m\lambda$

= 81.56 x 202.49 x 185 + 6361.76 x 210

= 4391260.214 KJ/hr.

Here, steam at 24 atm is used (Ts = 222°C).

1 of steam =1855.3 KJ/kg

Heat supplied by steam = Heat output Heat input

= (4391260 .214 + 2771367.7) 5.4569 x 10^6 KJ/hr.

m = 170572. 914 Kg/ hr.

4.2.3 Carbamate Condenser

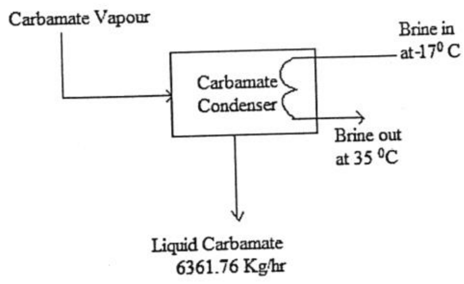


Fig.4.9: Energy flow across carbamate condenser

$M_v\lambda_v = mCp\Delta t$

Heat gained by coolant = $6.69*10^6$KJ/hr.

Putting the values we get,

$6.69*10^6$= m. x 3.2098 x 52

So, m_s= 39095.01 kg/h

4.2.4 Medium Pressure Separator

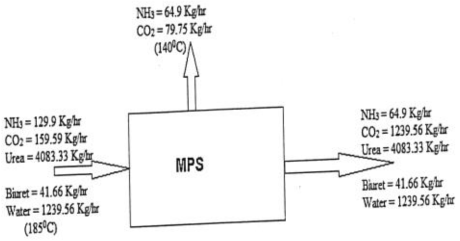


Fig. 4.10: Energy flow across medium pressure separator

Heat input = 2771367.7KJ/hr.

Outlet Stream

Table 4.4: Outlet Stream Composition of Liquid from Medium Pressure Separator

Material	Specific heat at 140°C	Molfractions (x)	Flow rate (Kmol/hr.)
NH_3	0.54 cal/gm°C = 38.4 KJ/Kmol°C	0.027	3.82
CO_2	0.23 cal/gm°C = 42.37 CO_2 KJ/Kmol°C	0.0127	1.813
Urea	0.493 cal/gm°C=123.84 KJ/Kmol°C	0.476	68.008
Water	1 cal/gm°C = 75.37 KJ/Kmol°C	0.4815	68.884
Biuret	170.92 KJ/Kmol°C	0.0028	0.4046
TOTAL			142.96

Cp of mixture = $\Sigma x_i Cp_i$

So, Cp = 0.027 x 38.4 + 0.0127 x 42.37 + 0.476 x 123.84 + 0.0028 x 170.92 + 0.4815 x 75.37

= 97.29 KJ/Kmol°C

Heat output = 142.96 x 97.29 x 140 KJ/hr.

= 1.947x 10^6KJ/hr.

λ of mixture = $\boldsymbol{\Sigma x_i \lambda_i}$

So, λ. = (0.6785 x 22.777 + 0.321 x 20.265) x 10^3

= 21.969 x 10^3 KJ/Kmol°C

Table 4.5: Outlet stream composition of Vapor escaping from top of Medium

Material	λ at 140°C	Molfractions (x)	Flow rate (Kmol/hr.)
NH_3	0.54 cal/gm°C = 38.4 KJ/Kmol°C	0.027	3.82
CO_2	0.23 cal/gm°C = 42.37 CO_2 KJ/Kmol°C	0.0127	1.813
TOTAL			5.65

Cp of mixture = $\Sigma x_i Cp_i$

So, Cp = 0.678 x 38.4 + 0.3215 x 42.37 KJ/Kmol°C= 39.676 KJ/Kmol°C

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

= 5.633 (39.676x 140 + 21.969 x 10^3)

= 155040.6 KJ/hr.

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

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

= 2771367.7$^-$ 1.947 x 10^6 KJ/hr.

mCp∆t = 669327.1KJ/hr.

m = 669327.1/ (4.187 x 25)

m = 6394.34 Kg/hr.

4.2.5 Low Pressure Separator

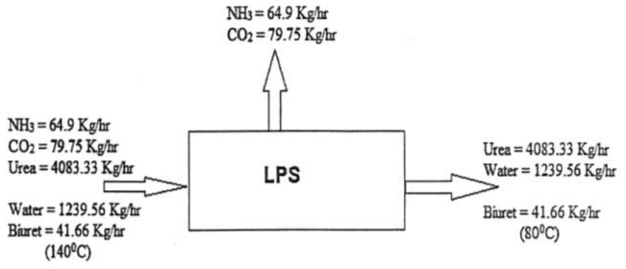


Fig. 4.11: Energy flow across low pressure separator

Heat input = 1.947x 10^6 KJ/hr.

Outlet Stream

Table 4.6: Outlet stream composition of Liquid from Low Pressure Separator

Material	Specific heat at 80°C	Molfractions (x)	Flow rate (Kmol/hr.)
Urea	0.429 cal/gm°C=107.76 KJ/Kmol°C	0.496	68.008
Water	1 cal/gm°C = 75.37 KJ/Kmol°C	0.5	68.884
Biuret	149 KJ/Kmol°C	0.004	0.4046
TOTAL			137.32

Cp of mixture = $\Sigma x_i Cp_i$

So, Cp = 0.496 x 107.76 + 0.5 x 75.37 + 0.004 x 149 KJ/Kmol°C

= 91.73KJ/Kmol°C

Heat output = 137.52 x91.73 x 80

= 1.009 x 10^6KJ/hr.

Table 4.7: Outlet stream composition of Vapor escaping from top of Low Pressure

Material	λ at 140°C	Molfractions (x)	Flow rate (Kmol/hr.)
NH_3	260 cal/gm°C = 18.51 x 10^3 KJ/Kmol°C	0.679	3.82
CO_2	85 cal/gm°C = 15.66 x 10^3 CO_2 KJ/Kmol°C	0.321	1.813
TOTAL			5.633

λ of mixture= $\Sigma x_i \lambda_i$

So, λ= (0.679 x 18.51 + 0.321 x 15.66) x 103 KJ/Kmol°C

= 17.6 x 10^3 KJ/Kmol

Cp of mixture = $\Sigma x_i Cp_i$

So, Cp = 0.679 x 37.013 + 0.321 x 38.69 KJ/Kmol°C

= 37.55 KJ/Kmol°C

Heat escaping from the top = m(Cp∆t +96) = 5.633(37.55 x 80 + 17.7 x 10^3)

= 1.66 x 10^5 KJ/hr.

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

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

= 1.947x10^6 - 1.009 x 10^6 KJ/hr.

mCp∆t = 772000KJ/hr.

m =772000/ (4.187 x 25)

m = 7375.21 Kg/hr.

4.2.6 Evaporator

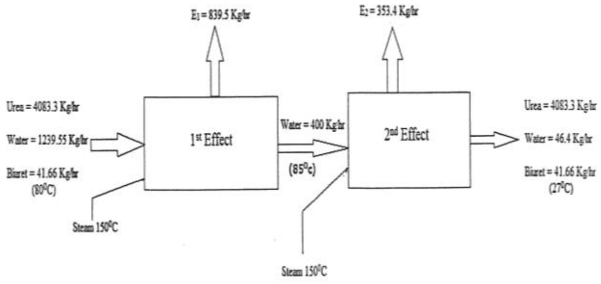


Fig. 4.12: Energy flow across evaporator

Table 4.8: Product stream coming out of 1St evaporator

Material	Specific heat at 85°C	Molfractions (x)	Flow rate (Kmol/hr.)
Urea	0.435 cal/gm°C=109.28 KJ/Kmol°C	0.75	68.056
Water	1 cal/gm°C = 75.37 KJ/Kmol°C	0.245	22.226
Biuret	149 KJ/Kmol°C	0.005	0.4045
TOTAL			90.6565

Cp of mixture = $\Sigma x_i Cp_i$

So, Cp = 0.75 x 109.28 + 0.245 x 75.37 + 0.005 x 149 KJ/Kmol °C

= 101.17 KJ/Kmol°C

$mCp\Delta t$ = 90.6565 x 101.17 x 85

= 779768.028 KJ/hr.

Heat Balance.

1st evaporator:

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

Heat input (feed) + $S_1\lambda_{S1}$= E_1H_{E1} + energy of the bottom product 1.009 x10^6 + 81 x 2123.2 = 839.55 x 2614.97 + 7.79 x10^5

S1 = 926.08 Kg/hr.

2nd evaporator:

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

Heat input (feed) + $S_2\lambda_{S2}$ = E_2H_{E2} + energy of the bottom product

7.79 x 10^5+ S_2x 2123.2 = 353.4 x 2545.7 + 10.91 x 10^6

S_2 = 144.307 Kg/hr.

4.2.7 Prilling Tower

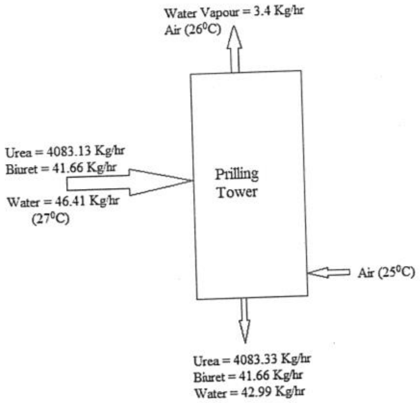


Fig. 4.13: Energy balance across prilling tower

Heat input= 10.91x 10^7 KJ/hr.

Outlet stream

Table 4.9: Outlet Stream Composition from prilling tower

Material	Specific heat at 30°C	Molfractions (x)	Flow rate (Kmol/hr.)
Urea	0.3758 cal/gm°C=94.4l KJ/Kmol°C	0.96	68.056
Water	1 cal/gm°C = 75.37 KJ/Kmol°C	0.34	2.38
Biuret	133.02 KJ/Kmol°C	0.005	0.4045
TOTAL			70.85

Cp of mixture = $\Sigma x_i Cp_i$

So, Cp = 0.96 x 94.41 + 0.034 x 75.37 + 0.006 x 133.02 KJ/Kmol°C

= 93.99 KJ/Kmol°C

Heat output = 70.85 x 93.99 x 25 = 166479.79 KJ/hr.

Assuming, humidity of air at 25°C = 0.01

Heat carried away by air = heat input heat output

$(mCp\Delta t)$ dry air = (1091-010) x 10^6

= 10.14 x106 KJ/hr.

m =10.14 x 10^6/ (1.009 x 1)

m = 10.053 x 10^6 Kg/hr.

So, flow rate of air =10.053 x 10^6 Kg/hr.

EQUIPMENT DESIGN

5.1 Reactor Design

Type of Reactor Plug Flow Reactor

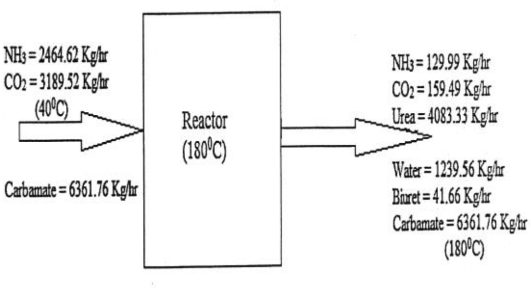


Fig. 5.1 Urea reactor

Values of activation energy and reaction rate constant are taken from the literature.

E = 139500 J/mole

K= 2.5[14].

We Know,

V/Vo = (1/K) x [-In (1-X_A)]

V = (12.417/2.5) x [-In 0-095)]

V = 14.88

t=V/F

= 14.88/12.417

t= 1.198 hr.

=71.88 min.

Taking, L/D = 6

V = (π/4) x D^2.L

14.88 = (π/4) x D^2 x 6D

D = 1.467

D = 1.5 m.

L=9m.

5.1.1 Thickness of Shell Data available:

Pressure inside the reactor = 154 atm.

Material of construction:

Low alloy carbon steel [IS-2002-1975]

Allowable stress = 121x 108 N/mm^2

Diameter of the reactor = 1.5 m t

= P-Di/ (213 -p) [16]

Where,

t = thickness of the shell

D_i = Internal diameter

J = joint efficiency

p = design pressure

f = permissible stress

Internal pressure=154 atm.

Design pressure= p = (10 % extra)

= 1.1 x 154

= 169.4 atm.

= 1.716 x107 N/m^2

= 17.16 N/mm^2

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

f= 121 N/mm^2

D_i = 1.5 m

So, t= (17.16 x 1.5)/ (2 x121-17.16)

t = 0.01145 m

Or, t = 1.2 cm.

5.1.2 Head Design

For 2:1 ellipsoidal dished head

t_h = pDV/2fJ

Where, p = interna1 design pressure

D = major axis of ellipse

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

k = major axis/minor axis So, th= (17.16 x 1.5 x1)/(2x121x1)

t_h= 0.0106m.

t_h = 1.2cm

5.1.3 Skirt Support for Reactor

Wt. of the reactor = wt. of material of construction + weight of the contents of the reactor

Wt. of material of construction = π.D.t.L. p

= π x 1.5 x 0.12x 9x7850

=39951.634 Kg

Weight of the reactor = 39951.634+12015.88

W = 51967.514 Kg

Wind Pressure = P1= 1226.25 N/m^2

Material of construction:

IS: 2002-1962 Grade 2B

Allowable tensile stress = 1.18 x 10^8 N/m^2

Yield stress = 2.55 x 10^8 N/m^2

Stress due to dead weight:

f_d = ΣW/ (π. $D_{ok.}$ t_{sk})

Where,

f_d = Stress due to dead weight

ΣW = Dead wt. of vessel

D_{ok} = Outside diameter of the skirt

t_{sk} = thickness of skirt

f_d= 51967.514/ (3.14 x1.5 x tsk)

= 11027.85 /t_{sk} N/m^2

Assuming height of skirt = 5 m

f_{wb} = M_w/Z

= 4M_w/ (π.D^2 ok t_{sk})

M_w = P_{1w} (h_1/2)

P_{1w}= kP_l h_1Do

= 0.7 x 1226.25 x 14 x 1.62

= 19467.945

Mw=19467.945/ (14/2)

=136275.615

F_{wb}= (4 x 136275.615)/ (π.1.5^2. t_{sk})

=66] 14.73 / t_{sk}

Stress due to seismic load:

f_{sb}= (2/3) x CWH/ (π R_{ok} t_{sk})

[Here C = 0.08]

f_{sb} = (2/3) x (0.08 x 51967 x 9)/((3.14 x0.81 x t_{sk})

f_{sb} = 9802.532 /tsk kg/m^2

Maximum tensile stress = f_{wb} - f_d

= 66114.73/t_{sk} 10210.97/t_{sk}

(ft) Max = 55903.76/t_{sk}

Now, permissible tensile stress = 1.18 x 108N/m^2=12028542.3 kg/ m^2

Now, permissible tensile stress = Maximum tensile stress

tsk = 55903.76/ 12028542.3

= 4.6476x10^3 m

= 4.64 mm = 6m

Maximum compressive stress:

$(fc)_{Max}$ = 661 14.73/tsk + 10210.91/tsk

$(fc)_{Max}$ = 76325.7/tsk

$(fc)_{max}$ =12572205.25 kg/ m^2

Or, t_{sk} = 76325.7/ 12572205.25

t_{sk} = 6.0704 x 10^{-3} m

= 6.1 mm

So, thickness to be used = 8 mm

5.2 Multiple Effect Evaporator Design

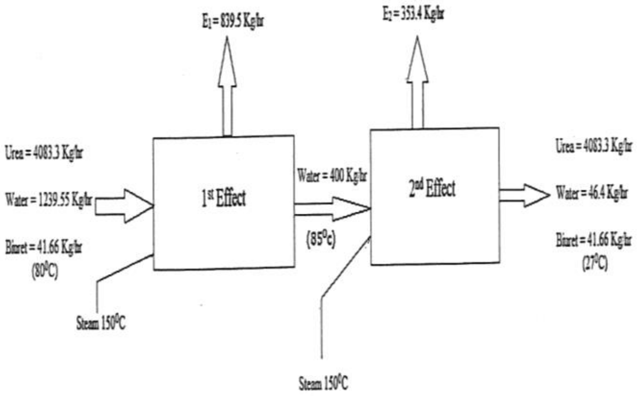


Fig. 5.2: Evaporator

Vapor space pressure = 0.23 atm

Vapor space temperature = 63.1 °C

BPR = 28.11°C (1)

Boiling point of liquid = 85 °C

Table 5.1: product stream coming out of 1st evaporator:

Material	Specific heat at 85°C	Molfractions (x)	Flow rate (Kmol/hr.)
Urea	0.435 cal/gm°C=109.28 KJ/Kmol°C	0.75	68.056
Water	1 cal/gm°C = 75.37 KJ/Kmol°C	0.246	22.226
Biuret	149 KJ/Kmol°C	0.005	0.4045
TOTAL			90.6865

Cp of mixture = $\Sigma x_i Cp_i$

So, Cp = 0.75 x 109.28 + 0.245 x 75.37 + 0.005 x 149 KJ/Kmol°C

= 101.17 KJ/Kmol°C

$mCp\Delta t$= 90.6865 x 101.17 x 85

= 7.797 x 10^5 KJ/hr.

Heat balance

1^{st} evaporator:

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

Heat input (feed) + S1151 = EIHEI + energy of the bottom product

For steam at 150 °C, 151 = 2123.2 KJ/kg

Putting the values we get

1.009 x106 + 31 x 2123.2= 839.55 x 2614.97 + 7.79 x10^5

S_1 = 926.08 Kg/hr.

Economy = 839.55/926.08 = 0.9066

Now,

U_1 value is obtained from fig: 5.3.At 63.1°C (145.58°F) the value of U_1 is 270 Btu/hr.ft^2.°F. Multiplying this value by 5.6783 gives the value of U_1 in W/m^2K

$A_1= S_1\lambda_{S1}/U_1\Delta T_1$

ΔT_1 = (AT) app BPR_1

= 150 63.1 -28.11

= 58.79 °C

So, A_1= 926.08 x 2123.2/ 1533 x 58.79

A_1= 21.8149 m^2

Similarly,

$A_2 = S_2\lambda_{S2}/U_2\Delta T_2$

ΔT_2 = (AT) app BPR_2

= 150 23.7787.097 = 38.903 °C

So, A_2= 144.307 x 2123.2 / 738 x 38.903

A_2= 10.6718 m^2

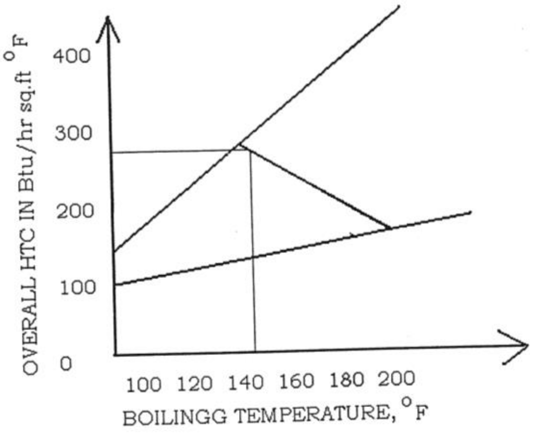


Fig. 5.3 Graph to find out heat transfer co-efficient [18].

5.2.1 For I^{st} Effect Evaporator.

Assuming:

Length = 6 m

Tube OD = 1 inch = 2.54 cm

Tube ID = 0.834 inch = 2.11836 cm

Minimum pitch = 1.25 x OD

= 1.25 x 25 = 31.25 mm

Let,

Pitch = 32 mm^2

Area=21.82m^2

No. of tubes (N):

21.82 = π x 0.025 x 6 x N

N = 46.293

Or

N=48

Let, OTL = D

So, (π/4) x D2 = 48 x $(0.032)^2$

D = 0.25016 m

Now,

D_{di}= OTL + 2C

= 0.25016 + 2 x 0.075

D_{di} = 0.40016 m

D_{di}= 400 mm

5.2.1.1 Wall Thickness Calculation

Material of construction: Mild Steel

Specification: IS 2002-1962 Grade-l

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

c = 0 m

J = 0.85

$t = P_d D_i / (2fj - p) + C$

Where,

t = thickness of the shell

D = internal diameter

J = joint efficiency

p_d= design pressure

f = permissible stress

C=Corrosion allowance

$P_d = 1.1 \times P_s$

P_s =4.5atm=4.413bar

$P_d = 4.854 \times 10^5$N/m^2

$t = (4.854 \times 105 \times 0.4)/ (2 \times 0.93 \times 10^8 \times 0.85\ 4.854 \times 10^5)$

= 1.23 mm

Assuming, t_{std} = 3.2 mm

Checking this thickness for critical buckling pressure:

Pc = critical external buckling pressure

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

$E = 2.2 \times 10^6$ N/cm^2

u = 0.30

L=6m

$D0 = Di + 2 \times t$

$= 0.4 + 2 \times 0.032$

= 0.4064 m

Putting values in the above equation of Pc we get

Pc = 2.13557 Kg/ cm^2

Taking factor of safety (FOS) = 4

Pall = Pc/4 = 2.13557/4

= 0.533 Kg/ cm^2

This is less than 1 Kg/ cm^2

Hence, this thickness is not acceptable.

Again, taking t_{std} = 5 mm

We get, Pc = 6.4 Kg/ cm^2

Pall = 6.4 /4 = 1.609 Kg/ cm^2

This is less than 1Kg/ cm^2

This is greater than 1Kg/cm^2

Hence, this thickness is acceptable.

So, **t_{min} = 5 mm**

t_{min} = 6 mm

5.2.1.2 Separator

Top head (Elliptical head)

For 2:1 ellipsoidal dished head

Di = 1 m

L = 4m

$t_h = pDV/2fJ$

Where,

p = internal design pressure

D = major axis of ellipse

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

k = major axis/minor axis

$p = 0.23$ atm $= 0.226 \times 10^5$ N/m^2

j=0.85, Di= 1.6m, k=2, v=1.5

For internal pressure:

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

$= 2.144 \times 10^{-4}$ m

= 0.214 mm

For external pressure:

P_{ext} = 1 Kg/cm^2

Corresponding internal pressure to be used to calculate

$t_h = 1.67 \times P_{ext}$

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

So,

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

Where,

p = Design external pressure

Rc = Crown radius for torispherica1& hemispherical heads and equivalent crown radius for elliptical head. (R_c = Do)

E = modulus of elasticity

p = Poisson's ratio

Putting the values we get

t_h = 1.84 m

So, **t_{std}= 2 mm**

5.2.1.3 Bottom Head Design

Assuming an apex angle of 60°

For, conical head

D=1m

t_h = pDV/2fJCos 01

Here, a = Half the apex angle

For, internal pressure:

p = 0.226 x 10^5

t_h = (0.226x 10^5 x 1.5 x 1) / (2 X 0.93 x10^8 x 0.85x 0.5)

= 0.429 mm

For, external pressure: ;

p = 1.67 x P_{ext}

= 1.67 x 1 kg/cm^2

So, t_h = (1.67x10^5 x 1.5 x 1) / (2 x 0.93 x10^8 x 0.85 x 0.5)

= 3.17 mm

t_{std} = 3.2 mm

Checking this thickness for critical buckling pressure:

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

Here,

L = D/2

Putting the values we get,

Pc = 25.92 kg/cm^2

So, Pall = Pc/4 = 25.92 / 4

= 6.4786 kg/ cm^2

Which is greater than 1 kg/cm^2

So, this thickness is acceptable

Hence, t_{min} = 3.2 mm

5.2.2 For 2nd Effect Evaporator

A = 10.6718 m^2

No. of tubes,

A= μ x OD x N x L

10.6718 = μ x 0.025 x N x 6

N = 22.64

No. of tubes = 24

(n/4) x D^2 = N x (Pitch)^2

D^2 = [24 x $(0.032)^2$ x 4] / μ

D = 0.1718 m.

D_{di} = D + (2 x C)

= 0.1718 + 2 x 0.075

= 0.3218 m

D_{di} = 325 mm

5.2.2.1 Wall Thickness Calculation

Material of construction:

Mild steel Specification: IS 2002-1962 Grade-l

f_{all}= 0.93 x 108 N/m^2

C= 0 mm

J=0.85

t= P_d Di/(2fj- p) + c

p_d= 1.1 x p_s

p_s= 4.5 atm = 4.413 bar

p_d= 4.854 x 105 N/m^2

So, thickness will be

t= (4.854 x 10^5 x 0.325)/ (2 x 0.93 x 10^8 x 0.85 4.854 x 10^5)

= 1.0008 x 10^{-3} m

t= 1 m

Assuming, t_{std} = 3.2 mm

Checking this thickness for critical buckling pressure:

Pc = critical external buckling pressure

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

E = 2.2 x 106 N/cm^2 [For mild steel]

μ = 0.30

L = 6 In

D0 = Di + 2 x t

= 0.325 + 2 x 0.0032

= 0.3314 m

Putting values in the above equation of Pc we get

$P_c = 4.3553\ Kg/cm^2$

Taking factor of safety = 4

$P_{all} = P_c/4 = 4.355/4$

$= 1.0888\ Kg/cm^2$

Which is greater than 1 Kg/cm^2

Hence, this thickness is acceptable.

So, $t_{min} = 3.2$ mm

COSTING

Capacity of Plant = 100 ton/day

Annual production days = 300 days

6.1 Raw material cost

1) Ammonia:

Cost of Ammonia = Rs. 13171.2 / ton

Annual cost = Flow rate of Ammonia* Cost of Ammonia * Total production days = 2213.9 lacs

2) Carbon dioxide:

Cost of Carbon dioxide = Rs. 1568 / ton

Annual cost = Flow rate of Carbon dioxide* Cost of Carbon dioxide * Total production days

= 342 lacs

6.1.1 Equipment cost

Table 6.1: Equipment cost

Sr. No.	Equipment		Quantity	Cost (in Lac.)
1.	Jacketed (PFR)		1	306.69
2.	Evaporator	1^{st} Effect	1	387.02
		2^{nd} Effect	1	329.73
3.	Prilling Tower		1	125
4.	Stripper		1	105
5.	Separator	MPS	1	103.75
		LPS	1	129.3
6.	Storage tanks	NH_3	1	319.6
		CO_2	1	226.7
7.	Condensers		10	15.6
8.	Compressor			
9.	Blower			
10	Pumps			
TOTAL				2096.68

Total equipment cost = 2096.68 lacs

Delivery and transport cost = 10% of total equipment cost = 209.6 lacs

Total equipment purchase cost (PEC) é 2306.35 lacs

Now we will calculate direct and indirect costs

6.2 Calculation of Direct cost

Table 6.2: Direct cost (D)

Sr. No.	Equipment	% of PEC	Cost (in Lac.)
1.	Purchased equipment installation	39	817.7
2.	Instrumentation and control	30	629
3.	Electrical systems	10	209.67
4.	Yard improvements	10	209.67
5.	Land	4	83.86
6.	Maintenance & Repair	1.5	31.45
7.	Piping (Installed)	100	2096.68
8.	Auxiliary cost	40	838.67
TOTAL			4916.7

6.2.1 Raw material cost (RMC):

NH_3 Cost + CO_2 Cost = 2213.97 + 342

= 2555.97 lacs.

6.2.2 Labor Cost (LC):

LC = 12% of RMC

= 306.721acs.

Total direct cost= 4916.7 + RMC + LC

Total direct cost = 7779.39 lacs

6.3 Calculation of Indirect cost

Fable 6.3: Indirect cost (1)

Sr. No.	Equipment	% of PEC	Cost (in Lac.)

1.	Engineering and supervision	32	670.9
2.	Construction expenses	34	712.87

Total indirect cost = 1383.77 lacs

Direct cost (D) + Indirect cost (I) = 7779.39 + 1383.77

= 9163.161acs

Contingency = 8 % (D +I)

= 733.05 lacs

Contractors fees = 2 % (D +1)

= 183.26 lacs.

Fixed capital investment (FCI) = D + 1+ Contractors fees + Contingency

= 10079.47 lacs.

Working capita1= 20 % of total capital investment

Total capital investment (TCI) = FCI + WCI

TCI = 10079.47 + 0.2 TCI

TCI = 12599.341acs

Hence working capital (WCI) = 2519.87 lacs.

6.4 Estimation of Production cost

Capacity = 100 TPD of Urea * 300 days

= 30000 TPA

Total product cost (C_{TPC})

$C_{TPC} = C_{MC} + C_{GEC}$ Where,

C_{MC} = Manufacturing Cost

C_{GEC} = General Expenses Cost

$C_{MC} = C_{POC} + C_{DPC} + C_{FC}$

Where,

C_{POC} = Plant overhead cost = 15 % labor cost

C_{DPC} = Direct Production cost

C_{FC}: Fixed charges = 10 % C_{DPC}

C_{DPC} = Direct Material cost + Direct labor cost + Power & utility cost

C_{DPC} = 2555.97 + 306.72 + 838.67

= 3701.361acs

C_{FC}: 10 % *3701.36

= 370.1 lacs.

C_{POC}= 15 % * 306.72

= 46.008 lacs.

So, $C_{MC} = C_{POC} + C_{DPC} + C_{FC}$

= 46.008 + 3701.36 + 370.1

= 4117.47 lacs

$C_{GEC} = C_U + C_{DC}$

Where, Cu = Contingency cost

C_{DC} = Depreciation cost = 5 % of C_{MC}

C_{GEC} = Cu + C_{DC}

= 733.05 + 5 % * 4117.47

= 938.92 lacs.

$C_{TTPC} = C_{MC} + C_{GEC}$

= 4117.47 + 938.92

Total production cost = 5056.39 lacs/annum

6.5 Estimation of depreciation using double declining balance method [22]

$Va = V * (l\text{-}f)^{a}$

Va = asset value for particular year V = Fixed capital investment

n = service life = 20 years

f = 2/n.

1) For 0th year

V = 10079.471acé

$Va_o = 10079.47*(1\text{-}0.1)^0$

= 1007947 lacs.

For next year if we use. FCI as asset value, formula becomes

Va = V *(I-f)

2) For 1St year

$Va_l = Va_o * (l\text{-}t)$

= 10079.47 * 0.9

= 9071.5 lacs

$D_1 = Va_o\ Va_l$

= 1007.9 lacs

3) For 2“d year

$Va_2 = Va_1 * (l\text{-}t)^{n}$

= 9071.5 * (1 0.1)2

= 7347.92 lacs

$d_2 = Va_1 – Va_2$

$= 1723.59$ lacs

Similarly we can calculate depreciation for all next year

6.6 Estimation of profit

Market price = 24599.04 Rs/ton

Gross Profit = Selling Price -Production cost

= 7379.71 - 5056.39

= 2323.32 lacs / annum

As there are no taxes on depreciation, we will calculate profit taxable.

Profit taxable = Gross Profit - Depreciation = 2323.32 1007-9

=1315.421acs

Consider 25 % taxes are applicable on profit including all types of tax will calculate net profit Net profit = Profit taxable - 0.25 * Profit taxable

= 0.75 * Profit taxable

= 986.57 lacs

Similarly we calculate asset values, depreciation, profitable tax, net profit for each year to calculate cash flow.

Cash flow = Net profit + depreciation asset value **Table 6.4 Cash Flow**

Year (a)	Asset value	Depreciation	Net profit	Cash Flow
0.	10079.47	0	0	-10079.5
1.	9071.523	1007.947	986.5298	-7077.05
2.	7347.934	1723.589	449.798	-5174.55
3.	5356.644	1991.29	249.0225	-3116.33
4.	3514.494	1842.15	360.8777	-1311.47
5.	2075.273	1439.22	663.0747	27.02161
6.	1102.885	972.3881	1013.199	882.7016
7.	527.5067	575.3787	1310.956	1358.828
8.	227.0743	300.4323	1517.166	1590.524
9.	87.97325	139.1011	1638.164	1689.292
10	30.67437	57.29887	1699.516	1726.14
11	9.625944	21.04843	1726.704	1738.126
12	2.718651	6.907293	1737.31	1741.498
13	0.691045	2.027606	1740.969	1742.306
14	0.158089	0.532956	1742.09	1742.405
15	0.032549	0.12554	1742.369	1742.489
16	0.006031	0.026518	1742.47	1742.491
17	0.001006	0.005026	1742.486	1742.49
18	0.000151	0.000855	1742.489	1742.49
19	2.04E-05	0.000131	1742.49	1742.49
20	2.48E-06	1.79E-05	1742.49	1742.09

NEW ADVANCED TECHNIQUE

7.1 History

NFL was established in 1974, as a public-sector undertaking of the Government of India. NFL initially operated two plants in Bathinda and Panipat. In 1978, the ownership and operations of the Nangal plant of FCI was transferred to NFL. The Govt. of India, in 1984, entrusted the Company to execute the country’s first inland gas based fertilizer project of 7.26 lakh tons Urea capacity in District Guna of Madhya Pradesh and commercial production started from 01-07-1988.The company built and commissioned its Vijaipur plant in Guna district of Madhya Pradesh. The plant's capacity was doubled to 1.45 million tons in 1997.NFL was granted the Mini-Navratna status with additional autonomy due to its profitable operations.

In 2001, the company went public, with listings on the Bombay Stock Exchange and the National Stock Exchange.

Industrial and Business operations

NFL has five gas based Urea plants viz Nangal & Bathinda in Punjab,Panipat in Haryana and two at Vijaipur (Madhya Pradesh). The plants at Panipat, Bathinda & Nangal have recently been converted from fuel oil feedstock to natural gas. Vijaipur plants have also been revamped for energy saving & capacity enhancement. Products NFL is engaged in manufacturing and marketing of Urea, Neem Coated Urea, Bio Fertilizers (solid & liquid) and other allied Industrial products like Ammonia, Nitric Acid, Ammonium Nitrate, Sodium Nitrite, Sodium Nitrate etc.

The value-added Neem Coated Urea developed by NFL & widely recognized for its effectiveness is being produced at its three units at Panipat, Bathinda & Vijaipur. NFL is the first company in India to be permitted by the Government of India to produce and market Neem Coated Urea.

NFL manufactures and markets three types of Bio-Fertilizers namely Rhizobium, Phosphate Solubilizing Bacteria (PSB) and Azetobactor.Sta1ting with a mere 23 MT production in 1995-96, the production has risen to 231 MT (Approx.) in 2010-11.

7.2 Future Projects

An agreement for a Joint Venture Company "Ramagundam Fertilizers And Chemicals Limited" has been signed on 14-Jan-2013_bet\Neen National Fertilizers Limited, Engineers India Limited (EIL) and FCIL for setting up new Ammonia and Urea plants of 2200 MT per day & 3850 MT per day capacity respectively at the existing site of Ramagundam Fertilizer Plant in Distt. Karim Nagar in Telangana. As per this joint venture agreement, NFL shall provide commissioning, operation, maintenance and marketing services to the Joint Venture Company and EIL shall provide Engineering, Procurement and Construction Management (EPCM) Consultancy Services for the project. The total capital expenditure of 5,000 crore is being funded in Debt: Equity of 70:30 percentage NFL and EIL are contributing equity of 26% each and 11% is from FCIL. The balance equity is in the process of being tied up. The Company also has a Joint Venture (33.33% share) “Urvarak Videsh Limited” with Ms. KRIBHCO and RCF as promoters. The main objective of the joint venture company is to explore investment opportunity abroad and within the country in nitrogenous, phosphatic and potassic sectors and to render consultancy services for setting up projects in India and abroad.

7.3 Safety and Environment Management

NFL remains focused towards achieving sustained energy efficient operations of its ageing manufacturing facilities at the same time maintaining pollution free environment and process safety. A11 manufacturing Units of the Company continue to be ISO 9001-2008, IS014001-2004 and OHSAS-18001 certified, which indicates Company’s commitment to Quality Management System, Environment Management System and Occupational Health and Safety Systems.

PLANT LAYOUT

After the process flow diagrams are completed and before detailed piping, structural, and electrical design can begin, the layout of process units in a plant and the equipment within these process units must be planned. This layout can play an important part in determining construction and manufacturing costs, and thus must be planned carefully with attention being given to future problems that may arise. Since each plant differs in many ways and no two plant sites are exactly alike, there is no one ideal plant layout. Although conditions for example client specifications, schedule constraints, and availability of information may change significantly among the projects, the designer’s style remains constant. However, proper layout in each case will include arrangement of processing areas, storage areas, and handling areas in efficient coordination. Typical plant layout for Urea manufacturing plant is shown in figure 7.1. This includes tank farm for storage of raw material, processing area with extra space allowance for future expansion. Location of plant utilities, fire station, administrative office, canteen, workshop, and laboratory are shown in the layout. Entry and exit for raw materials and finished product are also shown in the layout along with emergency exit for any contingency.

Scale drawings, complete with elevation indications can be used for determining the best location for equipment and facilities. Elementary layouts are developed first. These shows the fundamental relationships between storage Space and operating equipment. The next step requires consideration of the safe operational sequence and gives a primary layout based on the flow of materials, unit operations, storage, and future expansion. By analyzing all the factors that are involved in plant layout, a detailed recommendation can be presented, and drawings and elevations, including isometric drawings of the piping systems, can be prepared [22].

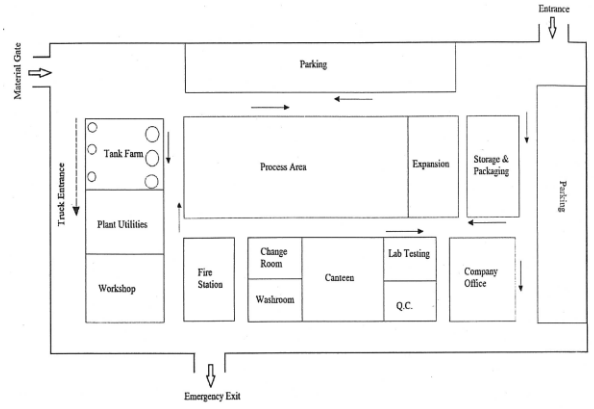


Fig. 7.1 plant layoutCONCLUSION

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 design of Plug flow reactor and multiple effect evaporators is done. Costing for whole plant is carried out and payback period is likely to be 6 years.

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