Design the manufacture of high purity anhydrous ammonia by a high pressure steam methane reforming process (Capacity -1200 tons/day.)

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Abstract

Process Description:

The steam –methane reforming process produces ammonia by steam reforming natural or refinery gas under pressure, followed by carbon monoxide shift, purification of raw synthesis gas and ammonia synthesis. In the process, saturated and unsaturated hydrocarbons are decomposed by steam according to the basis equation: $CH_4 + H_2O \rightarrow CO + 3H_2$

Feed streams high in olefins or sulphur require pretreatment. The primary reformer converts about 70% of a natural gas feed into raw synthesis gas, in presence of steam using a nickel catalyst. In the secondary reformer, air is introduced to supply the nitrogen required for production of ammonia. The heat of combustion of the partially reformed gas supplies energy to reform the remainder of the gas after reacting with oxygen in the air. High- pressure reforming conserves 30 to 40% in compressor horsepower over usual practices giving low –pressure synthesis gases. The mixture is then quenched and sent to the shift convertor. Here CO is converted to CO2 and H2. If heat is still available after satisfying the water requirement for the Shift reaction, a waste Boiler may be installed. Shift reactor effluent after heat recovery, is cooled and compressed, then goes to the gas purification section. CO2 is removed from the synthesis gas in a regenerative MEA (monoethanolamine) or other standard recovery system. After CO2 removal, CO traces left in the gas stream are removed by methanation. The resulting pure synthesis gas passes to the oil separators, is then passed through heat exchanger and charged to the catalytic ammonia convertor. Product gas is then passed through heat and exchanger and charged to the catalytic ammonia convertor. Product gas from the convertor is cooled primary separator and on further cooling, goes to the anhydrous ammonia then gets separated out in the primary separator and on further cooling, goes to the anhydrous ammonia product flash drum. The feed to the reforming section is normally is excess of 300 psig. The pressure however is not fixed and secondary reformers are 140^0 to 1800 ⁰F, while shift reaction temperatures are 700 to 850 ⁰F in the second stge. Ammonia synthesis is normally performed at inside the catalyst mass to allow a catalyst basket temperature gradient giving a maximum yield of ammonia per pass, regardless of production rate.

Key Words : Design, reforming process, Feed streams, Shift reaction, catalytic convertor, methanation, purification yield of ammonia,

Introduction

Contents:

The synopsis has been made on the basis of the following point as instructed by   IIChE- project.
Elaborating the whole idea of design Report has to be prepared on the following topics:

1. Literature Survey.
2. Detailed flow sheet.
3. Material balance(Assuming 95%operating factor)
4. Design of Heat Exchanger.
5. Design of Absorption column including Mechanical details.
6. Instrumentation and process control in the process plant.
7. Plant lay out.
8. Safety and pollution abetment aspects.
9. Cost Estimation.
10. Detailed Engineering Drawing of the Heat Exchanger and Absorption Column.

Modified Haber- Bosch Process

The ammonia concentration in the circulation gas leaving the catalyst is 10-11mole %. In the condenser the ammonia concentration is reduced by condensation to equilibrium at the exit temperature of the condenser. The condensed ammonia is separated from the circulating gas, and the gas is boosted in pressure by a circulating compressor to overcome the pressure drop in the synthesis loop.

    A brief outline of the process.

A typical modern ammonia-producing plant first converts natural gases (i.e., methane CH_4).The method for producing hydrogen from hydrocarbons is referred to as "Steam Reforming".Starting with a natural gas feedstock, the processes used in producing the hydrogen are:

(1) The first step in the process is to remove sulfur compounds from the feedstock because sulfur deactivates the catalysts used in subsequent steps. Sulfur removal requires catalytic hydrogenation to convert sulfur compounds in the feedstock's to gaseous hydrogen sulfide:

$H_2 + RSH \rightarrow RH + H_2S$ (gas)

(2) The gaseous hydrogen sulfide is then adsorbed and removed by passing it through beds of zinc oxide where it is converted to solid zinc sulfide:

$H_2S + ZnO \rightarrow ZnS + H_2O$
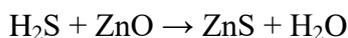
(3) Catalytic steam reforming of the sulfur-free feedstock is then used to form hydrogen plus carbon monoxide:

$CH_4 + H_2O \rightarrow CO + 3H_2$
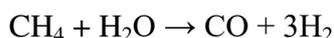
(4) The next step then uses catalytic shift conversion to convert the carbon monoxide to carbon dioxide and more hydrogen:

$CO + H_2O \rightarrow CO_2 + H_2$
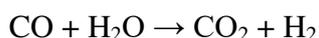
(5) The carbon dioxide is then removed either by absorption in aqueous ethanolaminesolutions or by adsorption in pressure swing absorbers (PSA) using proprietary solid adsorption media.

(6) The final step in producing the hydrogen is to use catalytic methanation to remove any small residual amounts of carbon monoxide or carbon dioxide from the hydrogen:

$CO + 3H_2 \rightarrow CH_4 + H_2O$
$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$
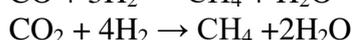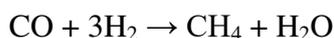
To produce the desired end-product ammonia, the hydrogen is then catalytically reacted with nitrogen (derived from process air) to form anhydrous liquid ammonia. This step is known as the ammonia synthesis loop (also referred to as the Haber-Bosch process):

$$3H_2 + N_2 \rightarrow 2NH_3$$
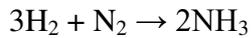
Due to the nature of the (typically multi-promoted magnetite) catalyst used in the ammonia synthesis reaction, only very low levels of oxygen-containing (especially CO, CO_2 and H_2O) compounds can be tolerated in the synthesis (hydrogen and nitrogen mixture) gas. Relatively pure nitrogen can be obtained by Air separation, but additional oxygen removal may be required.

Because of relatively low single pass conversion rates (typically less than 20%), a large recycle stream is required. This can lead to the accumulation of inert in the loop gas.

The steam reforming, shift conversion, carbon dioxide removal and methanation steps each operate at absolute pressures of about 25 to 35 bar, and the ammonia synthesis loop operates at absolute pressures ranging from 60 to 180 bar depending upon which proprietary design is used.

Ammonia Production

In 1909 Fritz Haber established the conditions under which nitrogen, $N_2(g)$, and hydrogen, H_2 (g), would combine using

- medium temperature (~500°C)
- very high pressure (~250 atmospheres, ~351kPa)
- A catalyst (a porous iron catalyst prepared by reducing magnetite, Fe_3O_4).
  Osmium is a much better catalyst for the reaction but is very expensive.

This process produces an ammonia, NH_3 (g), yield of approximately 10-20%.

The Haber synthesis was developed into an industrial process by Carl Bosch.

The reaction between nitrogen gas and hydrogen gas to produce ammonia gas is **exothermic**, releasing 92.4kJ/mol of energy at 298K (25°C).

$$N_2(g) + 3H_2(g) \xrightarrow{\text{heat, pressure, catalyst}} 2NH_3(g) \qquad \triangle H = -92.4 \text{ kJ mol}^{-1}$$

nitrogen + hydrogen → ammonia
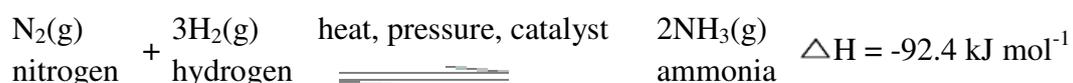
By **Le Chetalier's Principle:**

1. Increasing the pressure causes the equilibrium position to move to the right resulting in a higher yield of ammonia since there are more gas molecules on the left hand side of the equation (4 in total) than there are on the right hand side of the equation (2). Increasing the pressure means the system adjusts to reduce the effect of the change, that is, to reduce the pressure by having fewer gas molecules.
2. Decreasing the temperature causes the equilibrium position to move to the right resulting in a higher yield of ammonia since the reaction is exothermic (releases heat). Reducing the temperature means the system will adjust to minimize the effect of the change, that is, it will produce more heat since energy is a product of the reaction, and will therefore produce more ammonia gas. However, the rate of the reaction at lower temperatures is extremely slow, so a

higher temperature must be used to speed up the reaction which results in a lower yield of ammonia.

The equilibrium expression for this reaction(K_{eq}):
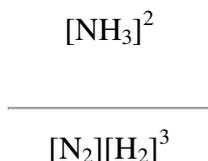
$$= \frac{[NH_3]^2}{[N_2][H_2]^3}$$

[As the temperature increases, the equilibrium constant decreases as the yield of ammonia decreases.]

Temperature (^{o}C)	K_{eq}
25	6.4×10^2
200	4.4×10^{-1}
300	4.3×10^{-3}
400	1.6×10^{-4}
500	1.5×10^{-5}

As the temperature increases, the equilibrium constant decreases as the yield of ammonia decreases.

Rate considerations:

1. A catalyst such as an iron catalyst is used to speed up the reaction by lowering the **activation energy** so that the N_2 bonds and H_2 bonds can be more readily broken.
2. Increased temperature means more reactant molecules have sufficient energy to overcome the energy barrier to reacting (activation energy) so the reaction is faster at higher temperatures (but the yield of ammonia is lower as discussed above). A temperature range of 400-500°C is a compromise designed to achieve an acceptable yield of ammonia (10-20%) within an acceptable time period.

(a) At 200°C and pressures above 750atm there is an almost 100% conversion of reactants to the ammonia product.

(b) Since there are difficulties associated with containing larger amounts of materials at this high pressure, lower pressures of around 200 atm are used industrially.

(c) By using a pressure of around 200atm and a temperature of about 500°C, the yield of ammonia is 10-20%, while costs and safety concerns in the building and during operation.
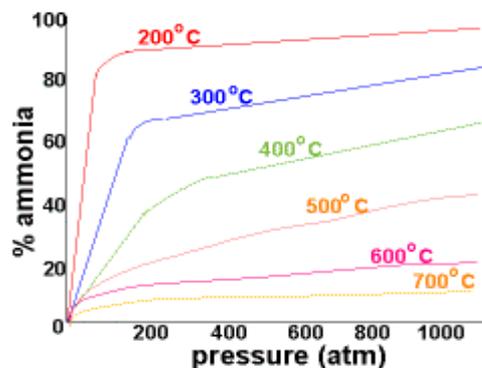
3. During industrial production of ammonia, the reaction never reaches equilibrium as the gas mixture leaving the reactor is cooled to liquefy and remove the ammonia. The remaining

mixture of reactant gases is recycled through the reactor. The heat released by the reaction is removed and used to heat the incoming gas mixture.

4. **Detailed flow sheet(Fig1: NH3 process)**
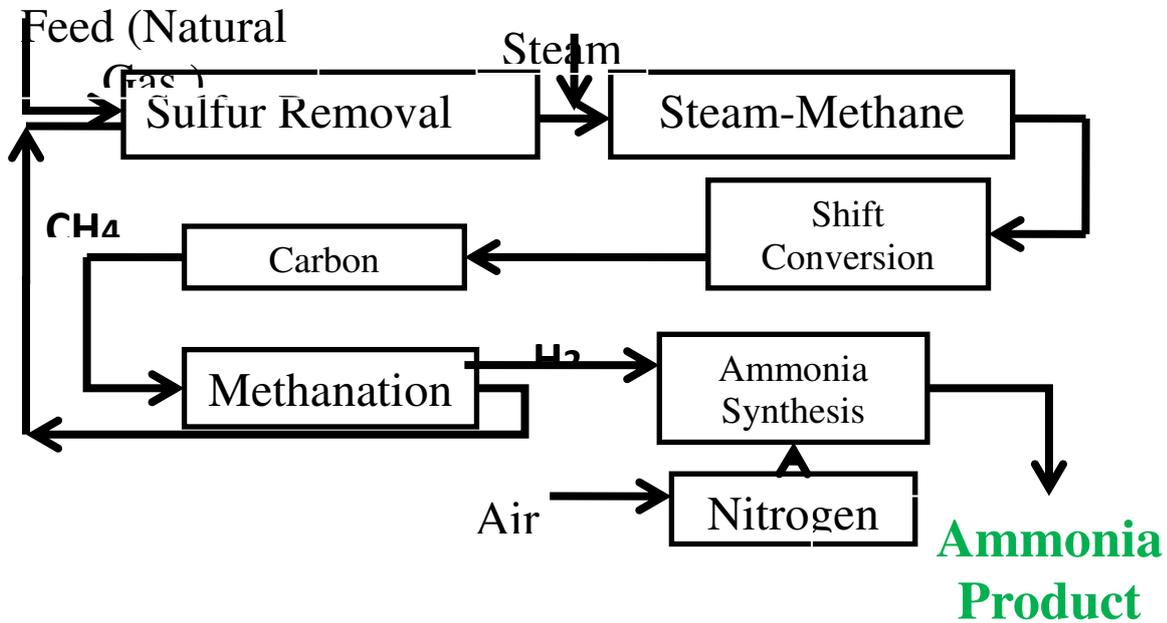
**A summary of process design of the equipment:**

A block diagram of the conventional steam reforming process is shown in Figure 1. (In some cases, a separate auxiliary boiler is required). For heavier feed stocks than naphtha (figure3), partial oxidation with oxygen is used in the synthesis gas production. A block diagram of a process description is given in the Figure1.About 85% of world ammonia production is based on steam reforming concepts A process description is given in Figure 2.

A summary of the mechanical design.

The following equipment is required for the mechanical design purposes as per figure 1 in the mechanical design:

(i)Compressor(ii) Reactor (a) Catalytic Beds (b) Heater.(iii)Heat Exchanger.(iv)Condenser.

(v)Refrigerated Unit.

FIGURES:
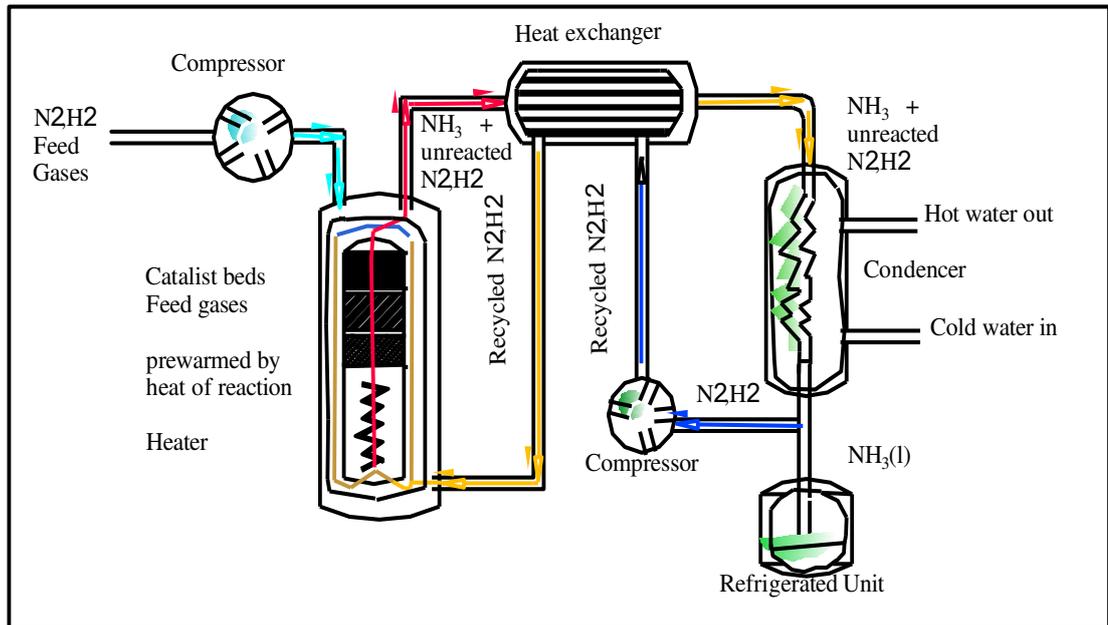
(Figure 2 : Auto cad design of modern NH3 production plant.)

More details the following equipment is required for the mechanical design purposes as per figure 2 in the mechanical design:

As per figure 1 the following mechanical equipment are required:

(1)Fired Heater (2) Reformer(3)Shift Convertor.(4)CO_2 Removal System.(5) Methanation Chamber.(6) Dryers and cold Box.(7)Synthetic gas compressor.(8)Ammonia Convertor.(9) Storage Tank.

Scope of the Design work required the following:Material balance of each of the segment, Material flow diagram of the preferred process, Heat balance of each of the segment, Mechanical design of the whole plant

➢ Material balance (Assuming 95%operating factor)

$$3H_2 + N_2 \rightarrow 2NH_3$$
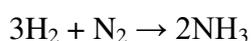
6   14   34

For 34 gm Ammonia we needs 6 gm Hydrogen

For 1200 Ton= 12 x10^8 gm Ammonia we need (6/34) x12x108= 211764705.88gm Of Hydrogen.

For 34 gm of Ammonia we need 14 gm of Nitrogen

For 1200 Ton= 12x10^8 gm Ammonia we need (14/34) x 12x10^8= 494117647.06 gm of Nitrogen.

As per the figure 9 flow chart the following calculation has done:

A. Material Balance

Ammonia Production = 1200 TPD=2941.18 kmol/hr
N_2+ 3H_2=   2 NH_3
Let , % conversion=20%
Which means, 1 kmol N_2 forms 2x0.2=0.4kmol NH_3
Assuming losses=8%
Amoeniaproduction=2941.18/0.92=3196.93kmol/hr
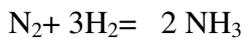
Hence, N_2 required=3196.93/0.4=7992.34kmol/hr
Air required=7992.34x100/79=10,117kmol/hr=10117x22.4=226621 Nm3/hr
 Siimilarly, H2 required=3x7992.34=23977kmol/hr

Material balance across H$_2$S absorber

Basis: 100 kmoles or Raw gas
CO = 42 kmole
H2= 51.23 kmole
H2S = 0.23 kmole
CO2= 5.7 kmole
N2= 0.18 kmole
Ar = 0.09 kmole
Now, maximum concentration of H_2S in H_2S free raw gas in limited to 0.3 ppm.
Amount of H_2S to be removed = 0.229997 kmole

In H_2S free gas

H_2= 51.23 kmole % of H2= 51.23/99.77x100 = 51.34

CO = 42 kmole % of CO = 42/99.31 X 100 = 42.09

H_2S =(0.23/99.31)x100= 0.3 ppm % of H_2S = 0.3 ppm
CO_2= 5.77 kmole % of CO_2= 5.77/99.31x 100 = 5.78

CH_4= 0.5 kmole % of CH_4= 0.5/ 99.31x 100 = 0.501

N_2= 0.18 kmole % of N2= 0.18/99.31x 100 = 0.18

Ar = 0.09 kmole % of Ar = 0.09/99.31x 100 = 0.09

Total no. of moles = 51.23 + 42.09 + 0.00003 + 5.77 + 0.5+0.18 + 0.09 = 99.77 kmol

Specification Sheet for Heat Exchanger design

Number required : 5

Function : To cool the gases coming out of shift Converter

Type : 1- 4 shell and tube with floating head.

Shell side:

Number of shell passes : 1

Shell side fluid : Hot Gases

Shell internal diameter : 1067 mm

Baffle spacing : 534 mm

Entering temp. of gases : 132^0C

Leaving temp. of gases : 40^0C

Tube side:

Number of passes : 4

Number of tubes : 738

Tube length : 4.88 m

Tube side fluid : Cooling Water

Tube pitch : 31.75 mm

Outside diameter of tube : 25.4 mm

Inside diameter of tube : 22.1 mm

Entering temperature of Water : 20^0C

Leaving temperature of Water : 35^0C

Reactor Design:A triply promoted (K_2O-CaO-Al_2O_3) iron oxide catalyst will be used. The iron oxide(Fe_2O_3-FeO) is in the form of non stoichiometric magnetite. It is made by fusing the magnetite with the promoters. The catalyst is reduced in situ, and the removal of oxygen yields a highly porous structure of iron with promoters present as inter phases between iron crystals and as porous clusters along the pore walls. The pores range from 500A to1000A and intra particle diffusion is diffusion to occur by the bulk mechanism. Alumina prevents sintering and corresponding loss of surface area and also bonds the K_2O, preventing its loss during use. The K_2O and CaO neutralize the acid character of Al_2O_3. Both K_2O and CaO decrease the electron work function of iron and increase it sability to chemisorb nitrogen by charge transfer to the nitrogen.
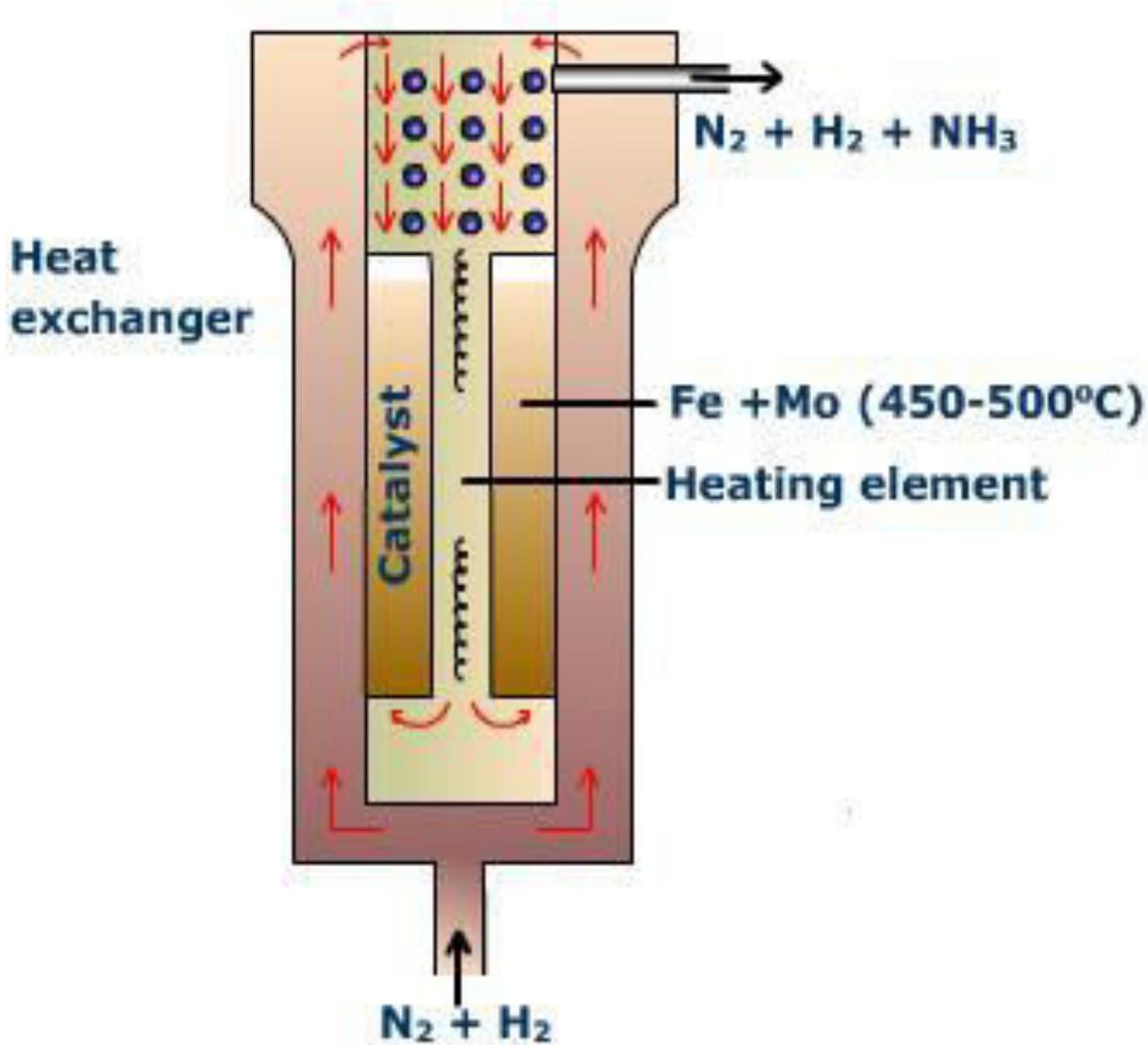
(Figure 3: Heat Exchanger)
CHEMISTRY AND KINETICS
The overall stoichiometric equation is
$1/2 N_2 + 3/2 H_2 = NH_3$
Extensive studies of ammonia synthesis on iron catalystssuggest that the reaction occur through the
following steps $N_2(g) = 2N$ (ads)
$H_2(g) = 2H$ (ads)
 N (ads) + H (ads)= NH (ads)
 NH (ads) + H (ads)= NH_2(ads)
 NH_2(ads) + H (ads)= NH_3(ads)
 NH_3(ads)= NH_3(g)
A rate equation based on nitrogen adsorption as the slow step is the most commonly used although
other forms have been development that also correlates the dataSince the effectivess factor of
ammonia catalyst is less that unity in a commercialsize pellets, it is desirafinely ground catalyst and
employ and effectiveness factor correction for other sizes. Ammonia synthesis is another example of
an old reaction with sufficient data existence to make this procedure feasible. The following equation
in termof activity has been recommended.
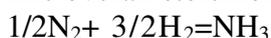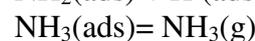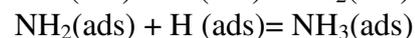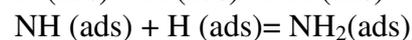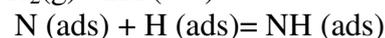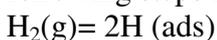
Input F12=28557.5kmol/hr $\rightarrow$ Plug Flow Reactor $\rightarrow$ Out put F13=25990.61 kmol/hr

Step -1
Reaction occurring in PFR
$N_2 + 3H_2 = 2NH_3$
Or A + 3B = C
Reactor Kinetics
The only reaction that needs to be considered is the formation of ammonia from H2 and N2,
$N_2 + 3H_2 \longleftrightarrow 2NH_3$
A common rate equation for ammonia synthesis is the Temkin-Pyzhev equation given
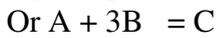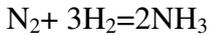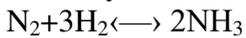
$$R = \int [\{(K_1 P_{N2} P_{H2}^{1.5}/P_{NH3}) - (K_2 P_{NH3}/P_{H2}^{1.5})\}]$$

Where Ris the rate of nitrogen consumption per unit volume of catalyst, f is a catalystactivity factor, and Pi is the partial pressure of component i in the gas. Values for K_1 and K_2:

$K_1 = k_{o1} \exp(-E1/RT)$

$K_2 = k_{o2} \exp(-E2/RT)$

$K_{o1} = 1.78954 \times 104$ kgmol/m3-hr-atm1.5

E1= 20,800 kcal/kgmol

$K_{o2} = 2.5714 \times 1016$ kgmol-atm0.5/m^3-hr

E2= 47,400 kcal/kgmol

One limitation of this rate equation is that the rate will be infinite if the amount of ammonia in the gas is zero, and this may occur in a reactor being fed with fresh make-upgas. To avoid this numerical problem, the rate equation may be multiplied by a factor $K_3 P_{NH3}/(1+K_3 P_{NH3})$; this will avoid the approach to infinity at low P_{NH3}
while having littleeffect at high ammonia pressures.
 This modification re-casts the rate equation into theLangmuir-Hinshelwood-Haougen-Watson (LHHW) form, which is one of the options for the RPLUG reactor block in Aspen

$$R = \int K_3 [\{(K_1 P_{N2}.P_{H2}^{1.5}) - (K_2 P_{NH3}^2/P_{H2}^{1.5})\}/\{(1+K_3 P_{NH3})\}]$$

Assume that f = 1.0 and
 $K_3 = 2$ atm-1.
Assume that the bulk density of the catalyst is 1200 kg/m^3 and that the catalyst costs\$12/kg.
 Now, reactor temperature is 450 degree Celsius
Therefore, k 1
= 0.00873 kmol/m^3-hr-atm1.5

K_2
=106.3 kmol-atm0.5/m^3-hr
Also, pi=pio(1-Xi)/(1+EiXi) ; where i=A,B,CE
A=(2-4)/2=-0.5 & X
B=3XA,XC=2XAUsing p

AO=0.2247*200=44.94atm,p

BO=0.6742*200=134.84atm,p

CO=0.06*200=12atm

Finally, we get

pA=44.94(1-XA)/(1-0.5XA)

pB=134.84(1-3XA)/(1-1.5XA)

pC=12(1-2XA)/(1-XA)

$$(-r\ A\ )=[\{1228.6(1-X_A)(1-3X_A)^{1.5}\}/\{(1-0.5X_A)(1-1.5X_A)^{1.5}\}-\{9.78(1-2X_A)2(1-1.5X_A)^{1.5}\}/\{\ (1-X_A)2(1-3XA)^{1.5}]/[1+24/(1-2X_A)/(1-X_A)]$$

 Now plug flow reactor design equation

$$dx_A/(-r\ A) = V/F_{AO}$$

Where F_{AO}-- feed rate of A

FromV/ F_{AO} =∫ dx_A/(-rA)

(limits from 0 to 0.2)

Site Layout   :(Figure 4: Plant Layout Blue Print)

Figure 4: Plant Layout Blue Print
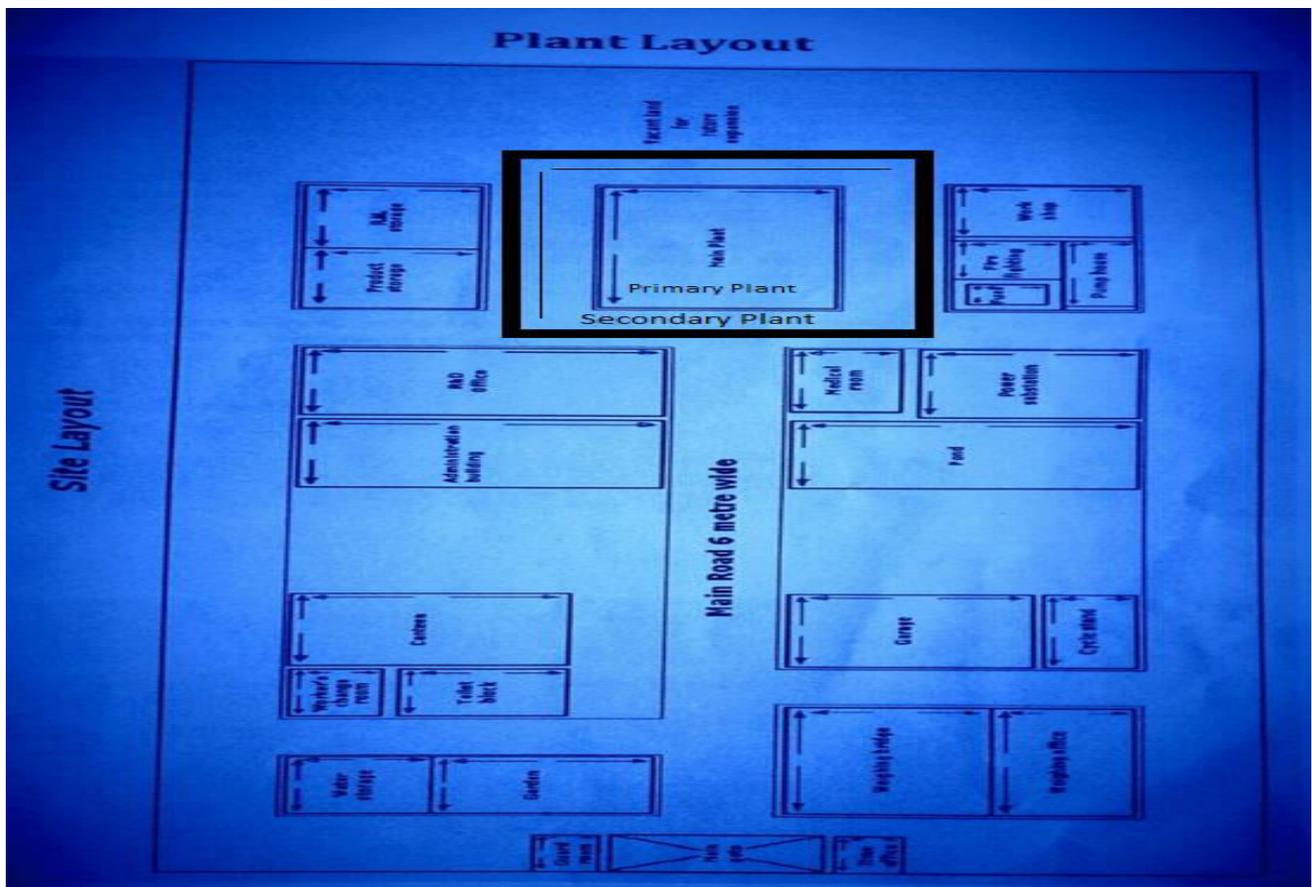
Cost Analysis

Cost estimation is a specialized subject and profession in its own right. Chemical plants a r e b u i l t to m a k e a profit. An estimate of the investment required and the cost of production needed before the profitability of project can be assessed.

Cost in plant can be classified as-

1.Fixed capital

2.Working capital

(1)Fixed capital:-

Fixed capital is the cost of the plant ready to start up.

It includes

1.Design

2. Equipment and interaction

3. Piping instrumentation and control system

4. Building and structure

5. Auxiliary facilities such as utilities

(2)Working capital

Working capital is additional investment needed over and above fixed capital to start up the plant. It includes

1. Start up 2. Catalyst cost 3. Raw material 4. Finished product inventories.

These costs are required to be corrected to the specific capacity and index price of the given year.Correction of index price

Cost in yr. A = Cost in yr. B x Cost index of yr. A /Cost index of yr. B
Correction of capacity $C2 = C1(S2/S1)^n$
$C2$ = Capital cost at capacity S_2
$C1$ = Capital cost at capacity S_1
n = power factor

(From Peter TimmerHaus, page 186)

Ammonia plant capacity = 100,000 ton/yr.= S_1
Fixed capital investment = 24 x 10^6 $ (Based on 1979)
Power factor for ammonia = 0.55
Our plant capacity = 1200TPD
Equation (2) $C_2 = C_1 (S2/S1)^n$

Taking that our plant runs for 335 days /yr.
deducting the time for normal shut down period.
Capacity = 1200 x 335 tons/yr. = 402000 tons/yr.=S_2
C_2=   24 x 10^6 x $(402000/100000)^{.55}$
= 51.59 x 10^6\$
Cost index = 1200
From "chemical engineering" journal

CI =585.9(based on 1979)

Cost in yr. 2017= cost in yr. 1979 x C.I. 2017/ C.I.1979

Thus,
Cost in yr. 2017 = (51.59 x 10^6) x 1200/585.9=105.663 x 10^6\$ =F.C
From Peter Timer Haus -   Working capital = 15% [fixed + working capital]

W.C.=(15/85)* F.C

W.C= (15/85)* 105.663 x 10^6\$

= 18.646 x 10^6\$

Therefore total capital investment= 105.663 x 10^6+ 18.646 x 10^6
= 124.31 x 10^6\$
Taking currency exchange rate as 1 \$ =60 Rs.
 Total capital investment
= 124.31 x 10^6x60=7458.6x 10^6
=Rs. 7458.6**crore**

Conclusion and Recommendation:

We have been asked to advice by assessing the economic profitability of the plant design to produce Ammonia against selling the hydrocarbon feed stock on the open market.

The basis for all decisions made in terms of the technical process design was based on two highly quantifiable factors efficiency and cost. In terms of the decisions we made, we choose to look first at the efficiency and then at the cost. In terms of efficiency, the quantifiable aspects we observed are conservations percentages, purity of separations and flow outputs.

In this work, it was presented an analysis of a distillation column with segmented weir sieve-tray of a 5 RT ammonia/water absorption refrigeration cycle. The simulations using the mathematical models were implemented in the program EES (Engineering Equation Solver). The method of Ponchon-Savarit, used to make the balance of mass and energy in the distillation column, appears appropriate as it provides accurate results. The stripping section provides a considerable increase in vapor concentration (51%),the enrichment section increases the concentration of vapor in 1.5% and the rectifier in 2.2%. The last two make a morere fined work on the purification process and should not be ignored.

The factor of reflux happens to be an important economic parameter in the design of distillation column because the number of plates in the column and the flow heat involved in the generator and

rectifier are significantly linked to it, much more for values near to 1. The study allowed calculating the geometry of the plate and the column. The distance between plates is sensitively linked to the height of the weir. The height of the column depends directly on the distance between plates, this secondone is a measure of the compaction of the column. The height of the weir influences directly the diameter of the column, the length of the weir, column efficiency, among others, and results are optimal for values between 4 mm and 8 mm, although for greater reliability it would to be used values greater than 8 mm. The diameter of the column is sensible to increase the vapor flow, and to a lesser extent to increase the liquid flow. For hole diameters larger than 4 mm, the geometry of the column shown callous, including pressure drop.

Our novel process is broken up into stages that effectively produce Ammonia at ASTM standards for shipment and sales. The process is broken up into single- step reaction scheme. The starting material Air or N2 or H2 are purified via distillation in a stage of reaction. The purified Ammonia is sent to the further reaction where hydrogen is removed via catalyst from CO to CO2 to next stage via separating using distillation columns. The ASTM standards state that the Ammonia purity must be at least 99.5wt%, with a maximum of 0.05wt% water and 0.5wt% of the total alcohols. The next final stage condenses the waste form the process to be sent off for disposal. Our major assumption was the composition of the waste feed from LDPE productions, which we assumed to be a hydrocarbon raffinate feed from the catalytic cracking of crude oil.

Findings from our Environmental, Health and hazard are the he materials involved in our process. Namely, they are hydrogen, Nitrogen, Ammonia. The next greatest hazard is the various ecological effects to the environment and toxicological effects in the work place of the materials. The equipment involved in our process design is built to contain these hazardous materials at all times under industry established stable conditions, with one instance of controlled and safe venting of a hydrogen-Ammonia vapour mixture. Health and safety of the plant personnel and the surrounding community are considered paramount and require OSHA regulated preventive, incident response, and disposal measures as describe in a section of this report.

Our economic analysis set us at a break –even point of 26.4 years with an NPV of $43.44MM at 12 years after start up. An NPV after a certain year after start up is defined as the amount of money earned above the investments and the earnings on those investments at a defined interest rate. In this way, this means that after the operating life of the plant has expired we still have not made a profit above the investor`s capital investment despite taking into account the time value of money. The breakeven point tell us that it will take almost two operating lives of the plant in order to make back the total capital investment given to us to start back the plant, which is not a possible option given that the typical operating period for a given design market, we found that that this generates revenue of $2.94MM per year. It is believed the operating costs and the capital cost to create a storage and shipment facility for the waste feed would be much cheaper than the revenue generated from selling it on the open market, which would create a profit for company.

In Summary IIChE team recommends that the NH_3 waste steam be sold on the open market rather than designing a plant to produce Ammonia under the conditions proposed. It should be noted that after making the changes outlined in the section alternatives, we estimate that costs will be lowered enough to profit from production of the NH_3. Further research into these changes should be conducted and

pilot scale experiments should be performed to evaluate the changes proposed in order to obtain more information about this process design.

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