UNIVERSITY OF DAYALA
COLLEGE OF ENGINEERING
CHEMICAL ENGINEERING DEPARTMENT

PRODUCTION OF AMMONIA

A PROJECT SUBMITTED TO THE COLLEGE OF ENGINEERING OF THE UNIVERSITY OF DIYALA IN PARTIAL FULFILLMENT OF THE REQUIREMENT FOR THE DEGREE OF BACHLOR OF SCIENCE IN CHEMICAL ENGINEERING

By

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بسم الله الرحمن الرحيم

"يَرْفَعِ اللَّهُ الَّذِينَ آمَنُوا مِنكُمْ وَالَّذِينَ أُوتُوا الْعِلْمَ دَرَجَاتٍ
وَاللَّهُ بِمَا تَعْمَلُونَ خَبِيرٌ "

صدق الله العظيم

سورة المجادلة الآية رقْم ١١.

الإهداء

إلى من خلقنا فسوانا ....الرحمن الرحيم.

إلى معلم الإنسانية ونبراس الهدى ....سيدنا محمد (صلى الله عليه واله وسلم).

إلى التي ترتاح لرؤيتها نفسنا وتسهل دعواتها كل امورنا، إلى التي بسط الله الجنة تحت قدميها ....الوالدة الحبيبة .

الى من نحت في صخر الحياة ليعبَد لنا درب العلم والمعرفة ،الذي كلامه يذلل لنا كل مصاعب الدنيا وهمومها ....الوالد الحبيب .

الى من لهما الفضل بعد الله تعالى ،الى من ساندنا ولم يدخرا جهدا في تعليمنا ....أساتذتنا الأعزاء.

الى الذين كانوا لنا خير سند وعون ....إخوتي وأخواتي .

الى كل من يحب هذه الأرض الطاهرة، ارض الأنبياء والأولياء لان جذوره غرست وامتدت فيها ....جميع العراقيين

نهدي ثمرة جهدنا المتواضع متمنيا القبول والحمد لله رب العالمين

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I wish also express my faithful and sincere respect to my honorable supervisor Lecture .Staar Ibrahem for his continuous guidance and permanent concern during my research.

Above all .thanks to God for the mercy and blesses he showed.

SUMMARY

Ammonia is one of the most highly produced inorganic chemicals, is alkaline gas has no color. Ammonia is also produced naturally from decomposition of organic matter, including plants, animals and animal wastes.

There is more than one way for the production of ammonia. In present project, A detailed literature survey, physical and chemical properties , and other specification were addressed.
we have to choose the way Haber in the production of ammonia to be the most common way as it is much easier than other economic. The Haber Process combines nitrogen from the air with hydrogen derived mainly from natural gas (methane) into ammonia. The reaction is reversible and the production of ammonia is exothermic.

the production capacity of the unit was 10000 ton / yr. this capacity is used as a basis in material & energy balance. Absorber column ,heat exchanger & reactor were designed .the dimensions of absorber column were 5m height & 0.87 diameter. the overall coefficient based on the outside area of the tube =891 w/m^2.c Shell diameter D_s =796mm in heat exchanger. Volume :27.81 m^3,Thickness of shell:34mm in reactor design.

Control system for CO_2 Absorber and Methanator were also designed.

Supervisor Certification

I bear witness that this project entitled " PRODUCTION OF AMMONIA " prepared under my supervision in the Chemical Engineering Department , Faculty of Engineering , University of Diyala , by (Shatha Mohamed , Shahad Ali, Eman Mahmood) as part of the requirements for obtaining a bachelor's degree in chemical engineering.

Signature :

Name :staar Ibrahem

Date : 2016 / 5 /10

A certificate from the Commission on coal

We are seeing that we read the project entitled "production " and as a committee examining We tested students (Shatha Mohamed , Shahad Ali , Eman Mahmood) contents which saw it meet the standards of a project for a bachelor's degree (Chemical Engineering) .

Signature :

Dr. Salah N. Farhan	***L.A Mustafa Sabah***
(Member)	***(Member)***

Assist. Prof. Dr. Anees A. Khadom

(Chairman)

(Chemical Engineering Department)

Date : 2016 / 5 /10

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Chapter one

Introduction

1.1 Introduction:

Ammonia (NH_3) is one of the most commonly produced industrial chemicals in the United States. It is used in industry and commerce, and also exists naturally in humans and in the environment. Ammonia is essential for many biological processes and serves as a precursor for amino acid and nucleotide synthesis. In the environment, ammonia is part of the nitrogen cycle and is produced in soil from bacterial processes. Ammonia is also produced naturally from decomposition of organic matter, including plants, animals and animal wastes(1).

Ammonia is alkaline gas has no color. Made up of one nitrogen and three parts hydrogen portion, which is lighter than air and has a characteristic pungent odor. It is the chemical symbol NH_3 and attend the distillation of coal or some nitrogenous substances. Ammonia gas does not burn in the air, but it ignites in oxygen occurs flame yellow weak Ammonia forms a minute proportion of the atmosphere; it is found in volcanic gases and as a product of decomposition of animal and vegetable matter. Because ammonia was formerly obtained by destructive distillation of horns and hooves of animals. Ammonia has also been called alkaline air and volatile alkali(2).

1.2 Ammonia production:

facilities provide the base anhydrous liquid ammonia used predominantly in fertilizers supplying usable nitrogen for agricultural productivity. Ammonia is one of the most abundantly-produced inorganic chemicals. There are literally dozens of large-scale ammonia production plants throughout the industrial world, some of which produce as much as 2,000 to 3,000 tons per day of anhydrous ammonia in liquid form. The worldwide production in 2006 was 122,000,000 metric tons. China produced 32.0% of the worldwide production followed by India with 8.9%, Russia with 8.2%, and the United States with 6.5%. Without such massive production, our agriculturally-dependent civilization would face serious challenges .

1.3 History:

Before the start of World War I, most ammonia was obtained by the dry distillation of nitrogenous vegetable and animal products; the reduction of nitrous acid and nitrites with hydrogen; and the decomposition of ammonium salt] by alkaline hydroxides or by quicklime, the salt most generally used being the ammonium chloride (sal-ammoniac).

There are scientific discoveries play an important role in the history of the world and of human activity and led to the growth of factories and communities. Thanks to these discoveries, the global community moved from one community to another simple scientifically advanced seeks to secure human needs through the rapid and massive production. Before the birth of the twentieth century a few years, the time factor as a key factor in the manufacturing process afternoon. Then scientists began in research to devise new ways of generating managed factory manufactured products at a faster pace than those adopted. Diversified products between the cars, chemicals and furniture and clothes and others. Between discoveries which aimed to increase the production of chemicals is one of the Haber process designed by the world, "Fritz Haber".

Haber sought, then, to develop a process for the manufacture of ammonia in large quantities of agricultural and military imperatives. This seemed impossible, but he managed to achieve this dream(3).

1.4 Uses of ammonia:.

About 80% or more of the ammonia produced is used for fertilizing agricultural crops in the form of aqua ammonia (an aqueous solution of ammonia), ammonium sulfate $(NH_4)_2SO_4$, ammonium phosphate $(NH_4)_3PO_4$, ammonium nitrate NH_4NO_3 and urea $(NH_2)_2CO$. Some anhydrous liquid ammonia is also used directly as a fertilizer(4). Ammonia is also used for:

- Manufacture of nitric acid (HNO_3)
- Manufacture of nylon and other polyamides
- Refrigerant in household, commercial and industrial refrigeration systems
- Manufacture of dyes
- Manufacture of explosive
- Cleaning solutions

1.5 Ammonia damage:

can be summarized ammonia damage are as follows:

If the inhalation of this gas, it can cause severe allergic respiratory and burning eyes with a severe cough, and may lead to the closure through the air and inflammation in the lungs with hoarseness, and if it is inhaled, a center may cause suffocation and then death. When the amount of it swallow it cause burns in the mouth, esophagus, stomach, and accompanied by severe abdominal pain and difficulty in swallowing, may be exposed to infected blood, accompanied by vomiting, and may cause a puncture in the esophagus and stomach. If this gas touched the skin it causes severe and serious burns. Ammonia intervention in creams for skin care industry, but these creams cause many damages including: skin color changed to yellow, and the appearance of dark spots, and causes blisters in the skin for those who suffer from allergies, as it causes hair loss; so it is advisable to stay away from all products contain ammonia. Other damage is still under study is that the ammonia gas cause cancer and infertility, but this has not been confirmed damage, and in spite of all the damage caused by ammonia gas which, however, has beneficial uses in human life(5) .

1.6 Physical Properties of Ammonia:

- Ammonia is a colorless gas.
- It has a pungent odor with and an alkaline or soapy taste. When inhaled suddenly, it brings tears into the eyes.
- It is lighter than air and is therefore collected by the downward displacement of air.
- It is highly soluble in water: One volume of water dissolves about 1300 volumes of ammonia gas. It is due to its high solubility

  in water that the gas cannot be collected over water.
- Liquid ammonia boils at 239.6 K (- 33.5°C) under one atmosphere pressure. It has a high latent heat of vaporization (1370 J per gram) and is therefore used in refrigeration plants of ice making machines.

- Liquid ammonia freezes at 195.3 K (-77.8°C) to give a white crystalline solid(6).

1.7Chemical Properties of Ammonia:

Thermal stability

Ammonia is highly stable. However, it can be decomposed into hydrogen and nitrogen by passing over heated metallic catalysts or when electric discharge is passed through it.

$$2NH_3 \rightarrow N_2 + 3H_2$$

Combustibility

Ammonia is combustible in air. However, it will burn in an atmosphere of oxygen

$$4NH_3 + 3O_2 \rightarrow 2N_2 + 6H_2O$$

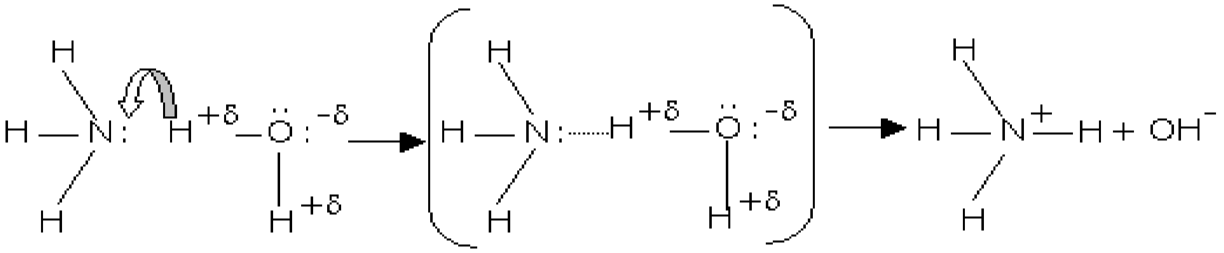
Nitric oxide is obtained when a mixture of ammonia and air is passed over platinum - rhodium catalyst at 800°C

$$4NH_3 + 5O_2 \rightarrow 4NO + 6H_2O$$

Basic character

Ammonia molecule has a strong tendency to donate its lone pair of electrons of nitrogen to other molecules. Thus, it acts like a strong Lewis base. In aqueous solutions, NH_3 ionizes in accordance with the reaction.

خطأ!



The equilibrium constant for this reaction at 298 K is 1.8 x 10^{-5}. Thus, ammonia ionizes to a very small extent in aqueous solution. The aqueous solution of ammonia acts as a weak base due to the presence of OH^- ions therein. Therefore, ammonia turns red litmus blue and reacts with acids to form salts.

For example,

$$NH_3 + HCl \rightarrow NH_4Cl$$

$$2NH_3 + H_2SO_4 \longrightarrow (NH_4)_2SO_4$$

With metal oxides

Ammonia gets oxidized to nitrogen, when passed over heated metal oxides.

$$3CuO + 2NH_3 \longrightarrow 3H_2O + N_2 + 3Cu$$

$$3PbO + 2NH_3 \longrightarrow 3H_2O + H_2 + 3Pb$$

With halogens

Ammonia reacts with halogens but the type of halogen and reaction conditions determine the nature of products.

Chlorine

Nitrogen and ammonium chloride are formed with a limited amount of chlorine. In the presence of excess of chlorine, nitrogen trichloride is formed.

$$8NH_3 + 3Cl_2 \text{ (limited)} \longrightarrow 6NH_4Cl + N_2$$

$$NH_3 + 3Cl_2 \text{ (excess)} \longrightarrow NCl_3 + 3HCl$$

Bromine

It gives ammonium bromide and nitrogen

$$8NH_3 + 3Br_2 \longrightarrow 6NH_4Br + N_2$$

Iodine

When rubbed with solid iodine, a dark colored precipitate of nitrogen tri-iodide is obtained

$$2NH_3 + 3I_2 \longrightarrow NH_3NI_3 + 3HI$$

With carbon dioxide (formation of urea)

Ammonia when heated under pressure with CO_2 gives urea.

$$2NH_3 + CO_2 \longrightarrow NH_2CONH_2 + H_2O$$

With alkali metals

When ammonia is passed over heated sodium or potassium, amides are formed and hydrogen is set free.

$2Na + 2NH_3 \longrightarrow 2NaNH_2 + H_2$

$2K + 2NH_3 \longrightarrow 2KNH_2 + H_2$

Alkali metal dissolved in liquid ammonia gives a blue solutions, which upon standing slowly, liberate hydrogen. The blue color of such solutions is due to the presence of solvated electrons ($e^-(NH_3)_n$). For instance, with sodium(6) .

$2Na + 2NH_3 \rightarrow 2NaNH_2 + H_2$

1.8 Methods production of Ammonia :

Ammonia is prepared in the industry in two ways:

- **1.8.1The first method (Way cyanamide) :**

Calcium cyanamide: is the inorganic compound with the formula $CaCN_2$. This calcium derivative of cyanamide (CN_2^{2-}) is used as fertilizer(7) .first synthesized in 1898 by Adolph Frank. It is commercially known as nitrolime.

Ammonia attend this way reactance CaC_2 calcium carbide powder and heated to 1100 ° C with nitrogen gas, was made up of calcium cyanamide $CaCN_2$ mixed with graphite

$$CaC_2 + N_2 \longrightarrow CaCN_2 + C$$

Heating calcium cyanamide with water under high pressure, ammonia and calcium carbonate consists.

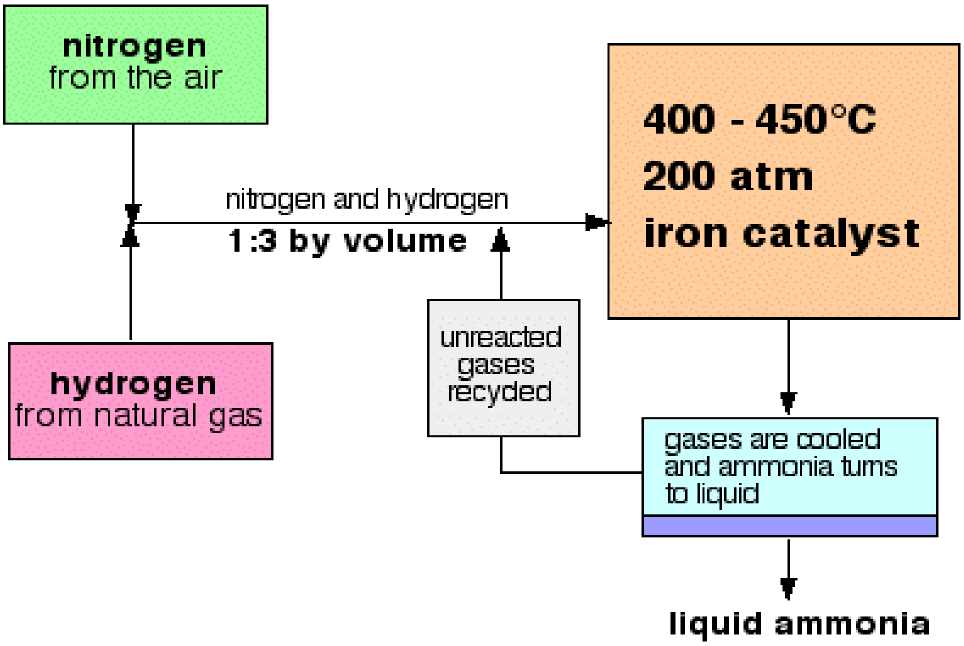
$$CaCN_2 + 3H_2O \longrightarrow CaCO_3 + 2NH_3$$

1.8.2The second method (Haber process):

The Haber Process combines nitrogen from the air with hydrogen derived mainly from natural gas (methane) into ammonia. The reaction is reversible and the production of ammonia is exothermic .

$$N_2(g) + 3H_2 \rightarrow 2NH_3(g) \qquad (\Delta H = -92.4\ kJ \cdot mol^{-1})$$

A flow scheme for the Haber Process looks like this:



Some notes on the conditions:

The catalyst

The catalyst is actually slightly more complicated than pure iron. It has potassium hydroxide added to it as a promoter - a substance that increases its efficiency.

The pressure

The pressure varies from one manufacturing plant to another, but is always high. You can't go far wrong in an exam quoting 200 atmospheres.

Recycling

At each pass of the gases through the reactor, only about 15% of the nitrogen and hydrogen converts to ammonia. (This figure also varies from plant to plant.) By continual recycling of the un reacted nitrogen and hydrogen, the overall conversion is about 98%.

The temperature

In order to get as much ammonia as possible in the equilibrium mixture, you need as low a temperature as possible. However, 400 - 450°C isn't a low temperature!

Rate considerations

In the absence of a catalyst the reaction is so slow that virtually no reaction happens in any sensible time. The catalyst ensures that the reaction is fast enough for a dynamic equilibrium to be set up within the very short time that the gases are actually in the reactor.

Starting with a natural gas feedstock, the processes used in producing the hydrogen are:

- 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 organic sulfur compounds (RSH) in the feed stocks to gaseous hydrogen sulfide (H_2S):

$$H_2 + RSH \rightarrow RH + H_2S \text{ (gas)}$$

- The gaseous hydrogen sulfide is then passed through a reactor containing zinc oxide (ZnO) where it is converted to solid zinc sulfide (ZnS) in a batch process (i.e., when all of the zinc oxide has been converted to zinc sulfide, it is removed and replaced with fresh zinc oxide):

$$H_2S + ZnO \rightarrow ZnS + H_2O$$

- Catalytic steam reforming of the sulfur-free methane (CH_4) feedstock is then used to form carbon monoxide (CO) plus hydrogen (H_2):

$$CH_4 + H_2O \rightarrow CO + 3H_2$$

- The next step then uses catalytic shift conversion to convert the carbon monoxide to carbon dioxide (CO_2) and more hydrogen:

$$CO + H_2O \rightarrow CO_2 + H_2$$

- The carbon dioxide is then removed either by absorption in aqueous ethanolamine solutions or by adsorption in pressure swing adsorbers (PSA) using proprietary solid adsorption media.
- 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 by converting them into methane:

$$CO + 3H_2 \rightarrow CH_4 + H_2O$$

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$$

- To produce the desired end-product ammonia, the hydrogen is then catalytically reacted with nitrogen (N_2) derived from process air to form anhydrous liquid ammonia (NH_3). This step is known as the "ammonia synthesis loop" (also referred to as the Haber-Bosch process):

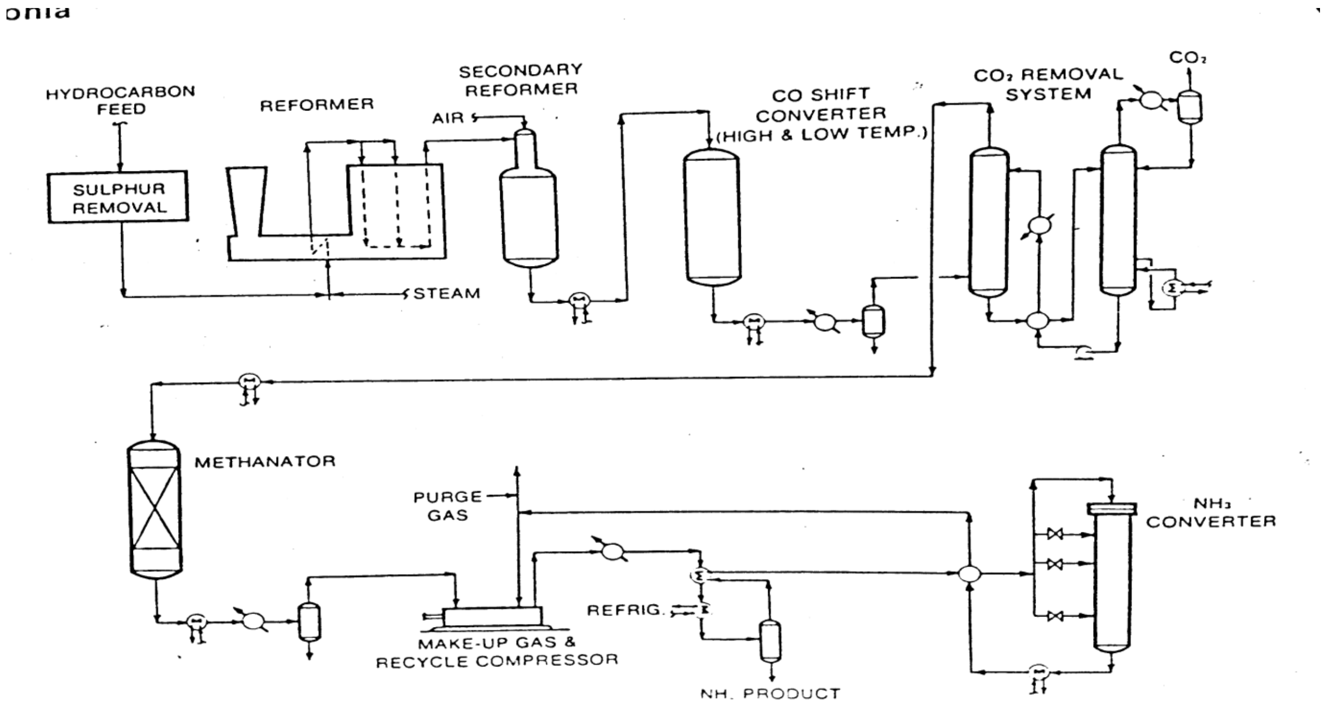
$$3H_2 + N_2 \rightarrow 2NH_3$$

The nitrogen required for the Haber-Bosch process is obtained in either of these two ways:

- The catalytic steam reformer has two sections: a primary reformer and a secondary reformer. By the introduction of air into the secondary reformer, the air reacts with some of the hydrogen produced in the primary reformer to form nitrogen and water :

$$H_2 + air \rightarrow N_2 + H_2O$$

- By the low-temperature distillation of air



Chapter two

Material & energy balance

2.1 Material balance:

production capacity of NH_3 = 10000 ton / yr

yr = 300 day
day = 24 hr
ton =1000 kg

P = (10000 ton / yr) * (1000 kg / ton)* (yr /300 day)* (day /24 hr)

P = 1389 kg /hr NH_3
M.W of NH_3 = 17 kg / k mol
P = 1389 /17
= 81.7059 k mol /hr

Reaction :
Primary reformer *conversion*
$CH_4 + H_2O \rightarrow CO + 3H_2$ $x_1 = 70\%$

Secondary reformer
$CH_4 + 1/2\ O_2 \rightarrow CO + 2H$ $x_2 = 29\%$ (8)

Shift converter
$CO + H_2O \rightarrow CO_2 + H_2$ $x_3 = 95\%$

Methanator
$CO + 3H_2 \rightarrow CH_4 + H_2O$ $x_4 = 4\%$
$CO_2 + 4H_2 \rightarrow CH_4 + 2\ H_2O$ $x_5 = 98\%$

Ammonia converter
$N_2 + 3H_2 \rightarrow 2NH_3$ $x_6 = 85\%$

Let:
Methane inlet to the process = F_1 kmol /hr
Conversion of CH_4 in reaction (1) = x_1
Conversion of CH_4 in reaction (2) = x_2
Conversion of CO in reaction (3) = x_3
Conversion of CO in reaction (4) = x_4
Conversion of CO_2 in reaction (5) = x_5
Conversion of H_2 in reaction (6) = x_6

H_2 produced
Primary reformer = $3F_1x_1$ kmol /hr

Secondary reformer = $2F_1x_2$

Shift converter $=(F_1x_1 + F_1x_2) x_3$

H_2 reacted
Methanator $=3(F_1x_1 + F_1x_2)x_4$

99% of CO_2 is removed in the CO_2 recovery unit

CO_2 inlet to the Methanator = CO_2 produced - CO_2 removed from *Shift* converter
$=(F_1x_1 + F_1x_2) x_3 - 0.99(F_1x_1 + F_1x_2) x_3$

$=0.01(F_1x_1 + F_1x_2) x_3$ kmol /hr

$=CO_2$ inlet to the methanator

H_2 reacted with CO_2 in methanator $=4(0.01)(F_1x_1 + F_1x_2) x_3x_5$
$=0.04(F_1x_1 + F_1x_2) x_3x_5$

H_2 inlet to the Ammonia converter = H_2 produced - H_2 reacted

$=3F_1x_1 + 2F_1x_2 + (F_1x_1 + F_1x_2) x_3$

$=3(F_1x_1 + F_1x_2) x_4 -0.04(F_1x_1 + F_1x_2) x_3x_5$

$= F_1[(3x_1 +2x_2) + (x_1 +x_2) x_3 -3(x_1 +x_2) x_4 -0.04(x_1 +x_2) x_3x_5]$

$=F_1[(3*0.7+2*0.29)+(0.7+0.29)(0.95)-3(0.7+0.29)(0.04)-0.04(0.7+0.29)(0.95*0.98)]$

H_2 inlet to the Ammonia converter = 3.4648324 F_1 kmol /hr

Ammonia converter:

NH_3 produced = 1389 kg/hr =81.7059 kmol /hr

H_2 reacted = (81.7059)*(3/2) =122.5589 kmol /hr

C0nversion of H_2 = reacted H_2 /input H_2

0.85 =122.5589 /3.4648324 F_1

F_1 = 41.6144 kmol /hr CH_4 *16

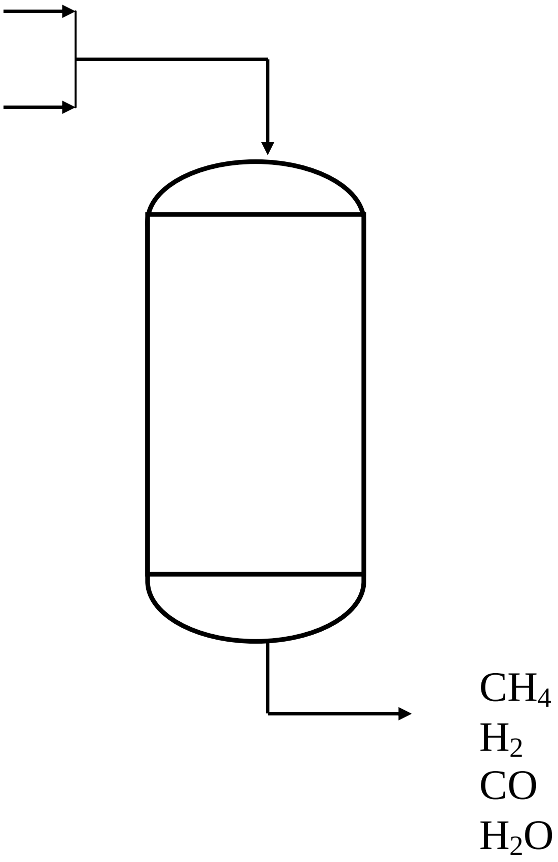
=665.83 kg/hr CH_4

<u>Material balance on Primary reformer :</u>

$CH_4 + H_2O \rightarrow CO + 3H_2$ $x_1 = 70\%$

41.6144 kmol /hr CH_4 F1

kmol /hr H_2O F2



Steam to methane ratio = 3/1
F_2/F_1 = 3/1
F_2/41.6144 = 3/1

F_2 =124.8432 kmol /hr H_2O *18
=2247.18 kg/hr

CH_4 reacted = F_1x_1
=41.6144 *(0.7)
=29.1301 kmol /hr *16

=466.08 kg/hr
H_2O reacted= 29.1301 kmol /hr *18
=524.34 kg/hr
CO Produce = 29.1301 kmol /hr * 28
=615.64 kg/hr

H_2 Produce =3*(29.1301)
=87.3903 kmol /hr *2
= 174.78 kg/hr

Exit product
CH_4 = inlet – reacted
=41.6144 -29.1301
=12.4843 kmol /hr *16
=199.75 kg/hr

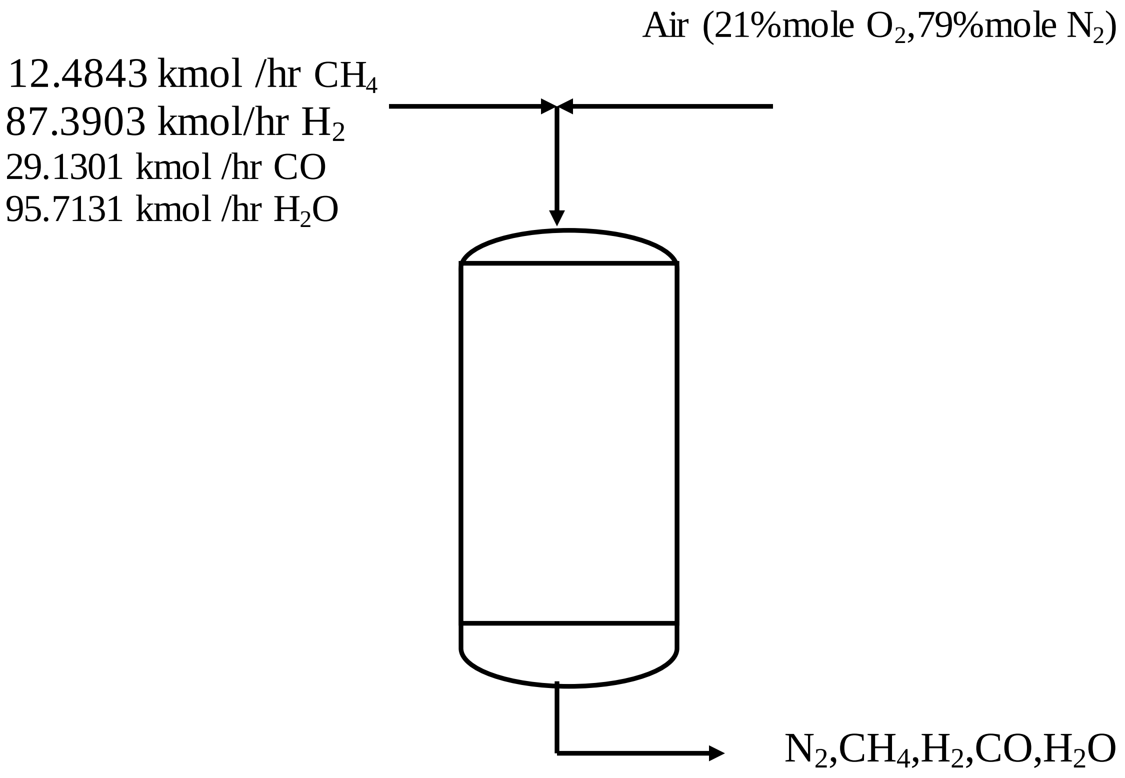
H_2O = inlet – reacted
=124.8432 – 29.1301
=95.7131 kmol /hr *18
=1722.84 kg/hr

CO =29.1301 kmol /hr *28
= 615.64 kg/hr
H_2 = 87.3903 kmol /hr *2
=174.78 kg/hr

Input			**output**	
Comp.	**Kmol/hr**	**Kg/hr**	**Kmol/hr**	**Kg/hr**
CH_4	**41.6144**	**665.83**	**12.4843**	**199.75**
H_2	**------**	**------**	**87.3903**	**174.78**
CO	**------**	**------**	**29.1301**	**815.64**
H_2O	**124.8432**	**2247.18**	**95.7131**	**1722.84**
	166.4576	**2913.01**	**224.7648**	**2913.01**

Material balance on secondary reformer:

$CH_4 + 1/2\ O_2 \rightarrow CO + 2H$ $x_2 = 29\%$



$H_2/N_2 = 3/1$(in ammonia converter)
H_2 inlet to ammonia converter =3.4648324 F_1
3.4648324(41.6144)=
144.1869 kmol /hr =

144.1869/N_2= 3/1
N_2 =48.0623 kmol /hr inlet to secondary reformer
= 48.0623 *18
1345.74 kg/hr=

O_2 inlet to secondary reformer

=48.0623(0.21/0.79)
=12.7761 kmol /hr *32
=408.84 kg /hr

CH_4 reacted = F_1x_2
=41.6144 (0.29)

=12.0682 kmol /hr

O_2 reacted =1/2(12.0682)
=6.0341 kmol /hr *32
=193.09 kg /hr

CO produce = 12.0682 kmol /hr *23
=337.91 kg /hr

H_2 produce = 2*(12.0682)

=24.1364 kmol /hr *2
=4827kg /hr

Exit product

CH_4 =inlet – reacted
=12.4843 – 12.0682
=0.4161 kmol /hr *16
=6.66 kg /hr

CO = inlet +produced
=29.1301+12.0682
=41.1983 kmol /hr *23
=1153.55 kg /hr

H_2= inlet + produced
=87.3903 +24.1364
=111.5267 kmol/hr *2
=223.05 kg /hr

O_2 =inlet – reacted
=12.7761-6.0341
=6.7420kmol/hr *32
=215.74 kg /hr

N_2 in = N_2 out
=22.6997 kmol /hr *28
=635.59 kg /hr

H_2O in = H_2O out
=95.7131 kmol /hr * 18
=1722.8 kg /hr

Input			***output***	
Comp.	***kmol /hr***	***kg/hr***	***kmol /hr***	***kg/hr***
N_2	**48.0623**	**1345.74**	**48.0623**	**1345.74**
O_2	**12.7761**	**408.84**	**6.7420**	**215.74**
CH_4	**12 .4843**	**199.75**	**0.4161**	**6.66**
H_2	**87.3903**	**174.78**	**111.5267**	**223.05**
CO	**29.1301**	**815.64**	**41.1983**	**1153.55**
H_2O	**95.7131**	**1722.84**	**95.7131**	**1722.84**
	285.5562	**4667.59**	**303.6585**	**4667.55**

Mol%
of CH4 dry basis =0.4161/(303.6585 -95.7131)*100%
=0.2%

Material balance on shift conversion:

48.0623Kmol/hr N_2

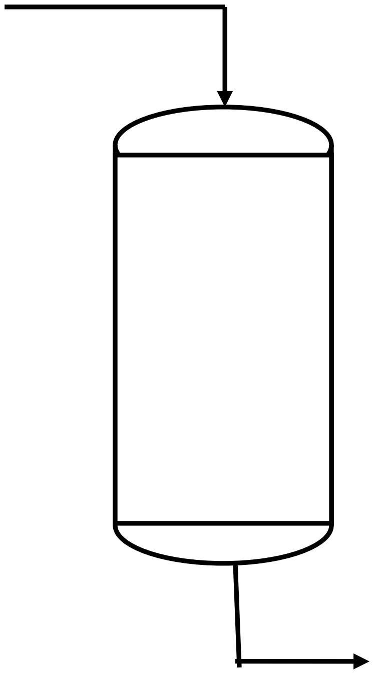
6.7420 Kmol/hr O_2

0.4161Kmol/hr CH_4

111.5267Kmol/hr H_2

41.1983Kmol/hr CO

95.7131Kmol/hr H_2O



N_2
O_2
CH_4
H_2
CO
CO_2
H_2O

$CO+H_2O \rightarrow CO_2+H_2$ X_3=95%

Co reacted = 41.1983(0.95)
= 39.1384Kmol/hr *28
= 1095.88Kg/hr
H_2O reacted = 39.1384Kmol/hr *18
= 704.49kg/hr
CO_2 produced = 39.1384 Kmol/hr *44
=1722.09 Kg/hr
H_2 produced = 39.1384 Kmol/hr *2
=78.27 Kg/hr
Exit products:

N_2 in = N_2 out

= 48.0623 Kmol/hr *28
= 1345.74Kg/hr
O_2 in = O_2 out
= 6.7420Kmol/hr *32
= 215.74 Kg/hr
CH_4 in = CH_4 out
=0.4161Kmol/hr *16
=666 Kg/hr

H_2= inlet + produced
= 111.5267 + 39.1384
= 150.6651 Kmol/hr *2
= 301.33 Kg/hr

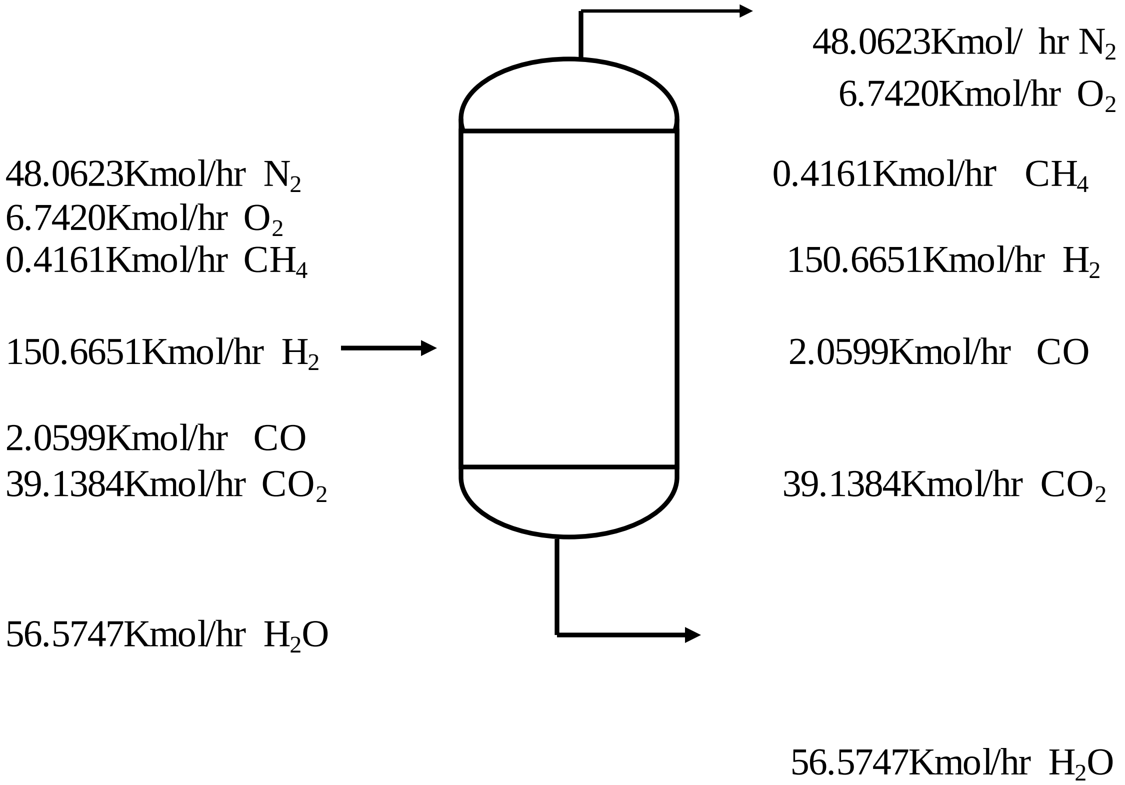
CO = inlet - reacted
= 41.1983-39.1384
= 2.0599Kmol/hr *28
= 57.68 Kg/hr

CO_2= inlet + produced
= 0+39.1384Kmol/h *44
= 1722.09 Kg/hr
H_2O in = inlet -reacted
= 95.7131-39.1384
= 56.5747 Kmol/hr *18
=1018.34 Kg/hr

Input			***output***	
Comp.	***kmol /hr***	***kg/hr***	***kmol /hr***	***kg/hr***
N_2	**48.0623**	**1345.74**	**48.0623**	**1345.74**
O_2	**6.7420**	**215.74**	**6.7420**	**215.74**
CH_4	**0.4161**	**6.66**	**0.4161**	**6.66**
H_2	**111.5267**	**223.05**	**150.6651**	**301.33**
CO	**41.1983**	**1153.55**	**2.0599**	**57.68**
CO_2	**-----**	**-----**	**39.1384**	**1722.09**

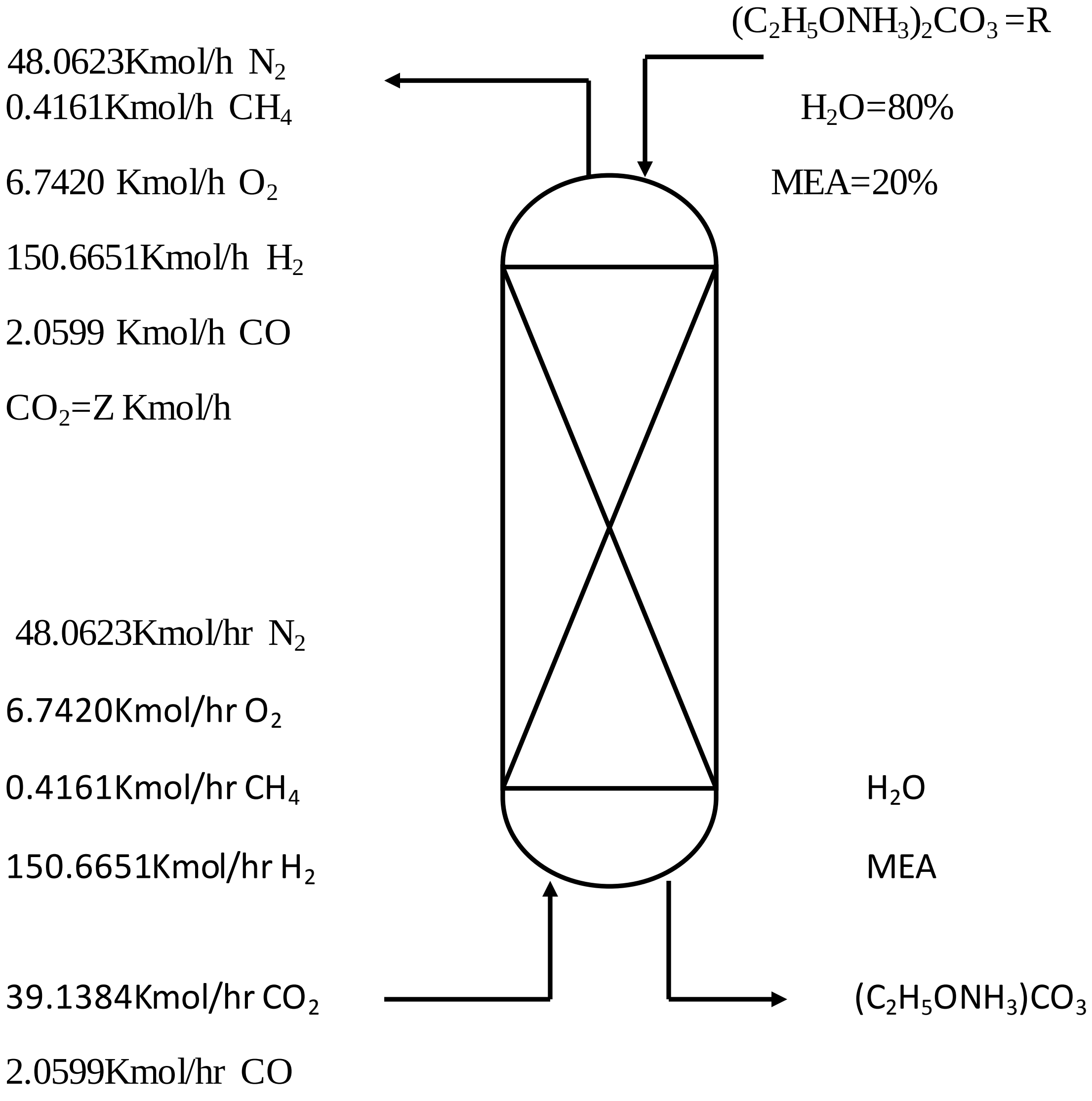
H_2O	**95.7131**	**1722.84**	**56.5747**	**1018.34**
	303.6585	**4667.58**	**303.6585**	**4667.58**

Material balance on gas liquid separator:



Material balance on absorber:

use 20 wt% MEA solution as solvent.
let recycle of $(C_2H_5ONH_3)_2CO_3$=R



$$2C_2H_7NO+CO_2+H_2O \rightarrow (C_2H_5ONH_3)_2CO_2$$

16 44 18 184

1Kmol of MEA per 0.40Kmol of CO_2 is used

Kmol MEA /Kmol CO_2 = 1/0.40
Kmol / 39.1384 = 1/0.40

MEA=97.8460 Kmol/hr inlet to absorber *61
=5968.61 Kg/hr
H_2O inlet to absorber = (5968.61) (0.80/0.20)
= 23874.44Kg/hr ÷18
=1326.3578Kmol/hr
99% recovery of CO_2
CO_2 recovery=0.99(39.1384)
= 38.7470 Kmol/hr

Un reacted CO_2 exit with gases :

Z=39.1384 - 38.7470
=0.3914Kmol/hr *44
=17.22 Kg/hr

MEA reacted =2(38.7470)
=77.4940 Kmol/hr *61
=4727.13 Kg/hr

H_2O reacted =38.7470 Kmol/h *18
= 697.446Kg/hr
$(C_2H_5ONH_3)_2CO_3$ produced = 38.7470Kmol/hr *184
=7129.45 Kg/hr

Exit liquid :
H_2O= inlet – reacted
=1326.3578 - 38.7470
=1287.6108Kmol/hr *18
=23176.99Kg/hr

MEA=inlet-reacted
= 97.8460 - 77.4940
= 20.3520 Kmol/hr *1 6
= 1241.47Kg/hr

$(C_2H_5ONH_3)_2CO_3$= inlet + produced
= R+38.7470 Kmol/hr*184
= 184R+7129.45 Kg/hr

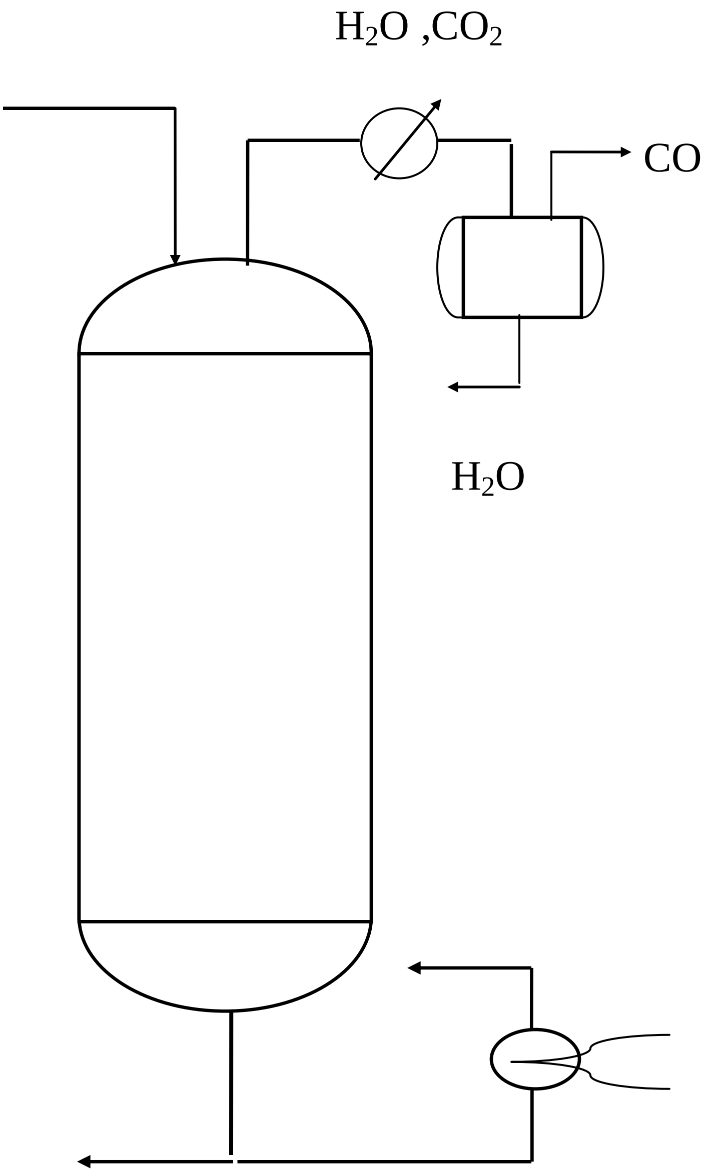
Input			*output*	
Comp.	*kmol /hr*	*kg/hr*	*kmol /hr*	*kg/hr*
N_2	**48.0623**	**1345.74**	**48.0623**	**1345.74**
O_2	**6.7420**	**215.74**	**6.7420**	**215.74**
CH_4	**0.4161**	**6.66**	**0.4161**	**6.66**
H_2	**150.6651**	**301.33**	**150.6651**	**301.33**
CO	**2.0599**	**57.68**	**2.0599**	**57.68**
CO_2	**39.1384**	**1722.09**	**0.3914**	**17.22**
H_2O	**1326.3578**	**23874.44**	**1287.6108**	**23176.9**
MEA	**97.846**	**5968.61**	**20.3520**	**1241.4**
$(C_2H_5ONH_3)_2CO_3$	**R**	**184R**	**R+38.7470**	**184R+7129.45**
		33492.29+184R		**33492.28+184R**

Material balance on stripper:

1287.6108Kmol/hr H_2O
20.3520Kmol/h MEA

R+38.7470Kmol/hr

$(C_2H_5ONH_3)_2CO_3$



H_2O

MEA

$(C_2H_5ONH_3)_2CO_3$

$(C_2H_5ONH_3)_2CO_3 \rightarrow 2\ C_2H_5NO + CO_2 + H_2O$

184 16 44 18

CO_2 produced = 38.7470 Kmol/hr *44
=1704.87 Kg/hr

H_2Oproduced = 38.7470Kmol/hr *18
= 697.45Kg/hr

MEA produced =2(38.7470)
=77.4940 Kmol/hr*61
=4727.13Kg/hr

$(C_2H_5ONH_3)_2CO_2$ reacted = 38.7470Kmol/hr *184
=7129.45Kg/hr

99% recovery of $(C_2H_5ONH_3)_2CO_3$
0.99 (R+38.7470) = 38.7470
R = 0.3914Kmol/hr $(C_2H_5ONH_3)_2CO_3$*184
=72.02 Kg/hr

Exit liquid :

H_2O = inlet + produced
= 1287.6108 + 38.7470
= 1326.3578Kmol/hr *18
= 23874.44Kg/hr

MEA= inlet+ produced
= 20.3520 + 77.4940
= 97.8460 Kmol/hr*61
= 5968.61 Kg/hr

$(C_2H_5ONH_3)_2CO_3$ =inlet-reacted
=(0.3914+38.7470)-38.7470
= 0.3914Kmol/hr *184
=72.02 Kg/hr

Input			*Output*	
Comp.	***kmol /hr***	***kg/hr***	***kmol /hr***	***kg/hr***
CO_2	**-----**	**-----**	**38.7470**	**1704.87**
H_2O	**1287.6108**	**23176.99**	**1326.357**	**23874.44**
MEA	**20.3520**	**1241. 47**	**97.8460**	**5968.61**
$(C_2H_5ONH_3)_2CO_3$	**39.1384**	**7201.47**	**0.391**	**72.02**
		31619.93		**31619.93**

Material balance on methanator :

48.0623Kmol/h N_2

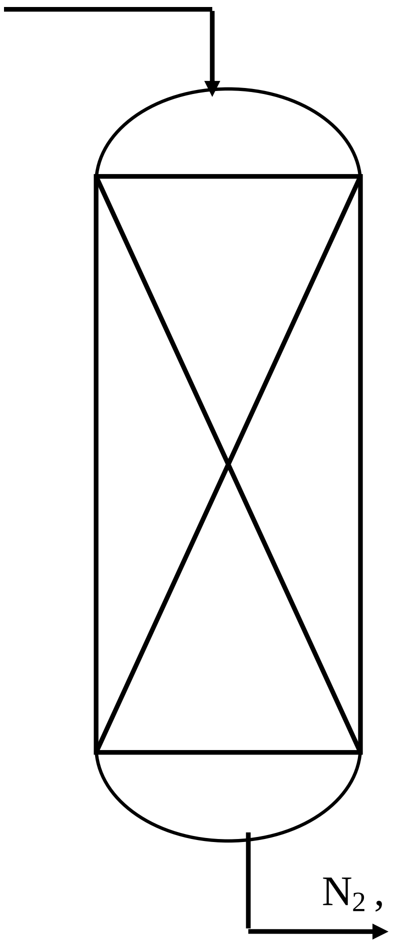
6.7420Kmol/h O_2

0.4161Kmol/h CH_4

150.6651Kmol/h H_2

2.0599Kmol/h CO

0.3914 Kmol/h CO_2



N_2 , H_2, CH_4,

O_2, CO, CO_2

H_2O

Reaction 4

$CO+3H_2 \rightarrow CH_4+H_2O$

$CO_2+4H_2 \rightarrow CH_4+2H_2O$

Co reacted = 41.1983X $_4$
=41.1983 (o.o4)
= 1.6479 Kmol/hr *28 = 46.14 Kg/hr
H_2 reacted = 3(1.6479)
= 4.9437 Kmol/hr *2
= 9.89 Kg/hr

CH_4 produced = 1.6479 Kmol/hr *16

=26.37 Kg/hr

H_2o produced = 1.6479 Kmol/hr *44

= 29.66Kg/hr

Reaction 5

Co_2 reacted = O .3914 X_5

=O.3914 (0.98)

= O.3836 Kmol/hr *44

= 16.88 Kg/hr

H_2 reacted = 4(0.3836)

= 1.5344 Kmol/hr *2

= 3.07 Kg/hr

CH_4 produced = 0.3836 Kmol/hr *16

=6.14Kg/hr

H_2O produced = 2(0.3836)

= 0.7672Kmol/hr *18

= 13.81 Kg/hr

Exit Products

N_2 = 48.0623Kmol/hr

= 1345.74 Kg/hr

O_2 = 6.7420Kmol/hr

= 215.74 Kg/hr

CH_4 = inlet + produced

= 0.4161 + (1.6479 + 0.3836)

= 2.4476 Kmol/hr *16

= 39.16 Kg/hr

H_2 = inlet – reacted

= 150.6651 – (4.9437 + 1.5344)

= 144.1870 Kmol/hr *2

= 288.37 Kg/hr

Co = inlet – reacted

= 2.0599 – 1.6479

= 0.4120 Kmol/hr *28

= 11.54 Kg/hr

Co_2 = inlet – reacted

= 0.3914 – 0.3814

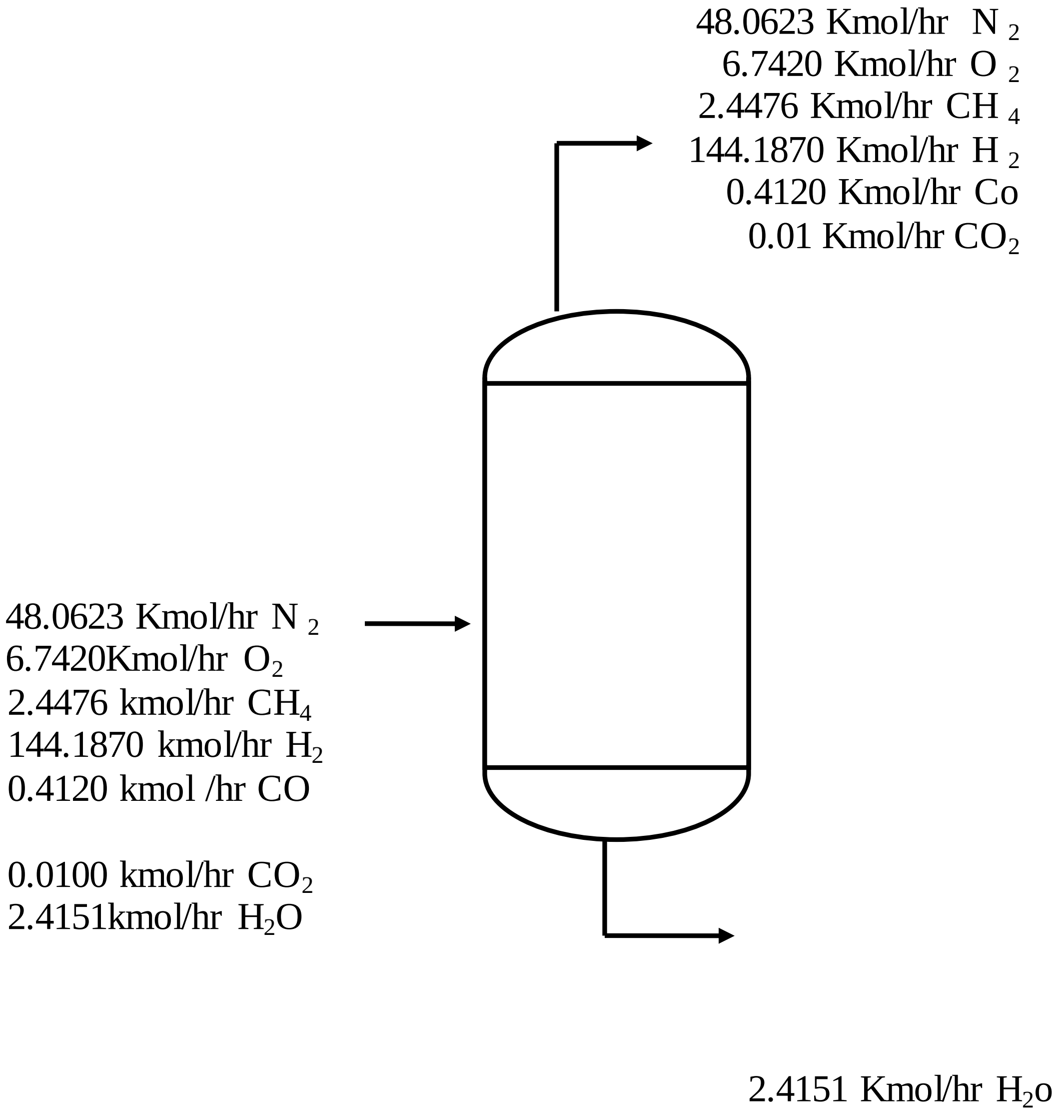
= 0.0100 Kmol/hr *44

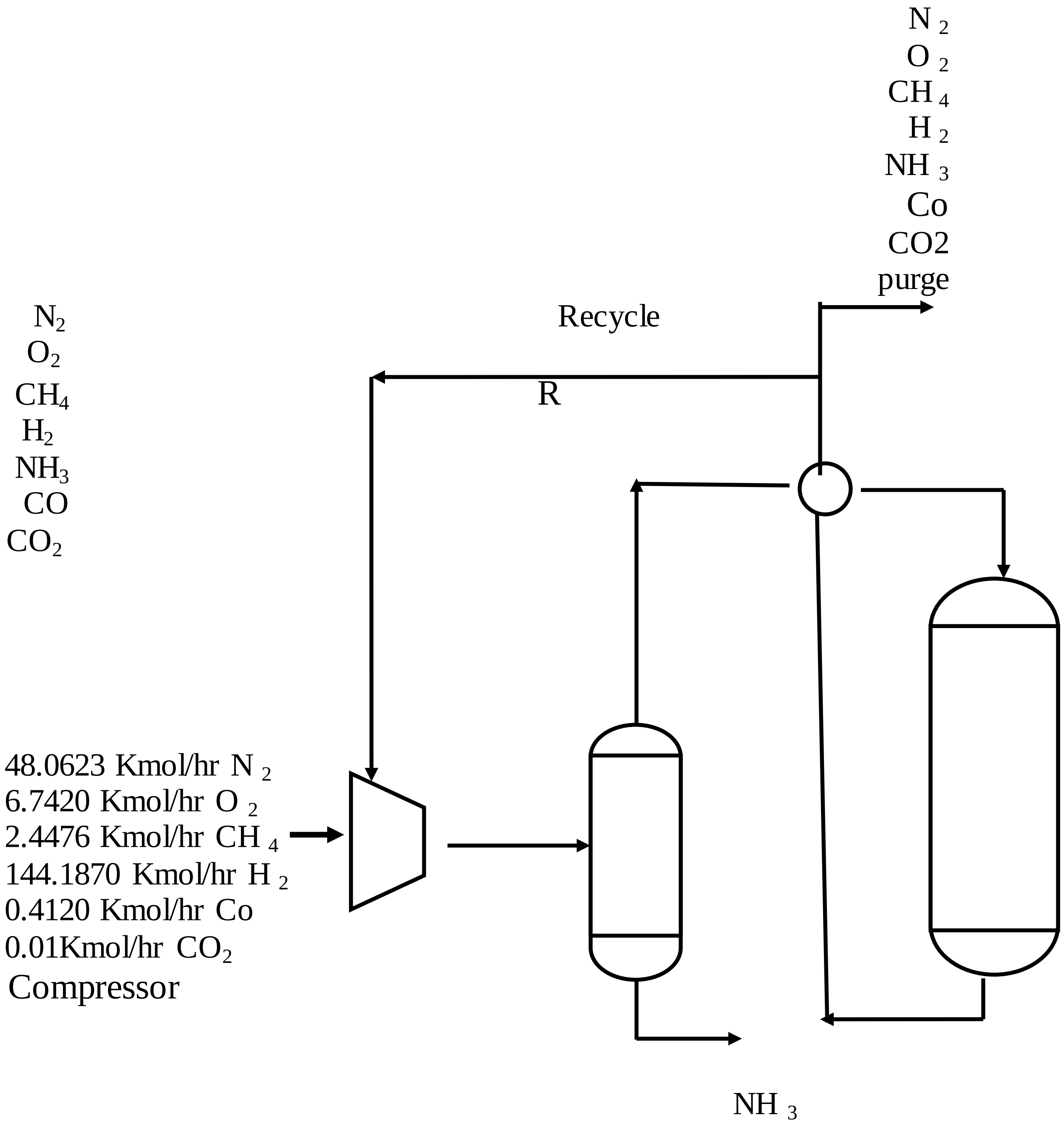
= 0.44 Kg/hr

H_2o = inlet + produced
= 0 + (1.6479 + 0.7672)
= 2.4151 Kmol/hr *18
= 43.47 Kg/hr

input			**Output**	
Comp.	**Kmol/hr**	**Kg/hr**	**Kmol/hr**	**Kg/hr**
N_2	**48.0623**	**1345.74**	**48.0623**	**1345.74**
O_2	**6.7420**	**215.74**	**6.7420**	**215.74**
CH_4	**0.4161**	**6.66**	**2.4476**	**39.16**
H_2	**150.6651**	**301.33**	**144.1870**	**288.37**
Co	**2.0599**	**57.68**	**0.4120**	**11.54**
Co_2	**0.3914**	**17.22**	**0.0100**	**0.44**
H_2o	**-------**	**------**	**2.4151**	**43.47**
		1944.37		***1944.46***

Material balance on gas_liquid separator 2:





REACTION

$N_2 + 3H_2 \rightarrow 2NH_3$ $x_6 = 85\%$

Overall conversion of H_2 =85%

H_2 reacted = 144.1870 X 6
= 144.1870 (0.85)

= 122.5589 Kmol/hr *2
= 245.12 Kg/hr
N_2 reacted = 1/3(122.5589)
= 40.8530 Kmol/hr *28
= 1143.88 Kg/hr
NH_3 produced = 2/3 (122.5589)
= 81.7059 Kmol/hr *17
= 1389 Kg/hr

PURGE

N_2 = inlet – reacted
= 48.0623 – 40.8530
= 7.2093 Kmol/hr
O_2 = 6.7420 Kmol/hr (inert)
CH_4 = 2.4476 Kmol/hr (inert)
H_2 = inlet – reaction
= 144.1870 – 122.5589
= 21.6281 Kmol/hr
Co = 0.4120 Kmol/hr (inert)
Co_2 = 0.0100 Kmol/hr (inert)
NH_3 = e Kmol/hr

Single pass conversion 0f H_2 = 20%
Single pass conversion = H_2 reacted / H_2 input to reactor
Let H_2 in recycle stream = r Kmol/hr
H_2 input to reactor = 144.1870 + r Kmol/hr ∴
H_2 reacted = 122.5589 Kmol/hr

0.2 = 122.5589/144.1870 + r

r = 468.6075 Kmol/hr H_2
Recycle to purge ratio R/B

R/B = 468.6075 / 2106281 = 21.6666

Recycle stream
N_2 = 21.6666 (7.2093)
= 156.2010 Kmol/hr
O_2 = 21.6666 (6.7420)
= 146.0762 Kmol/hr
CH_4 = 21.6666 (2.4476)
= 53.0312 Kmol/h

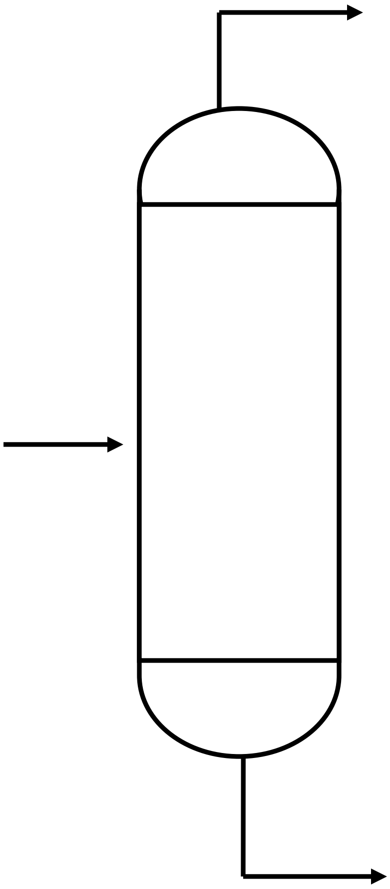
H_2 = r = 468.6075 Kmol/hr
Co = 21.6666 (0.4120)
= 8.9266 Kmol/hr
Co_2 = 21.6666 (0.0100)
= 0.2167 Kmol/hr
NH_3 produced from reactor = 81.7059 Kmol/hr

NH_3 in recycle stream = 81.7059 * 21.6666/ 21.6666 + r

= 78.1012 Kmol/hr
NH_3 in purge stream e = 81.7059 – 78.1012
= 3.6047 Kmol/hr

Material balance on separator :

To reactor
204.2633b Kmol/hr N_2
152.8182 Kmol/hr O_2
55.4788 Kmol/hr CH_4
612.7945 Kmol/hr H_2
9.3386 Kmol/hr Co
0.2267 Kmol/hr Co_2



Fresh feed + recycle
204.2633 kmol/hr N_2

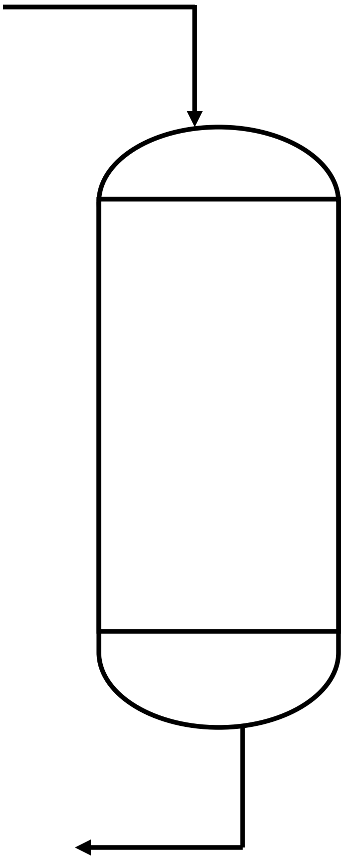
152.8182 kmol/hr O_2
55.4788 kmol/hr CH_4
612.7945 kmol/hr H_2
9.3386 kmol/hr Co
0.2267 kmol/hr Co_2
78.1012 kmol/hr NH_3

78.1012 Kmol/hr *17 NH_3
= 1327.72

Material balance on reactor:

All stream are calculated

204.2633 Kmol/hr N_2
152.8182 Kmol/hr O_2
55.4788 Kmol/hr CH_4
612.7945 Kmol/hr H_2
9.3386 Kmol/hr Co
0.2267 Kmol/hr Co_2



163.4103 Kmol/hr N_2
152.8182 Kmol/hr O_2
55.4788Kmol/hr CH_4
490.2356 Kmol/hr H_2
9.3386 Kmol/hr Co
0.2267 Kmol/hr Co_2
81.7059 Kmol/hr NH_3

Input			output	
Comp.	Kmol/hr	Kg/hr	Kmol/hr	Kg/hr
N_2	204.2633	5719.37	163.4103	4575.49
O_2	152.8182	4890.18	152.8182	4890.18
CH_4	55.4788	887.66	55.4788	887.66
H_2	612.7945	1225.59	490.2356	980.47
Co	9.3386	261.48	9.3386	261.48
Co_2	0.2267	9.97	0.2267	9.97
NH_3	--------	-------	81.7059	1389
		12994.25		*122994.25*

2.2 Energy Balance:

Gas heat capacity :

$Cp = a + bT + cT^2 + dT^3$

Where : Cp Gas heat capacity KJ/K mol .k

T: temperature k

a, b , c and d constants

component	**A**	**B**	**c**	**d**
N_2	**31.150**	$-1.356*10^{-2}$	$2.679*10^{-5}$	$-1.168*10^{-8}$
O_2	**28.106**	$-3.680*10^{-6}$	$1.745*10^{-5}$	$-1.065*10^{-8}$
CH_4	**19.251**	$5.212*10^{-2}$	$1.197*10^{-5}$	$-1.131*10^{-8}$
H_2	**27.143**	$9.273*10^{-3}$	$-1.38*10^{-5}$	$7.645*10^{-9}$
CO	**30.869**	$-1.285*10^{-2}$	$2.789*10^{-5}$	$-1.271*10^{-8}$
CO_2	**19.795**	$7.343*10^{-2}$	$-5.6*10^{-5}$	$1.715*10^{-8}$
NH_3	**27.315**	$2.383*10^{-2}$	$1.707*10^{-5}$	$-1.184*10^{-8}$
H_2O	**32.243**	$1.923*10^{-3}$	$1.055*10^{-5}$	$-3.596*10^{-9}$

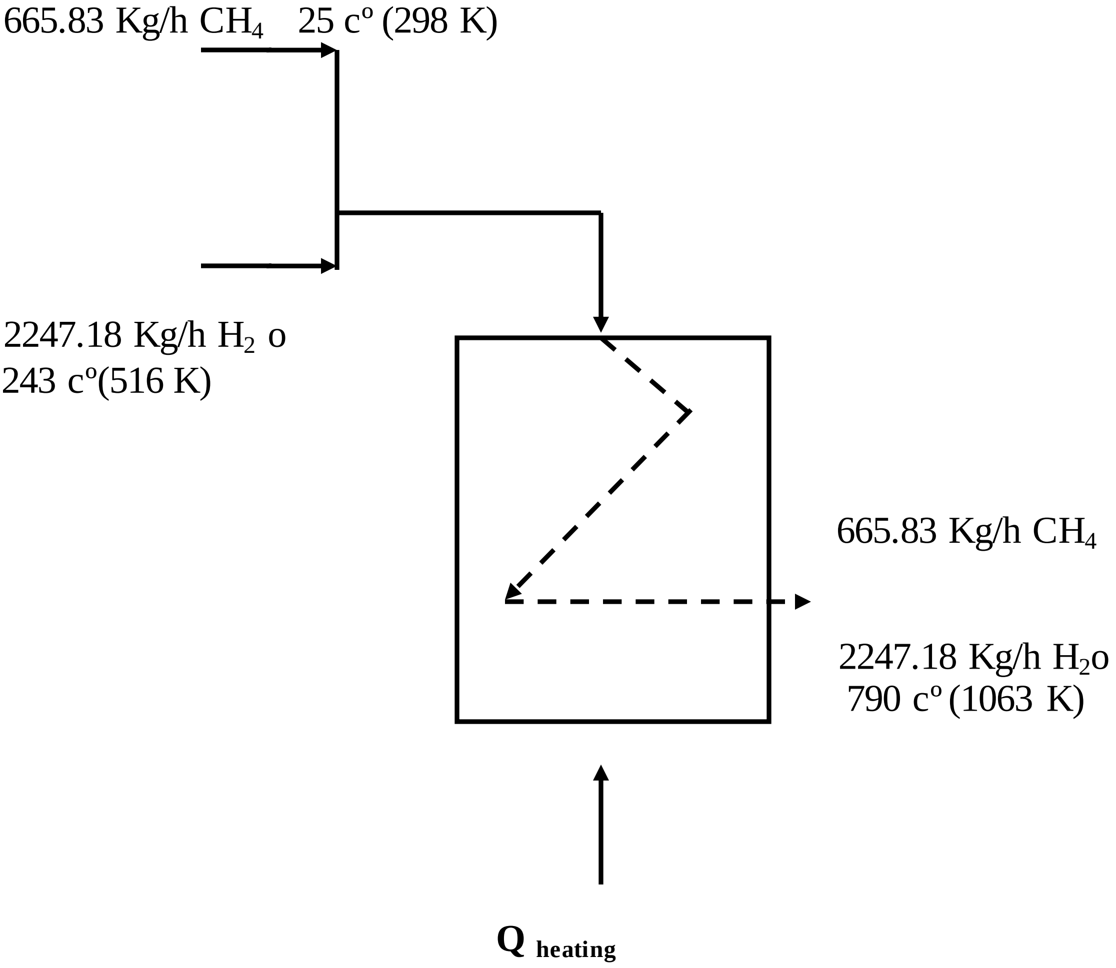
component	ΔHf_{298} **KJ/mol**
CH_4	**-74.86**
CO	**-110.62**
CO_2	**-393.77**
NH_3	**-45.72**
H_2O	**-242**

Energy balance furnace
Methane at 25 °C &35 Kg/cm^2 (35 bar)
saturated steam at (35bar)

(243 °C) from steam table.

The two feeds are heated to 790 °C .

(Reaction temperature of primary reformer)



Energy balance

$Q_{in} + Q_{heating} = Q_{out}$

Q_{in} = inlet enthalpy KJ/h
Q_{out} = outlet enthalpy KJ/h
$Q_{heating}$ = heat supplied KJ/h

$Q = n\ \Delta H$

$\Delta H = \int_{T\ ref}^{T} \mathbf{Cp\ dt}$

$Q = n \int_{Tref}^{T} \mathbf{Cp\ dt}$

n = Molar flow rate Kmol/h
Cp = gas heat capacity
dT = temperature difference

Q = Kmol/h *KJ/Kmol *k = KJ/h

$Cp = a+bT+cT^2+dT^3$

$\int_{T\ ref}^{T} Cp\ dt = \int_{T\ ref}^{T} (a+bT+CT^2+dT^3)\, dT$

$=[a(T-T_{ref})+b/2(T-T_{ref})+C/3(T-T_{ref})]$

T ref =25 c° = 29

$Q_{in} = n \int_{298}^{298} Cp\ dT + n \int_{298}^{516} Cp\ dT$

For CH_4

$\int_{298}^{516}$ **cp**dt = [32.243 (516 – 298) + 1.923 $*10^{-3}$ *(516^{2} – 298^2) /2 + 1.055$*10^{-5}$ *(516^3 -298^3)/3-3.596$*10^{-3*}$ (516^4 -298^4/4]

=7531 KJ/Kmol

n = M/Mw
= 2247.18/18 =124.8433 Kmol/h

$n= \int_{298}^{516} Cp\ dT = (124.8433)*(7531)$
= 940195 KJ/h

Q_{in} = 0 + 940195
Q_{in} = 940195 KJ/h

$Q_{out} = \sum ni \int_{298}^{1063} Cpi\, dT$

Component	**M Kg/h**	**n Kmol/h**	$\int_{298}^{1063} Cpi\, dT$	**n** $\int_{298}^{1063} Cpi\, dT$
CH_4	**665.83**	**41.6144**	**42959**	**1787713**
H_2o	**2247.18**	**124.8433**	**28657**	**3577634**

$Q_{out} = \sum ni \int_{298}^{1063} Cpi\, dT$

$= 5365347$ KJ/h

$940195 + Q_{heating} = 5365347$

$Q_{heating} = 4425152$ KJ/h

This heat is supplied by burning of natural gas

NHV =1005 Btu/ft^3

ρ= 0.0471 Ib/ft^3

NHV = Net heating value

NHV = 1005 Btu/ft^3 *1.055 KJ/Btu *ft^3/0.0471 Ib *2.2Ib/Kg

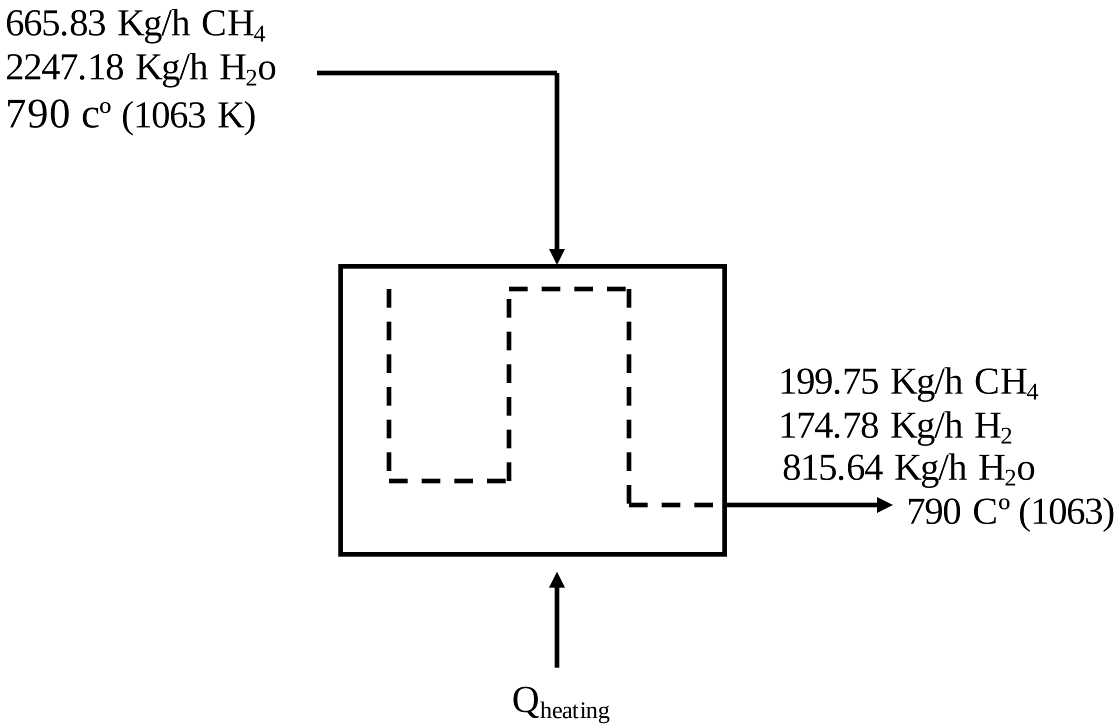
NHV = 49525 KJ/kg

$Q_{heating} = m$ (NHV)

$M = Q_{heating}$/NHV =4425152/49525

= 89.35 Kg/h

Energy balance on primary reformer :



Reaction

$CH_4 + H_2o \rightarrow Co + 3H_2$

Energy balance

$Q_{in} + Q_{heating} = Q_{out} + Q_{reaction}$
$T_{raf} = 25C^o$ (298K)

$$Q_{in} = \sum ni \int_{298}^{1063} Cpi\, dT$$

$$Q_{out} = \sum ni \int_{298}^{1063} Cpi\, dT$$
$$\sum ni \int_{298}^{1063} Cpi\, dT = 6138415 \text{ KJ/h}$$

$Q_{out} = 6138415$ KJ/h

$\Delta Hr_{298} = \sum \Delta Hf_{298} - \sum \Delta Hf_{298}$

$\Delta Hr_{298} = 0 + (-110.62) - (-242) - (-74.86)$

$= 206.24$ KJ/mol $= 206240$ KJ/Kmol

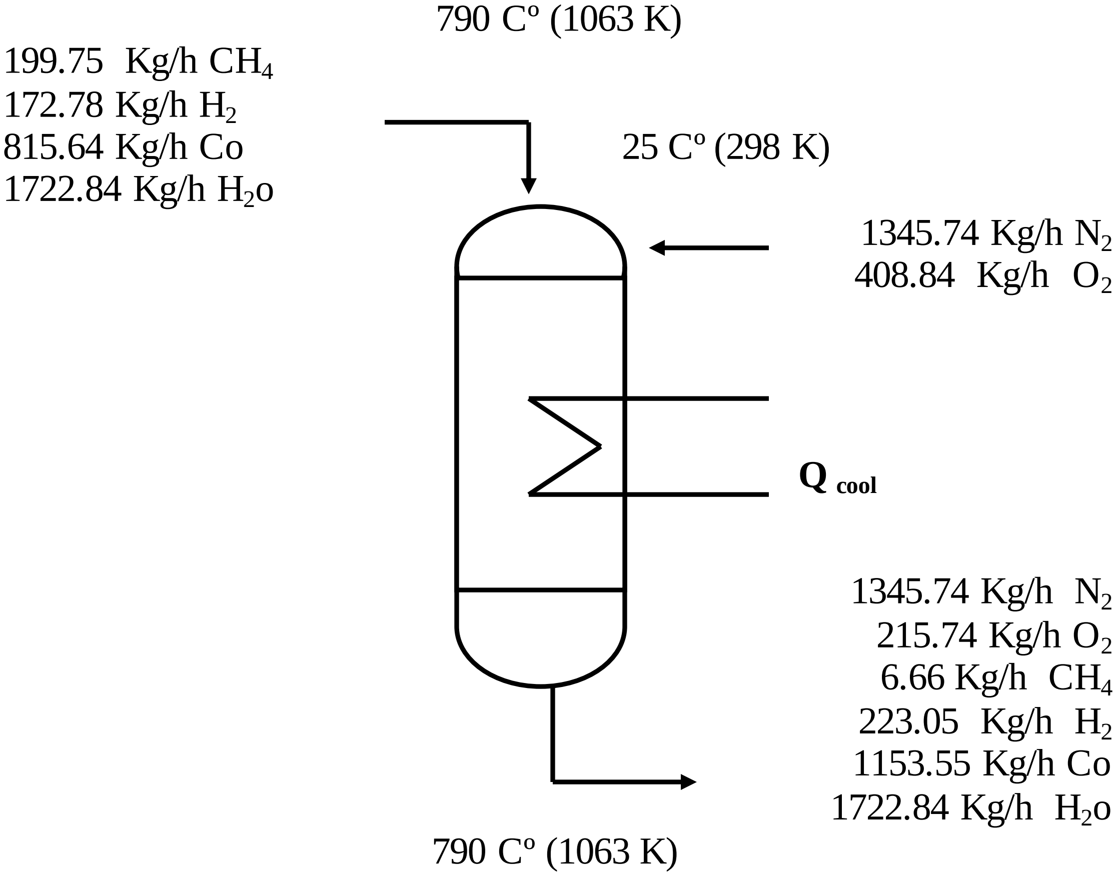
CH_4 reaction n= 29.1301 Kmol/h

$Q_{reaction} = n\,\Delta Hr$
$= (29.1301) * (206240) = 6007792$ KJ/h

$5365347 + Q_{heating} = 6138415 + 6007792$
$Q_{heating} = 6780860$ KJ/h

$Q_{heat} = m\,(NHV)$
m= Q/NHV = 6780860/49525 = 137 Kg/h

Energy balance on secondary reformer:



Reaction

$CH_4 + 1/2\,O_2 \rightarrow Co + 2H_2$

Energy balance

$Q_{in} = Q_{out} + Q_{reaction} + Q_{cool}$

T_{ref} = 25 C° = 298K

$Q_{in} = \sum ni \int_{298}^{298} \boldsymbol{Cpi\ dT} + \sum ni \int_{298}^{1063} \boldsymbol{Cpi\ dT}$

$\sum ni \int_{298}^{298} \boldsymbol{Cpi\ dT} = 0$

$\sum ni \int_{298}^{1063} \boldsymbol{Cpi\ dT} = 6138415$ KJ/h

Q_{in} = 0 + 6138415

= 6138415 KJ/h

$Q_{out} = \sum ni \int_{298}^{1063} \boldsymbol{Cpi\ dT}$

= 6259144 KJ/h

$\Delta Hr_{298} = \sum \Delta Hf_{298} - \sum \Delta Hf_{298}$

= 0 + (-110.62) – 0 – (-74.86)

= -35.76 KJ/mol CH_4

= -35760 KJ/Kmol CH_4

CH_4 reaction n = 12.0682 Kmol/h

$Q_{reaction}$ = n ΔHr

= (12.0682)*(-35760) = -431559 KJ/h

6138415 = 6259144 – 431559 + Q_{cool}

Q_{cool} = 310830 KJ/h
This heat is removed by cooling

Water inlet at 25 C° and 20 bar and exit saturated steam at 20 bar for steam tables

T_{sat} = 212 C°

λ= 1890.2 KJ/Kg

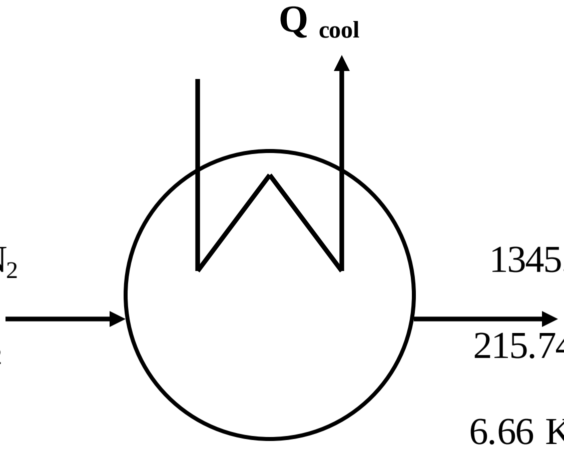
$Q_{cool} = m\ Cp\ \Delta T + m\ \lambda$

$310830 = m * 412 *(214 - 25) + 1890.2m$

m= 116 Kg/h H_2o

Energy balance on water heat boiler

The gases are cooled from 790 to 232 C°



1345.74 Kg/h N_2

215.74 Kg/h O_2

6.66 Kg/h CH_4
223.05 Kg/h H_2
1153.55 Kg/h Co
1722.84 Kg/h H_2o
790 C° (1063K)

1345.74 Kg/h N_2

215.74 Kg/h O_2

6.66 Kg/h CH_4
223.05 Kg/h H_2
1153.55 Kg/h Co
1722.84 Kg/h H_2o

Energy balance

$Q_{in} = Q_{out} + Q_{cool}$

$T_{ref} = 25$ C° $= 298$ K

$Q_{in} = \sum ni \int_{298}^{1063} Cpi\ dT$

$= 6259144$ KJ/h

$Q_{out} = \sum ni \int_{298}^{505} Cpi\ dT$

$= 1886185$ KJ/h

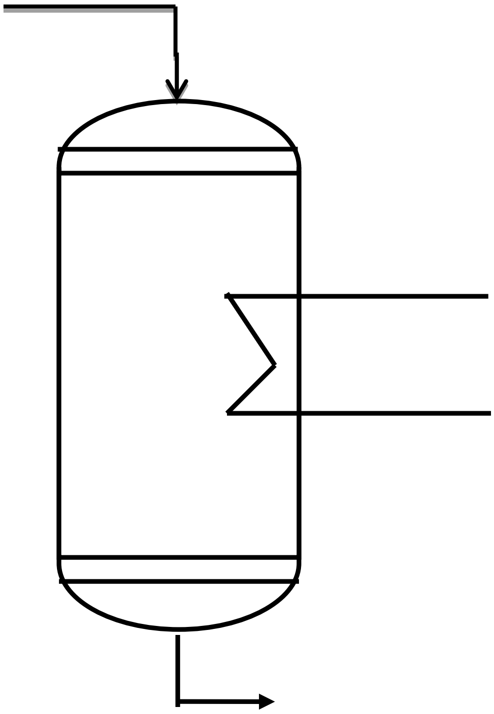
6259144 = 1886185 + Q_{cool}
Q_{cool} = 4372959 KJ/h
This heat is removal by cooling water inlet at 25 C° and 105 Kg/Cm^2 (103 bar) and exit
Saturated steam at 103 bar from steam tables

λ= 1294.1 KJ/h , T_{sat} =314 C°

$Q_{cool} = m\,Cp\,\Delta T + m\,\lambda$
4372959 = m * 4.2 *(314-25) +1`294.1m
m= 1744 Kg/h, = 0.48 Kg/h H_2o

Energy balance on shift converter:

1345.74 Kg/h N_2
215.74 Kg/h O_2
6.66 Kg/h CH_4
223.05 kg/h H_2
1153.55 Kg/h Co
1722.84 Kg/h H_2o
232 C° (505K)



1345.74 Kg/h N_2

215.74 Kg/h O_2
6.66 Kg/h CH_4
301.33 Kg/h H_2
57.68 Kg/h Co
1722.09 kg/h Co_2
1018.34 Kg/h H_2o

Reaction

$Co + H_2O \rightarrow Co_2 + H_2$

Energy balance

$Q_{in} = Q_{out} + Q_{reaction} + Q_{cool}$
$T_{ref} = 25\ C^o = 298\ K$

$Q_{in} = \sum ni \int_{298}^{505} Cpi\ dT$

$= 1886185\ KJ/h$

$Q_{out} = \sum ni \int_{298}^{505} Cpi\ dT$

$= 1894616\ KJ/h$

$\Delta Hr_{298} = 0 + (-393.77) - (-242) - (-110.62)$

$= -41.15\ KJ/mol$

$= -41150\ KJ/Kmol$

Co reaction n = 39.1384 Kmol/h

$Q_{reaction} = n\ \Delta Hr$

$= (39.1384) * (-41150)$

$= -1610545\ KJ/h$

$1886185 = 1895616 - 1610545 + Q_{cool}$

$Q_{cool} = 1601114\ KJ/h$

This heat is removed by cooling water at 25 C° and 10 bar and exit saturated steam at 10 bar, $T_{sat} = 180\ C^o$

$\lambda = 2113.2\ KJ/h$

$Q_{cool} = m\ Cp\ \Delta T + m\ \lambda$

1601114 = m*2.4 *(180 -25) +2113.2 m

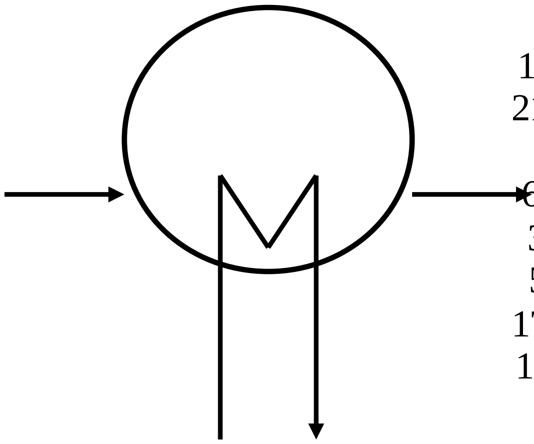
M= 579 Kg/h H_2O

<u>Energy balance on waste heat boiler:</u>

The gases are cooled from 232 to 150 C°

1345.74 Kg/h H_2
215.74 Kg/h O_2

6.66 Kg/h CH_4
301.33 kg/h H_2
57.63 Kg/h Co
1722.04 Kg/h Co_2
1018.74 Kg/h H_2O
232 C° (505K)



1345.74 Kg/h H_2
215.74 Kg/h O_2

6.66 Kg/h CH_4
301.33 kg/h H_2
57.63 Kg/h Co
1722.04 Kg/h Co_2
1018.74 Kg/h H_2O

Energy balance

$Q_{in} = Q_{out} + Q_{cool}$

$T_{ref} = 25$ C° = 298 K

$Q_{in} = \sum ni \int_{298}^{505} Cpi\, dT$

=1895616 KJ/h

$Q_{out} = \sum ni \int_{298}^{423} Cpi\, dT$

$Q_{out} = 1080501$ KJ/h
$1895616 = 1080501 + Q_{cool}$

Q_{cool} = 815115 KJ/h

This heat is removed by cooling water at 25 C° and 10 bar and exit saturated steam at

10 bar

T_{sat} = 180 C°

λ= 2113.2 KJ/h

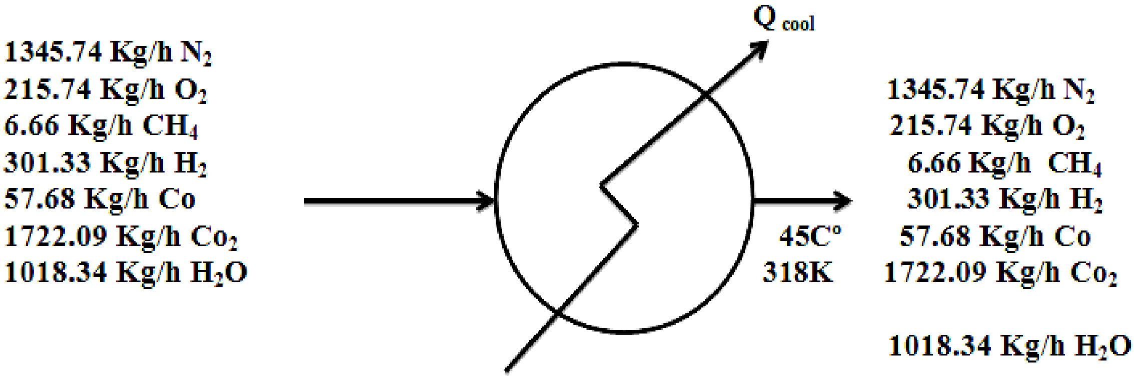
$Q_{cool} = m\ Cp\ \Delta T + m\ \lambda$

$815115 = m*4.2 *(180-25) + 2113.2\ m$

m= 295 Kg/h

Energy balance on partial condenser:

The gases are cooled from 150 to 45 C° where water is condenser



Energy balance

$Q_{in} = Q_{out} + Q_{cool}$

T_{ref} =25 C° = 298 K

$Q_{in} = \sum ni \int_{298}^{423} Cpi\ dT + m\ \lambda$

$\sum ni \int_{298}^{423} Cpi\ dT = 1080501$ KJ/h

$M\lambda = (1018.34)*(2256.9)$

$= 2298292$ KJ/h

$Q_{in} = 1080501 + 2298292$

$= 3378793$ KJ/h

$$Q_{out} = \sum ni \int_{298}^{318} Cpi\, dT + m\, Cp\, \Delta T$$

$$\sum ni \int_{298}^{318} Cpi\, dT = 131514 \text{ KJ/h}$$

$m\, Cp\, \Delta T = (1018.34)*(4.20)*(318-298)$
$= 85541$ KJ/h

$Q_{out} = 131514 + 85541$
$= 217055$ KJ/h

$3378793 = 217055 + Q_{cool}$
$Q_{cool} = 3161738$ KJ/h
$Q_{cool} = m\, Cp\, \Delta T$

For cooling water

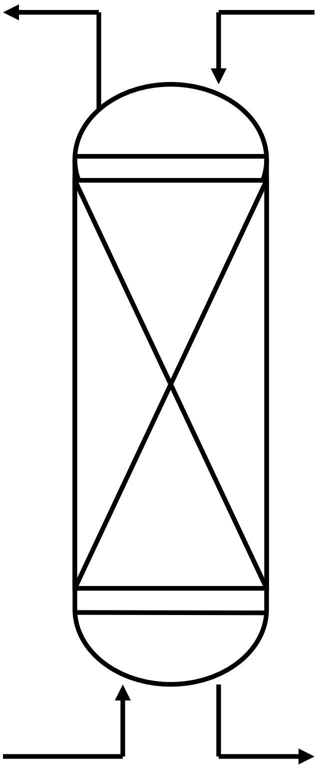
$T_{in} = 25$ C°
$T_{out} = 40$ C°
$m = Q_{cool} / Cp\, \Delta T$
$m = 3161738 / 4.2 *(40.25)$
$m = 50186$ Kg/h
$= 13.9$ Kg/s

Energy balance on absorber

5968.61 Kg/h MEA
23874.44 Kg/h H_2
72.02 Kg/h $(C_2H_5ONH_3)_2CO_3$

1345.74 Kg/h N_2
215.74 Kg/h O_2
6.66kg/hr CH_4
301.33 Kg/h H_2
57.68 Kg/h Co
17.22 Kg/h Co_2

P = 30 atm



1345.74 Kg/h N_2
215.74 Kg/h O_2
6.66 Kg/h CH_4
301.33 Kg/h H_2
57.68 Kg/h Co
17.22 Kg/h Co_2
45 C^o (318K)

23176.99 Kg/h H_2O
1241.47 Kg/h MEA

7201.47 Kg/h $(C_2H_5ONH_3)$ Co_3

Reaction
$C_2H_7NO + H_2O + Co_2 \rightarrow (C_2H_5ONH_3)_2 Co_3$

Component	**Cp $_L$ KJ/Kg .k**
H_2O	**4.20**
M E A	**2.67**
$(C_2H_5ONH_3)_2$ Co_3	**2.93**

Energy balance

$Q_{in} = Q_{out} + Q_{reaction}$

$T_{ref} = 25\ C^o = 298\ K$

$Q_{in} = \sum ni \int_{298}^{318} Cpi\ dT + \sum mi\ Cp\ i\ \Delta T$

$\sum ni \int_{298}^{318} Cpi\ dT = 131514$ KJ/h

$\sum mi\ Cp\ i\ \Delta T = \{(23874.44*4.20) + (5968.61*2.67) +(72.02*2.93)\}*\{318-298\}$

$= 2328397$ KJ/h

$Q_{in} = 131514 + 2328397 = 2459911$ KJ/h

By trial and error assume exit temperature = 70 C° (333 K)

Mean heat Capacity of gases between 25.7

Component	**Cp KJ/Kg .k**
N_2	**1.04**
O_2	**0.92**
CH_4	**2.31**
H_2	**14.39**
Co	**1.04**
Co_2	**0.86**

$Q_{out} = \sum mi\ Cp\ i\ \Delta T + \sum mi\ Cp\ i\ \Delta T$

$=[(1345.74*1.04)+(215.74*0.92)+(6.66*2.31) +(301.33*14.39)+(57.68*1.04)+(17.22*0.86)]$

$*[T-298]+[(23176.99*4.20)+(1241.47*2.67)+(7201.47*2093)]*[T-298]$

$Q_{out} = 6024.3701\ (T-298)+121758.39\ (T-298)$

Q_{out} = 127782.76 (T-298)KJ/h

ΔHr = -825 BTU/Ib Co_2

ΔHr =(-825 BTU/Ib CO_2) * (1.055 KJ/BTU) * (2.2 Ib/Kg) *(44 Kg/Kmol)

ΔHr = -84252 KJ/Kmol Co_2 reaction

Co_2 reacted n= 38.7470 Kmol/h

$Q_{reaction}$ = n ΔHr

= (38.7470)*(-84252) = -3264512 KJ/h

2459911 = 127782.79 (T-298) – 3264512

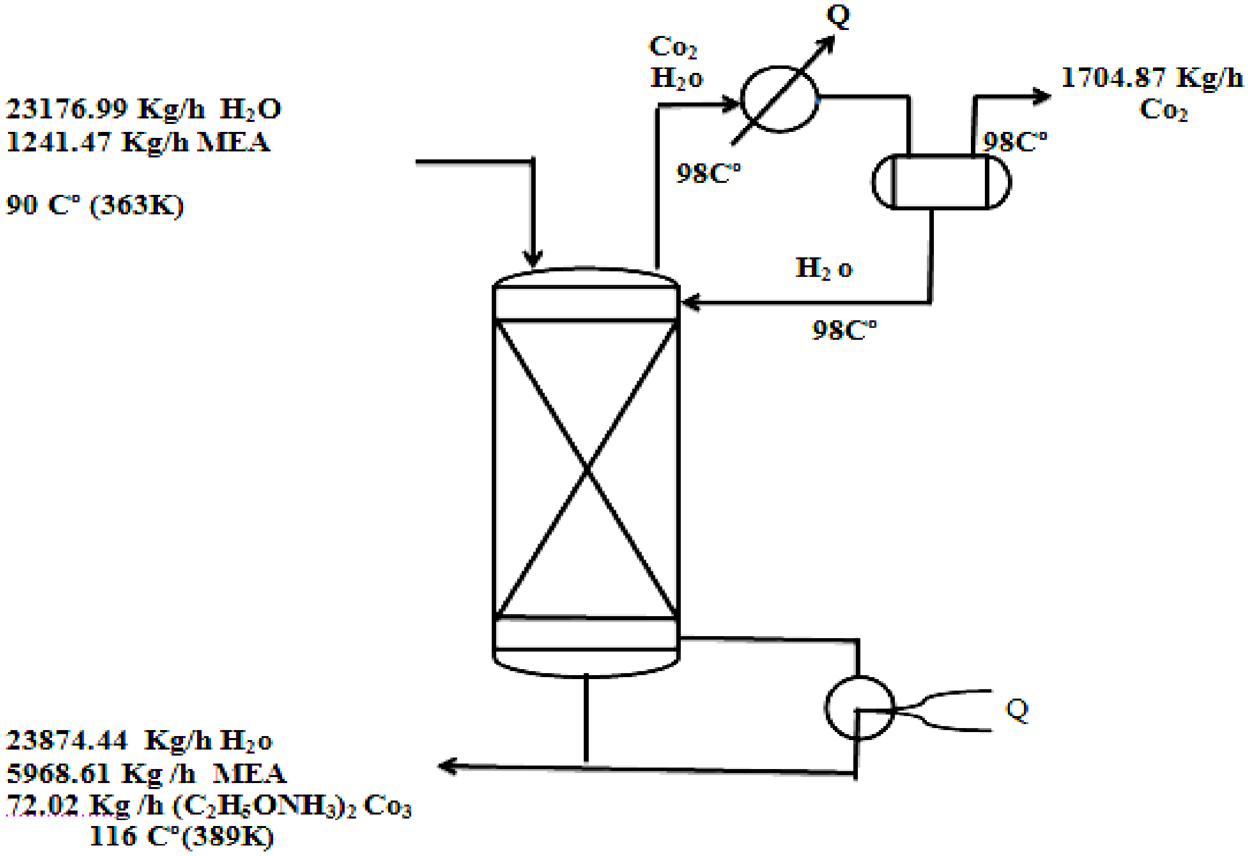
5724423= 127776.6 *(T-298)

T = 342.8 K = 69.8 C°

T assume = 70 C°

Energy balance on stripper operating conditions:

Top	Bottom
T=208 F°	T= 240 F°
= 98 C°	=116 C°
P =20 psia	P= 24 psia
= 137.3 Kpa	=165.4Kpa



Reaction

$(C_2H_5ONH_3)_2\ Co_3 \rightarrow C_2H_7NO + H2O + Co_2$

Energy balance

$Q_{in} + Q_{req} = Q_{out} + Q$

$T_{ref} = 25\ C^o$

$Q_{in} = \sum mi\ Cp\ i\ \Delta T$

$=[(23176.99*4.20) + (1241.47*2.67) + (7201.47*2.93)]*[363-298]$

$Q_{in} = 7914295$ KJ/h

$Q_{out} = n\int_{298}^{371} Cp\, dT + \sum mi\, Cp\, i\, \Delta T$

$n\int_{298}^{371} Cp\, dT = 112010$ KJ/h

$\sum mi\ Cp\ i\ \Delta T = [(23874.44*4.2)+(5968.61*2.67)+(72.02+2.93)]*[389-298]$

$=10594207$ KJ/h

$Q_{out} = 112010 + 10594207$

$=10706212$ KJ/h

Calculation of Q_c

$Y_w = P_w / P$

Y_w =mol fraction of water

P_w =partial pressure of water

P = total pressure Kpa

Assume exit Co_2 is saturated with water vapor at to top conditions

T = 98 C°

P = 137.8 Kpa

At saturation

Partial pressure of water vapor = vapor pressure of water vapor

$P_w = P^*_w$

$Y_w = P^*_w/P$

Vapour pressure of water at 98 C°

$P^*_w = 94.3$ Kpa

$Y_w = 94.3/137.8 = 0.6843$

Y_w = Kmol H_2o/total Kmol

H_2o = w Kmol/h

Co_2 = 1704.87/44= 38.7470 Kmol/h

W= 83.99Kmol H_2o *18

= 1511.76 Kg/h

$Q_c = m\ \lambda$

At 98 C° λ=2262.2 KJ/Kg

Q_c = (1511.76)*(2262.2) =3419904 KJ/h

7914295 + Q_{reb} = 1070 6217 +3419904

Q_{reb}= 6211826 KJ/h

$Q_c = m\ Cp\ \Delta T$

For cooling water

T_{in} = 25 C°
T_{out} = 40 C°

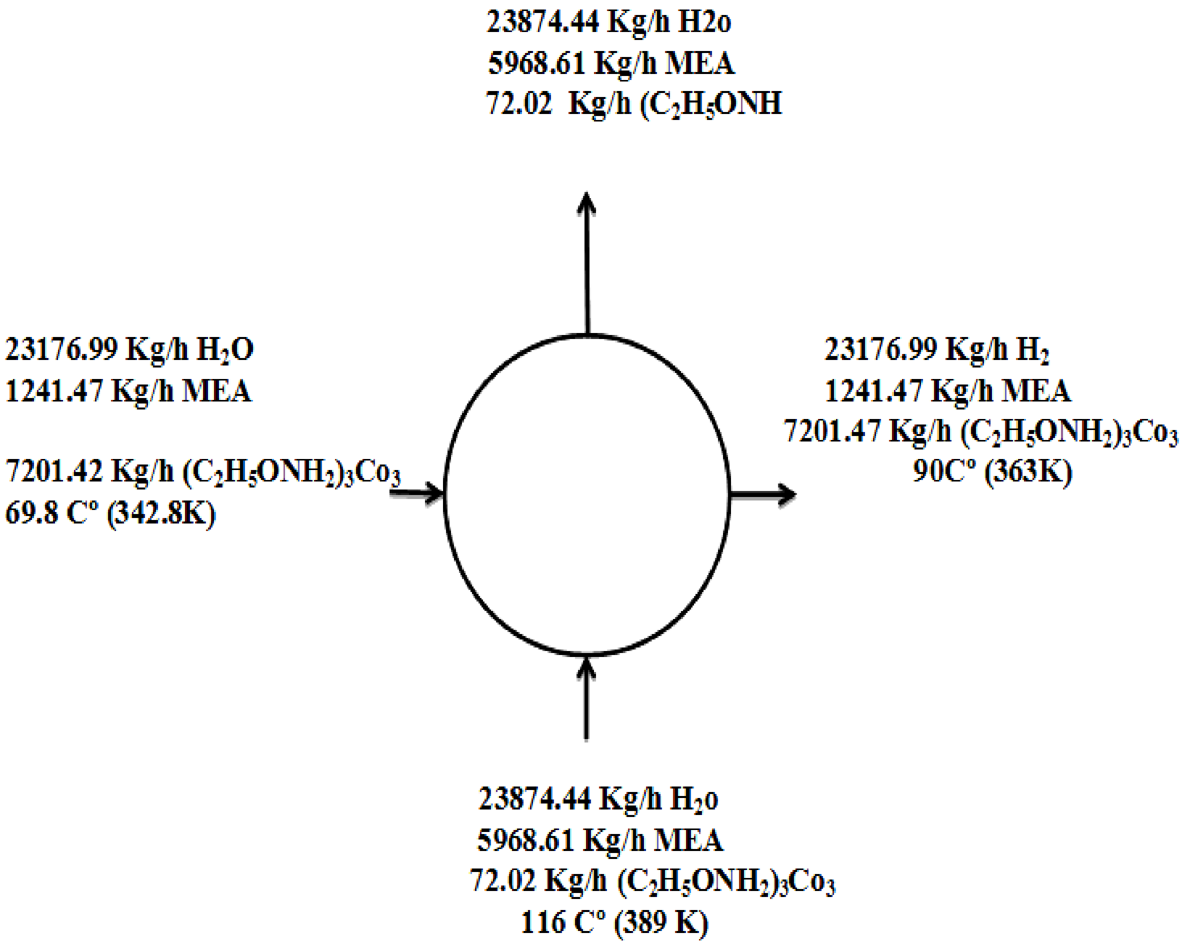
m = $Q_{c/}$ Cp ΔT = 3419904/4.2*(40.25)
= 54284 Kg/h
= 1511 Kg/s

$Q_{reb} = m\ \lambda$

Using saturated steam at 180 C°
λ= 2113.2 KJ/Kg
m= Q_{reb}/λ = 6211826/2113.2

= 2940 Kg/h
= 0.82 Kg/s

Energy balance on heat exchanger bottom of Absorber



Energy balance

$Q_{in} = Q_{out}$

$T_{ref} = 25C^o = 298K$

$Q_{in} = \sum mi\ Cp\ i\ \Delta T + \sum mi\ Cp\ i\ \Delta T$

$\sum mi\ Cpi\ \Delta T = 5724423$ KJ/h

$\sum mi\ Cpi\ \Delta T = 10594207$ KJ/h

$Q_{in} = 5724423 + 10594207$

= 16318632 KJ/h

$Q_{out} = \sum mi\ Cpi\ \Delta T + \sum mi\ Cpi\ \Delta T$

$\sum mi\ Cpi\ \Delta T = 7914295$ KJ/h

$\sum mi\ Cpi\ \Delta T = [(2384.44*4.2)+(5968.61*2.67)*(72.02*2.93)]*[T-298]$

$= 116419.86\ (T-298)$ KJ /h

$16318632 = 7914295 + 116419.86(T-298)$

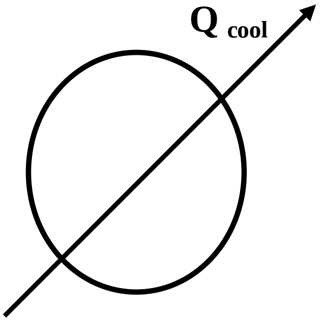
$8404337 = 116419.86(T-298)$

T= 370 K

=97C° to absorber

<u>Energy balance on cooler :</u>

23874.44 Kg/h H_2
5968.61 Kg/h MEA
72.02 Kg/h $(C_2H_5ONH_3)_2$ Co_3
45 C°



23874.44 Kg/h H_2o
5968.61 Kg/h MEA
72.02 Kg/h $(C_2H_5ONH_3)_2Co_3$
97 C°
Energy balance

$Q_{in} = Q_{out} + Q_{cool}$

T_{ref} =25 C° (298)

$Q_{in} = \sum mi\ Cpi\ \Delta T$

$= 8404337$ KJ /h

$Q_{out} = \sum mi\ Cpi\ \Delta T$

$= 2328397$ KJ/h

$8404337 = 2328397 + Q_{cool}$

$Q_{cool} = 6075940$ KJ/h

$Q_{cool} = m\ Cp\ \Delta T$

For cooling water

$T_{in} = 25\ C^o$

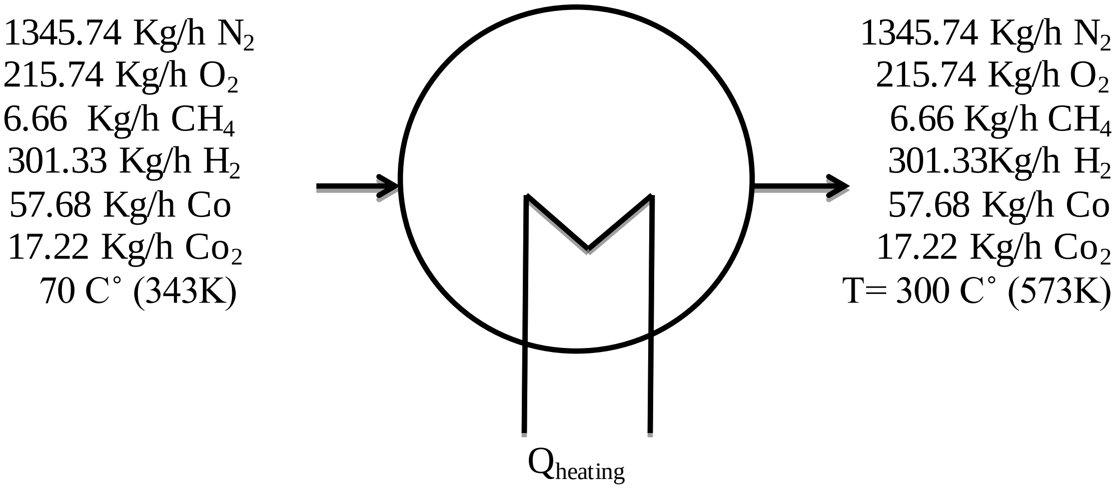
$T_{out} = 40\ C^o$

$m = Q_{cool} / Cp\ \Delta T$

$=6075940/4.2\ *(40.25) = 96443$ Kg/h

$=26.8$ Kg/s

Energy balance on heat exchanger



Energy balance

$Q_{in} + Q_{heating} = Q_{out}$

$T_{ref} = 25C^o = 298k$

$Q_{in} = \sum ni \int_{298}^{343} Cp\, dt$

$Q_{in} = 5724423$ KJ/h

$Q_{out} = 38480843$ KJ/h

$5724433 + Q_{heat} = 38480843$

$Q_{heat} = 32756420$ KJ/h

$Q_{heat} = m\lambda$

Using saturated steam at 350 C°

$\lambda = 895.6$ KJ/kg from steam table

$m = Q_{heat}/\lambda$

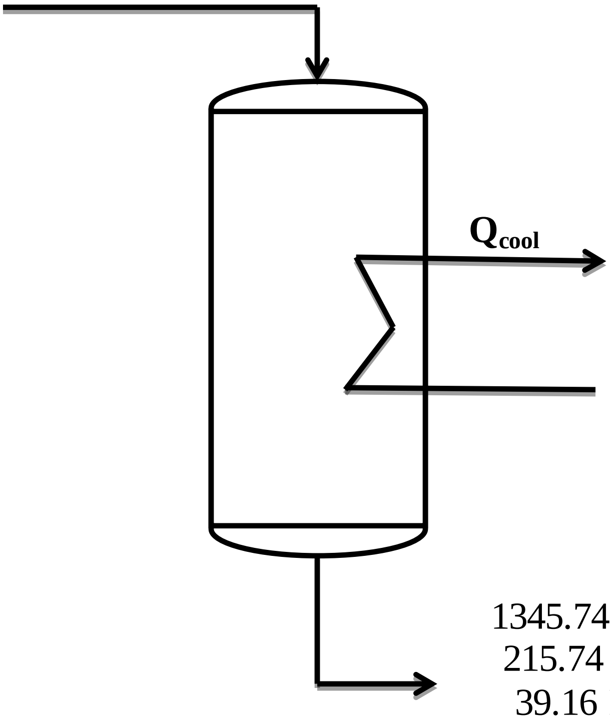
$m = 32756420/895.6 = 36575$ Kmol/h

$= 10.16$ Kg/s

Energy balance on methanomator

1345.74 Kg/h N_2
215.74 Kg/h O_2
6.66 Kg/h CH_4
301.33 Kg/h H_2
52.68 Kg/h Co
17.22 Kg/h Co_2

300C°(573K)



1345.74 Kg/h N_2
215.74 Kg/h O_2
39.16 Kg/h CH_4
288.37 Kg/h H_2
11.54 Kg/h Co
0.44 Kg/h Co_2
43.47 Kg/h H_2o

Reaction

$Co + 3H_2 \rightarrow CH_4 + H_2O$

$Co_2 + 4H_2 \rightarrow CH_4 + 2H_2O$

Energy balance

$Q_{in} = Q_{out} + Q_{cool} + Q_{reaction}$

$T_{ref} = 25C^o = 298K$

$Q_{in} = \sum ni \int_{298}^{573} Cpi\, dT$

$Q_{in} = 38480843$ KJ/h

$Q_{out} = \sum ni \int_{298}^{573} Cpi\, dT$

$= 37711226$ KJ/h

$\Delta Hr_{298} = (-242)+(-74.86) -0- (-110.62)$

$= -206.24$ KJ/mol

$= -206240$ KJ/Kmol

Co reacted ni = 106479 Kmol/h

$n\Delta Hr_1 = (1.6479) *(-206240) = -339863$ KJ/h

$\Delta Hr_2 = 2*(-242)+(-74.86)-0-(-393.77)$

$\Delta Hr_2 = -165.09$ KJ/mol

$= -165090$ KJ/Kmol

CO_2 reacted $n_2 = 0.3836$ Kmol/h

$n_2 \Delta Hr_2 = 0.3836 *(-165090)$

$= - 63329$ KJ/h

$Q_{reaction} = -339863 – 63329$

$= -403192$ KJ/h

$38480843 = 37711226 – 403192 + Q_{cool}$

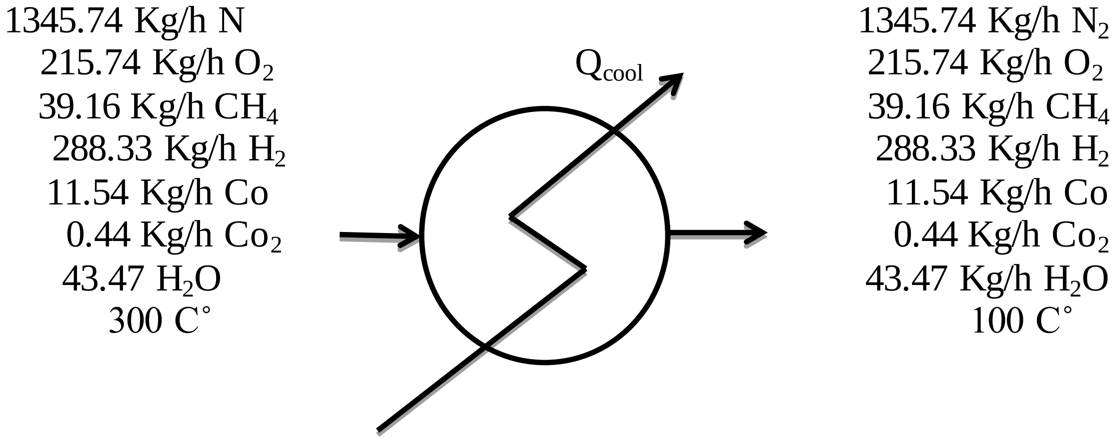
Q_{cool} = 1172809 KJ/h

$Q = m\,Cp\,\Delta T + m\lambda$

1172809 = m*(100-25) + 2256.9 m

M= 456 Kg/h H_2O

Energy balance on waste e that boiler



Energy balance

$Q_{in} = Q_{out} + Q_{cool}$

T_{ref} = 25 C = 298 K

$Q_i = \Sigma\, ni \int_{298}^{573} cpidt$

=37711226 KJ/h

$Q_{out} = \Sigma ni \int_{298}^{573} cpidt$

=10071982 KJ/h

37711226 = 10079182 + Q_{cool}

Q_{cool}= 27632044 KJ/h

Q_{cool} =mcp$\Delta T + m\lambda$

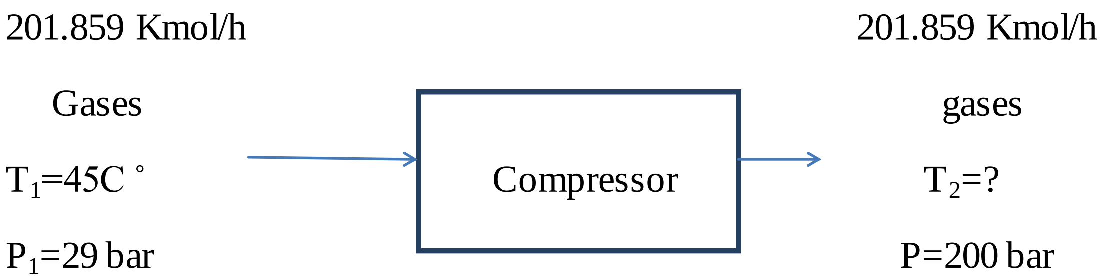
27632044= m+ 4.2* (100-25)+ 2256.7m

M=10744 Kg/h H_2O

=2.98 Kg/s

Energy balance on Compressor t

The gases are compressed from 29 to 200 bar



$(T_{2/}T_1)=(P_{2/}P_1)^m$

m = (ɤ-1)/epɤ

ɤ=Cp/Cv≅ 1.4

E_p =efficiency of compressor = 80%

M=(1.4-1)/(0.8*1.4)= 0.35

$(T_2/318)=(200/29)^{0.35}$

T_2=625 K°

=352 C°

Energy balance on NH_3 convertion

5719.37 Kg/h N_2

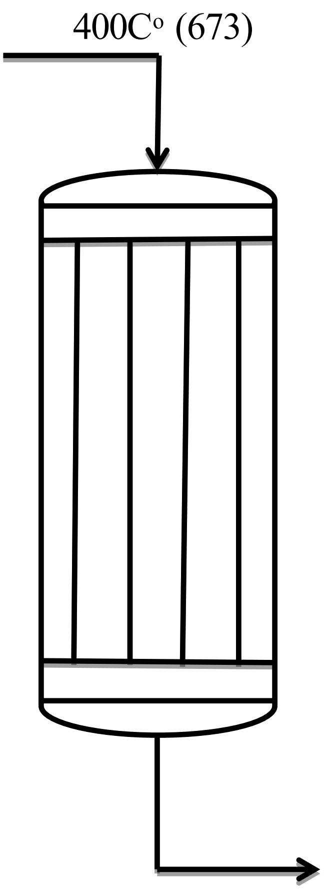
4890.18 Kg/h O_2

887.66 Kg/h CH_4

1225.59 Kg/h H_2

261.48 Kg/h Co

9.92 Kg/h Co_2



457.49 Kg/h N_2

4890.18 Kg/h O_2

887.66 Kg/h CH_4

980.47 Kg/h H_2

261.48 Kg/h Co

9.97 Kg/h Co_2

1389 Kg/h NH_3

Reaction

$N_2+3H_2\leftrightarrow 2NH_3$

Energy balance

$Q_{in} = Q_{out}+Q_{reaction}+Q_{cool}$

$T_{ref}=25C^{\circ}=298\ K^{\circ}$

$Q_{in}=\Sigma\ ni\int_{298}^{673} cpidt$

$Q_{in}=11866100\ \ KJ/h$

$Q_{out}=\Sigma ni\int_{298}^{698} cpidt$

$Q_{out}=12783745\ \ KJ/h$

$\Delta Hr=(-45.72)-0-0$

$=-45.72\ KJ/mol$

$=-45720\ KJ/mol$

N_2 reacted n=40.8530 Kmol/h

$Q_{reaction}=(40.8530)\text{-}(-45720)$

$=-1867799\ \ KJ/h$

$11866100=12783745-1867799+Qcool$

$Q_{cool}=950154\ \ KJ/h$

$Q_{cool}=m\ Cp\ \Delta T+m\lambda$

$950154=m*\ 4.2*(100-25)+2256.9\ m$

M =369.4 Kg/h H_2O

3.1 Introduction to *Equipment design*

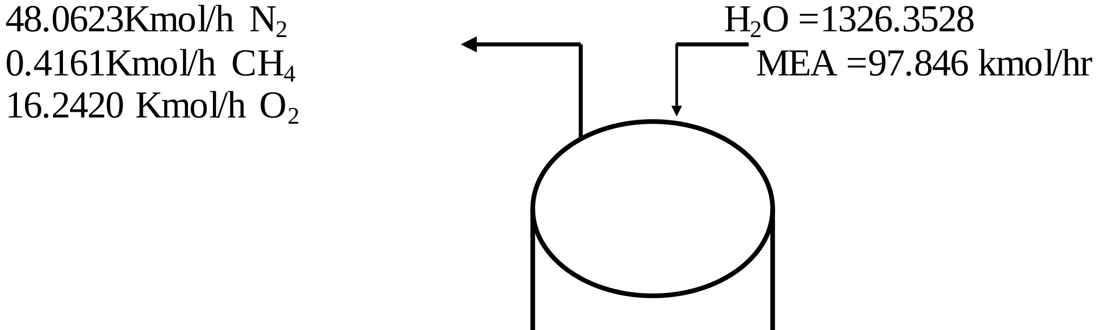
Chemical Industries involve problems in process design, unit operations, equipment design and overall plant design. In design of a chemical plant these problems cannot be segregated. However, these problems may be advantageously segregated for study and development because of different principles involved in it. The course goes deeper into the various aspects of mechanical design in the chemical process plant. It stresses upon the design and analysis of the basic process equipment viz. vessels, distillation column, absorption column, heat exchanger, driers and evaporators etc.

The course emphasizes on the development of design skills among the students to take design related decisions. Whatever be the earlier conception, today a chemical engineer is expected to be able to make complete design of a piece of chemical equipment. The course will be very useful to undergraduate students and practitioners. A number of problems will be solved to illustrate the concepts clearly.

Chapter Three

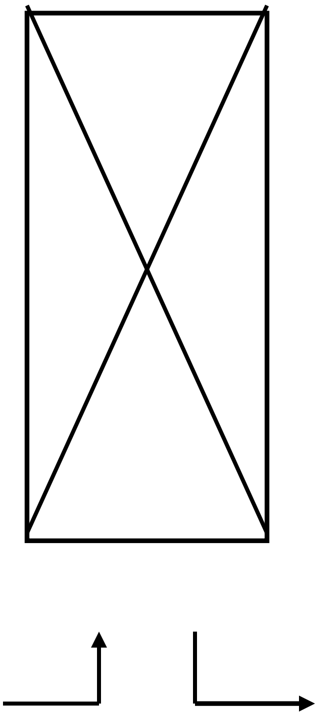
Equipment design

3.2 Absorber design:



150.6651Kmol/h H_2
2.0599 Kmol/h CO

CO_2=0.3914 Kmol/h



48.0623Kmol/hr N_2

6.7420Kmol/hr O_2

0.4161Kmol/hr CH_4

150.6651Kmol/hr H_2
39.1384Kmol/hr CO_2

2.0599Kmol/hr CO

H_2O=1287.6108 Kmol/hr

MEA=20.3520 Kmol/hr

<u>Input :</u> let A= CO_2 = 39.1384 Kmol/hr

B=207.9 Kmol/hr

y_1 =A/B =39.1384/207.9 =0.188

<u>0utput</u> :A= CO_2 =0.3914 Kmol/hr

B=207.9 Kmol/hr

y_2= A/B =0.3914 /207.9 =0.0018

Gs= B=207.9 Kmol/hr

Gs y_1 +Ls x_2 =Gs y_2 +Ls x_1

x_2 = 0
207.9*0.188 =207.9*0.0018 + x_1 *1347.1
x_1 =0.03
from (x_1, y_1) & (x_2 , y_2)
m = 4.3

Φ = m Gs/ Ls =4.3*207.9/1347.1 =0.66
y_1/y_2 = 0.188/0.0018 =104.4
from fig.(3.1) (appendix A)

NOG =10
HTU = 0.5 m (table3.1)(appendix B)
Where :
HTU :the height of a transfer unit

Z = 0.5*10 =5m
Gs = 207.9 kmol /hr =0.057 kmol/s
Ls = 1347.1 kmol /hr =0.374 kmol/s
Where :
Gs: is the mole rate of the inlet gas , kmol/s
Ls: is the mole rate of the liquid solvent, kmol/s
ρ_L = 998 kg/m^3
μ = 10^{-3} NS/m^2
F_ρ = 170m^{-1}
ρ_v = 1.094 kg/m^3

F_{LV}= Ls /Gs $\sqrt{(\rho_v / \rho_L)}$
Gs = 1.653 kg/s
Ls = 0.618 kg/s

0.618/1.653$\sqrt{(1.094/998)}$ = 0.012
Design for pressure drop of 20mm H_2O
From fig.(3.2) (appendix A)
K_4 = 1.51
At flooding K_4 =1.64
Percent flooding =$\sqrt{(1.51/1.64)}$*100 =95%

V_w^* = $\left[K_4\, \rho_v(\rho_L - \rho_v)/13.1\, F_p(\mu / \rho_L) \right]$
Where:
V_w^*: gas mass flow-rate per unit column cross-sectional area, kg/m^{2}s
F_ρ : packing factor, characteristic of the size and type of packing, see Table (3.1), m^{-1}.
μ: liquid viscosity, Ns/m^2
ρ_L& ρ_v: liquid and vapor densities, kg/m^3

=1.7
Col. area required =1.013 /1.7 =0.59

Diameter $=\surd 4/\Pi *0.59) =0.87$m

Column area= $\Pi/4\ d^2 = \Pi/4(0.87)^2 =0.56$

Packing size to column diameter ratio $=0.87/(38*10^{-3}) =22.8$

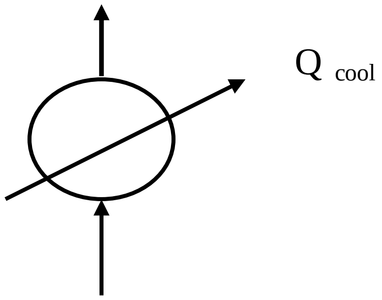
3.3 Heat exchanger design:

45°C (318k)

23874.44 kg /hr H_2O

5968.61 kg /hr MEA

72.02 kg /hr$(C_2H_5NH_3)_2CO_3$



23874.44 kg /hr H_2O

5968.61 kg /hr MEA

72.02 kg /hr$(C_2H_5NH_3)_2CO_3$

97°C

Q_{cool} = 6075940 kJ/hr

t_{in} = 25°C

t_{out} =40°C

Q_{in} =8404337 kJ/hr

Q_{out} =2328397 kJ/hr

M =96443 kg /hr

Mean tube temperature $t_m = t_1 + t_2 / 2$

$t_m = 97 +45 / 2 = 71$°C

physical properties at 71°C of 15% MEA

sp. gr=0.986

ρ =986 kg /m^3

$\mu =4.9 *10^{-4}$ kg/m .s

cp mix = cp /m ΔT

cp mix = 6075940 /29915.07 (97-45) =3906 J/kg .c

=3.906 kJ/ kg .c

Thermal conductivity :

H_2O $k_1=156.2*10^{-5}$ cal /cm .s . °C

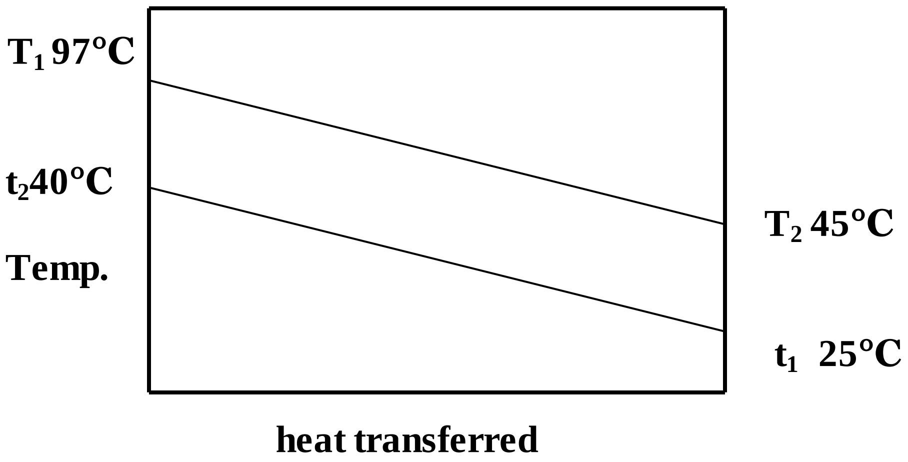
=0.654w / m. °C

MEA k_2=0.297 w/m. °C

K mix $=k_1x_1 +k_2x_2$

=0.654 *0.85 +0.277*0.15

=0.597 w/m. °C



ΔT_{1m} =(97 - 40) - (45-25) / ln (97- 40)/(45-25)
=35.3°C
R = $(T_1 - T_2)/(t_2 - t_1)$=97-45/ 40-25
=3.46
Use 0ne shell pass &two tube passes
S =$(t_2 - t_1)/(T_1 - t_1)$ = 40 -25 /97-25 =0.21
R: is equal to the shell-side fluid flow-rate times the fluid mean specific heat; divided
by the tube-side fluid flow-rate times the tube-side fluid specific heat.
S: is a measure of the temperature efficiency of the exchanger.

From fig. (3.3) ….. Ft = 0.85
Where : Ft = the temperature correction factor.
ΔT_m = true temperature difference, the mean temperature difference

$\Delta T_m = \Delta T_{1m} * Ft$
ΔT_m = 0.85 *35.3 = 30°C
U =860 $w/m^2.c$

Heat load = 96443/ 3600 *3.906(97- 45)
=5441.3 k w
A=5441.3*10^3 / (30*860)
=210 m^2
Choose 20mm 0.d ,16mm i.d ,4.88m long
Allowing for tube –sheet thickness ,take L=4.83m
Area of one tube =4.83 *20*10^{-3} Π =0.303 m^2
Number of tube =210 /0.303=693
As the shell –side fluid is relatively clean use 1.25 trangular pitch
Tube bundle diameter D_b =20 $(693/0.249)^{1/2.207}$

=727 mm
Use a split-ring floating head type from fig.(3.4)(appendix A)

bundle diametrical clearance = 69 mm
Shell diameter D_s =727+69 =796mm

Tube –side coefficient :

Tube –cross –sectional area $=\Pi/4*16^2 = 201mm^2$
Tube per pass =693/2 =347

Total flow area $=347*201*10^{-6} = 0.069\ m^2$
Cooling flow =5441.3 /3.906 (40-25)
=5441.3/585.9 =92.8

Linear velocity =1344/995=1.35 m /s
Mean mass velocity $=92.8 / 0.069 = 1344\ kg/s.m^2$

The coefficient can also calculated using equ. 12.15 ,this is done to illustrate use of this method .
$h_i\ d_i / k_f = j_f\ Re\ Pr^{\ 0.33}$
k (thermal conductivity)=0.59 w/m .c
$Re = \rho\ u\ d/\mu = 993.6 * 1.35 * 16 *10^{-3} / 7.53*10^{-4} = 28501$
$Pr = cp\ \mu / k_f = 3.906*10^3*7.53 *10^{-4} = 4.98$

$L/d_i = 4.83*10^3 /16 = 302$
From fig (3.6) $j_h = 3.5*10^{-3}$

$hi = 0.59/16*10^{-3} *3.5*10^{-3}*28501 *4.98^{0.33}$
$=6248\ w/m^2 .c$

Shell – side coefficient:

Choose baffle spacing = Ds/5 =796/5 =159.2mm
Tube pitch Pt =1.25*20 =25mm
Cross –flow area As $=(Pt - d_0)\ Ds\ l_B / Pt$

Where:
Pt = tube pitch,
d_0= tube outside diameter,
Ds =shell inside diameter, m,
l_B = baffle spacing, m.

$As = 25-20/25 * 796 * 159.2 * 10^{-6} = 0.025\ m^2$
Mass velocity Gs = Ws/As
Where : Ws= fluid flow-rate on the shell - side, kg/s

$= 96443/3600 * (1/0.025)$
$= 1071\ kg/s.m^2$
Equivalent diameter $= 1.1/20(25^2 - 0.917*20^2) = 14.4$ mm
Mean shell side temp. =71°C

$Re = Gs\ d_e / \mu = 1071*14.4*10^{-3} / 7.53*10^{-4}$
=20481
Pr =4.98
Choose 25 percent baffle cut
From fig (3.5)(appendix A)
$j_h = 4.6*10^{-3}$
$hs = 0.59 / 14.4*10^{-3} * 20481*4.98^{1/3} * 4.6*10^{-3}$
$= 6591\ w/m^2.c$

Overall coefficient :

$1/U_0 = (1/h_o) + (1/hod) + d_o \ln(d_o/di) / 2 kw + (d_o/di)*(1/hid)$
$+ (d_o/di)*(1/hi)$

Where:
U_o = the overall coefficient based on the outside area of the tube, $w/m^2.c$
h_o = outside fluid film coefficient, $w/m^2.c$
hi = inside fluid film coefficient, $w/m^2.c$
hod = outside dirt coefficient (fouling factor), $w/m^2.c$
hid = inside dirt coefficient, $w/m^2.c$
kw = thermal conductivity of the tube wall material, w/m .c
di = tube inside diameter, m,
d_o = tube outside diameter, m.

$1/U_0 = (1/6591) + (1/5000) + 20*10^{-3} \ln(20/16)/2*50$
$+ (20/16)*(1/3000) + (20/16)*(1/4022)$
$= 891\ w/m^2.c$
well above assumed value of 860 $w/m^2.c$

pressure drop:

tube – side :

from fig 12.24 for Re =28501
$j_f = 3.5*10^{-3}$
$\Delta P_t = N_p [8 j_f (L/d_i) + 2.5) \rho u_t^2 / 2$

Where:
ΔP_t = tube-side pressure drop, N/m^2 (Pa),
N_p = number of tube-side passes,
u_t = tube-side velocity, m/s,
L = length of one tube.

$\Delta P_t = 2(8*3.5*10^{-3}*(4.83*10^{3}/16)+2.5)*993.6*1.35^{2}/2$
$=17569$ N/ m^2

Shell –side:

Linear velocity =Gs /ρ =1071 /993.6 =1.077m/s
from fig(12.30) (appendix A) for Re =20481
$j_f = 4.6*10^{-3}$
$\Delta P_s = 8\, j_f*(Ds/d_e)(L/l_B)\,\rho\, u_s^{2}/2$
where :
L = tube length,
l_B = baffle spacing

$\Delta P_s = 8*4.6*10^{-3}(796/14.4)*(4.83*10^{3}/159.2)*(993.6*1.077^{2})/2$
$=35564$ N/ m^2

3.4 Reactor design :

Catalyst:

Atriply promoted (K_2O-CaO|-AL_2O_3)iron oxide catalyst will be used The iron oxide (Fe_2O_3-FeO) is in the from of nonstoichiometric mangetite .It is made by fusing the mangetite with the promoters.

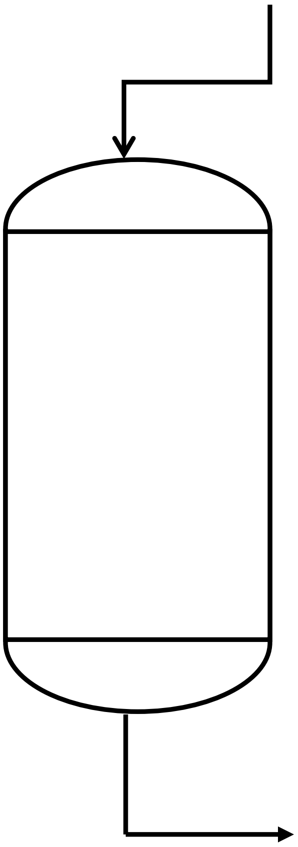
The iron oxide is catalyst

Properties:

1-partical size =6 -10 mm 2- Bulk density =1200 kg/m^3 3-partical density =305Ib/ft (4.9 g/cm^3)

204.2633 Kmol/h N_2
152.8182 kmol/h O_2
55.4788 kmol/h CH_4
612.7945 kmol/h H_2
9.3386 kmol/h CO
0.2267 kmol/h CO_2

P=200bar



400C˚(673)K˚

163.4103kmol/h N_2
152.8182kmol/h O_2
55.4788kmol/h CH_4
490.2356kmol/h H_2
9.3386 kmol/h CO

0.2267kmol/h CO_2
81.7059 kmol/h NH_3

425C˚(698)K˚

$1/2\ N_2+3/2\ H_2\leftrightarrow NH_3$
$N_2+3H_2\leftrightarrow 2NH_3$

The volume can be evaluated using method as SIMPSONS one –third Rule :

$V = \int_0^X \frac{FA^{\circ}}{-rA} dx \;=\; h/3\,\{f(x)_0+4f(x)_1+f(x)_2\}$

$H=X_2-X_0$

$-RA=K\ ca$

$K_1=k_1{}^{\circ}\ \exp -E/RT$

$K_1{}^{\circ}=1.78954*10^{-4}$ kmol/m^3.hr

$K_2 =K_2{}^{\circ}\exp -E/RT$

$K_2{}^{\circ}=2.5714*10^{6}$ Kmol/m^3.hr

E_1=20,800 Kcal /Kg mol

E_2=47,400 kcal/kg mol

$K_1=1.78959*10^{4}\ \exp -20800/8.314*673$

=0.0434 Kmol/m^3.h

K_2=2.5714*106 exp -47400/8.314*673

=538.398 Kmol/m^3.h

$\varepsilon=yA\ ^{\circ}\delta$

$\varepsilon=1*(2-4/2)$

$= -\ 0.5$

$CA^{\circ}=yA\ ^{\circ}p/\ RT^{\circ}$

CA°=1*197.2 atm/8.314*673

=0.0352 Kmol/m^3

FA°=204.2633 Kmol/h

Simpsons rule:

X	**(1-x)/(1+εx)**	**(1+x)/(1-x)=f(x)**

0	**1**	**1**
0.4	**0.75**	**2.33**
0.8	**0.33**	**9.1**

H=0.4 - 0= 0.4

F(x)=0.4/3 {1+4*2.33+9.1}

=2.58

V= 204.2633 /(538.398*0.0352) * 2.58

V= 27.81 m^3

V=Π /4 * D^2 *L

27.81= Π /4 *D^2*(4D)

D^3=8.852

D=2.06 m

Hence length (L)= 4*2.06 = 8.24 m

Catalyst is divided in 2 beds in ratio 1:2.5

Volume of catalyst in 1st bed =27.81/3.5 =7.94

Volume of catalyst in 2nd bed = 27.81- 7.94 =19.87

Bulk density of catalyst = 1200 Kg/m^3

Weight of catalyst in 2nd bed =19.87*1200

= 23844 tonne

Weight of catalyst in 1st bed = 7.94*1200 = 9528 tonne

Now giving allowance for space for gas movement

Upward and downward and insulation be 0.5 m

Hence diameter of reactor becomes =2.06-0.5 =1.56 m

Pressure at which reactor works = 200 Kg/m^2

Let factor of safety =20%

Design pressure = 1.2*200

= 240 Kg/m^2

We use carbon steel with internal lining of titanium

Allowable stress=66,000 psi {hesse and ruston}

=4494 Kgf/cm^2

NOW allowable stress =$p_D(1/k_2-1)$ from(Hesse)(9)

Where p_D= design pressure

4494=240 $(1/k_2-1)$

K=1.0337

D° = K Di

D°=1.0337*2.06

D°=2.129 m

Thickness of shell= (2.129-2.06)/2

= 32 mm

Operating pressure: 200 Kg/m^2

Design pressure: 240 Kg/m^2

Temperature : 673-698 k°

Allowable stress: 4494 Kgf/cm^2

Diameter inside of converter : 2.06 m

Bulk density of catalyst: 1200 Kg/m^3

Volume : 27.81 m^3

Thickness of shell: 34mm

Volume of 1st catalyst bed : 7.94 m^3

Volume of 2nd catalyst bed : 19.87 m^3

Chapter four

Process control

4.1 Introduction

Ammonia is one of the top 10 chemicals produced today. Most ammonia is produced by steam methane reforming of a natural gas feedstock. 85% of the ammonia produced is used for fertilizer in one form or another.

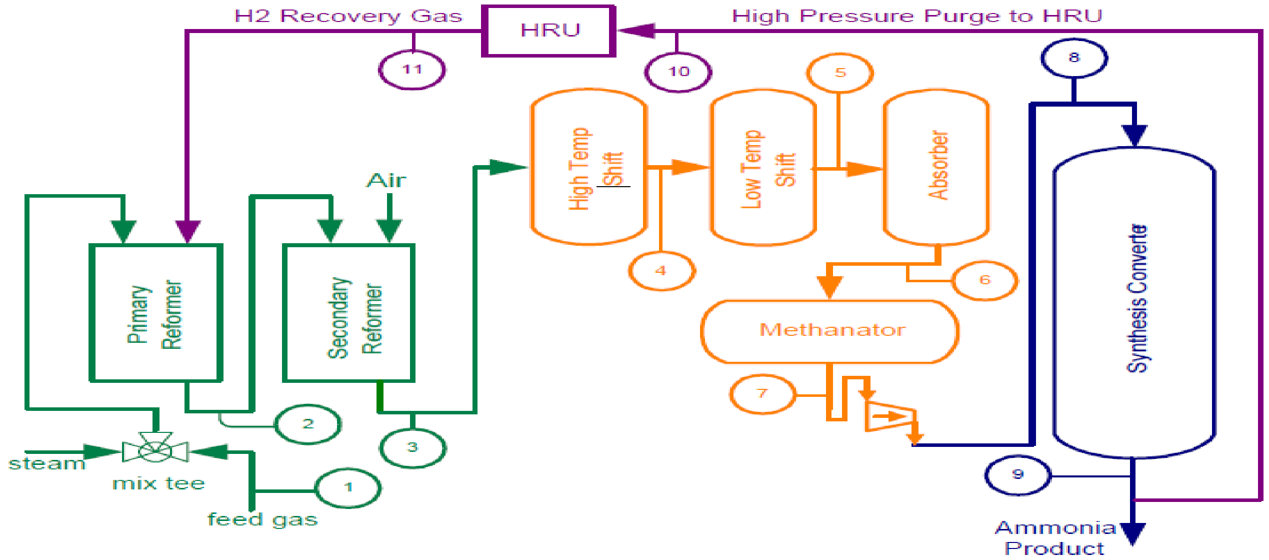
Other industrial uses for ammonia include fibers, plastics, coatings and resins. In looking
to improve process efficiency and cut costs ammonia manufacturers have adopted a closed loop control strategy.

The key to the success of this strategy is the Extrel CMS MAX300-IG Process Mass Spectrometer. The MAX300-IG is a rugged Quadrupole mass spectrometer utilizing sophisticated electronics and offering fully automatic operation.

The MAX300-IG is a universal analyzer offering stable real time data acquisition and a dynamic range of 10 ppb to 100%. The exact concentrations of the stream components are immediately available for integration into the control scheme.

Due to its reliability, speed and accuracy the MAX300-IG provides unsurpassed performance in ammonia production. The MAX300_IG allows for tight control of four important process variables: the steam to carbon ratio, the H_2:N_2 ratio, the methane slippage and the process inert gases.

Maintaining tight control of these variables results in three significant economic benefits to ammonia producers: lower fuel costs, higher yields and an overall optimized efficiency. Figure 1 is a simple diagram of the ammonia process indicating the typical streams analyzed by the MAX300-IG.



Feed Gas

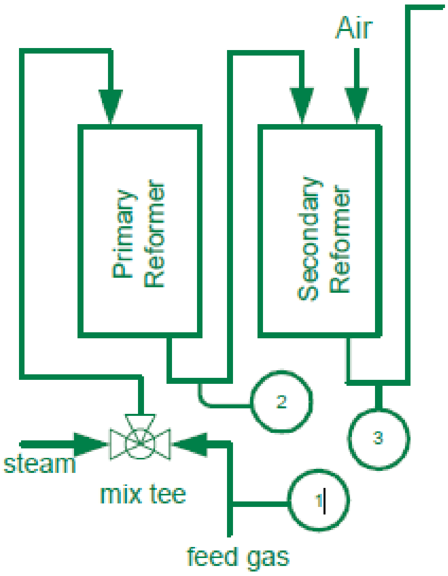
In the Primary and Secondary Reformers, a hydrocarbon feed stream and steam are mixed and passed over a reformer containing a nickel-based catalyst at 1500F. This results in the formation of carbon monoxide and hydrogen according to the following equation.

$CH_4 + H_2O \rightarrow CO + 3H_2$

Controlling Steam Generation Costs through Control of the Steam to Carbon Ratio to a tolerance of ±0.02%

The formation of carbon, known as coking, is detrimental to the catalyst. To prevent coking, an excess amount of steam is mixed with the feed gas than what the reaction requires. A minimum ratio of 3.0 lb. moles of steam to 1.0 lb. moles of carbon, i.e., 3:1 should be maintained. However, to avoid dropping below this ratio and degrading the catalyst, plants normally operate with generously elevated steam to carbon ratios.

The generation of steam is one of the most expensive processes in ammonia production. Typically, steam is produced at a level 4 times that of the ammonia product. Tight control of the steam to carbon ratio can significantly decrease production costs.

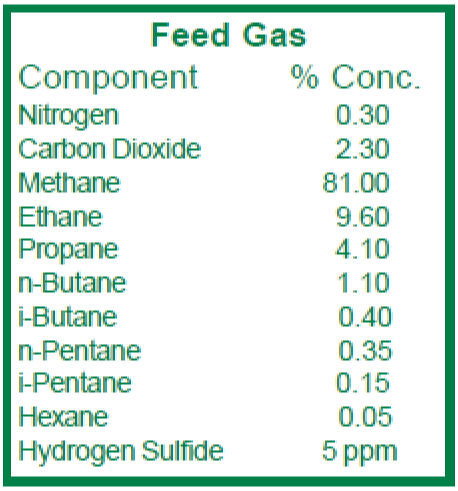


The MAX300-IG performs rapid analysis of the feed stream for hydrocarbon concentration and calculates the BTU value.

This information is used to control the steam flow; allowing rapid adjustments and avoiding potential converter problems. The MAX300-IG typically analyzes hydrocarbons up to and including hexane in a Natural Gas Feed Stream.

The instrument has the ability to analyze the butane and pentane isomers separately. This results in better accuracy for the calculated BTU value. Using the MAX300-IG as part of a closed loop strategy, Steam to Carbon ratio tolerances of ±0.02 have been demonstrated.

Using the electron multiplier dual detector option, the MAX300-IG can also analyze the Feed Gas stream for the presence of sulfur containing components such as hydrogen sulfide and mercaptans. Sulfur components are poisons and will de-activate the catalyst.



Feed Gas

Component	% Conc.
Nitrogen	0.30
Carbon Dioxide	2.30
Methane	81.00
Ethane	9.60
Propane	4.10
n-Butane	1.10
i-Butane	0.40
n-Pentane	0.35
i-Pentane	0.15
Hexane	0.05
Hydrogen Sulfide	5 ppm

4.2 Primary Reformer Effluent and Secondary Reformer Effluent:

The amount of un-reacted methane in the Primary Reformer Effluent and Reformer Effluent is referred to as methane slippage. Methane slippage, also called leakage, is an indication of the reformer efficiency in the conversion of methane to hydrogen and carbon dioxide.

The control objective is to maintain the methane slippage out of the secondary reformer at a very low and constant value, typically 0.3% methane or less. The analysis of the Primary and Secondary Reformers requires both a highly sensitive and very stable sensor because of the complex chemical makeup.

A wide dynamic range is also required as the methane conversion starts at 90% methane and ends at 0.3% methane. The sensitivity and the stability of the MAX300-IG enables control tolerances of 0.05% to be achieved.

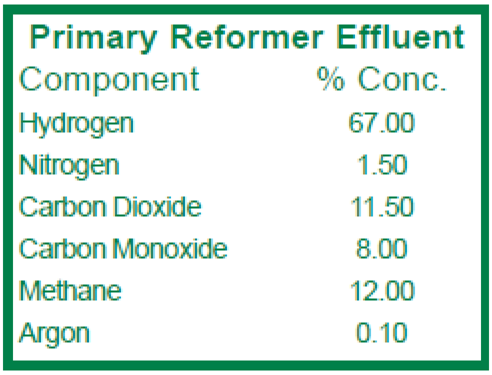
These tight tolerances allow the operator to minimize the changes in methane slippage thereby maximizing yield and lengthening equipment life. Increases in run time and equipment life of up to 20% have been observed in plants utilizing closed loop control with an Extrel CMS MAX300-IG.

Maximizing Yield and Equipment Life through Unsurpassed Control of Methane Slippage

Air is used as the source of the nitrogen required for the ammonia reaction. Air is introduced into the process at the secondary reformer **in** an amount to yield an exit gas that has a 3:1 hydrogen to nitrogen ratio. The oxygen in the air reacts with the hydrocarbon feedstock in combustion and supplies the energy to reform the remainder of the gas. The secondary reformer reaction increases the hydrogen content and oxidizes most of the CO to CO_2.

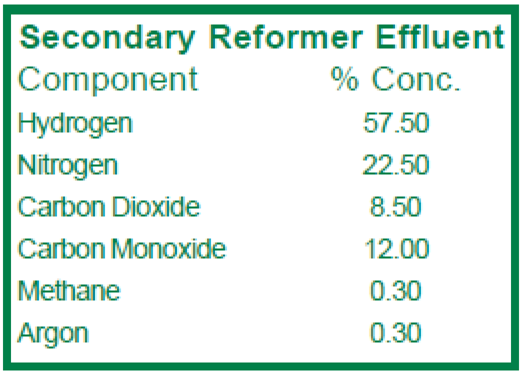
CO_2 is preferred over CO as it is much more soluble in water and can easily be removed further along in the process. The Secondary Reformer Outlet also contains a small amount of methane and argon which enter the synthesis loop as inert gases.

$CH_4 + H_2O \rightarrow CO + 3H_2 + (CH_4)$
$H_2 + Air \rightarrow 3H_2 + N_2 + (CO_2 + CO)$



Primary Reformer Effluent

Component	% Conc.
Hydrogen	67.00
Nitrogen	1.50
Carbon Dioxide	11.50
Carbon Monoxide	8.00
Methane	12.00
Argon	0.10



Secondary Reformer Effluent

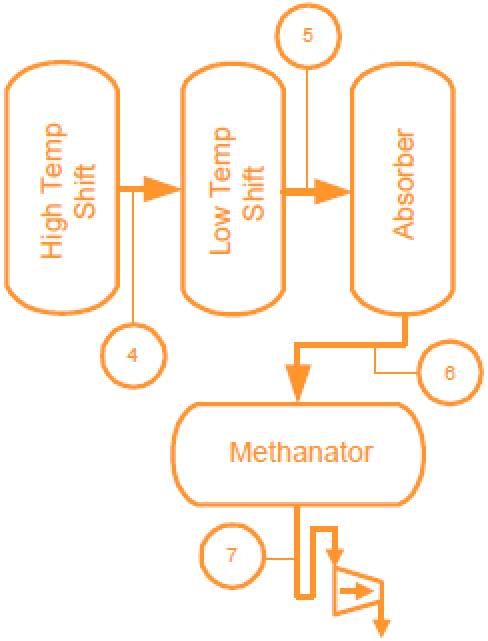
Component	% Conc.
Hydrogen	57.50
Nitrogen	22.50
Carbon Dioxide	8.50
Carbon Monoxide	12.00
Methane	0.30
Argon	0.30

4.3 CO2 Absorber and Methanator:

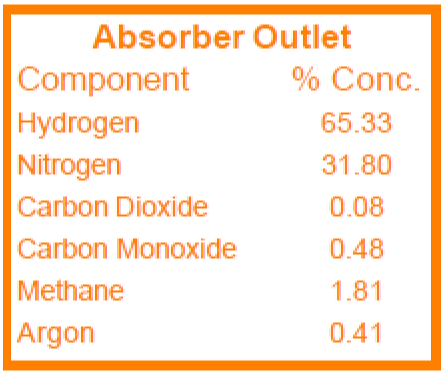
The CO_2 is generally removed by absorber-regenerator units containing many types of absorbent including monoethanolamine solution (MEA), sulfinol , propylene carbonate and others. After this process, CO in the product gas is reduced to less than 100 ppm. CO_2 and CO collectively known as carbon oxides, are poisons for many types of catalysts.

The trace amounts of CO and CO_2 remaining after the absorber must be removed from the synthesis gas. This is done by converting them to methane over a nickel or ruthenium catalyst in the presence of hydrogen

in the Methanator. The amount of methane produced by this process is insignificant compared to the level of the un-reacted methane slip.
After this process, the residual levels of the carbon oxides in the process gas will be less than 5 ppm.

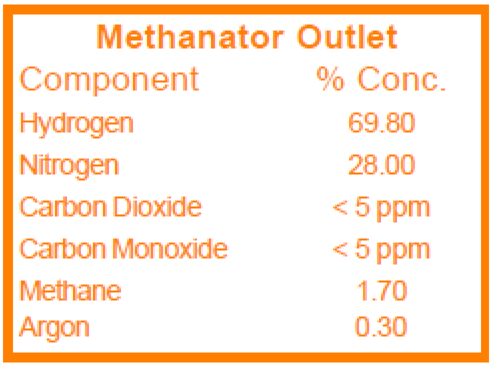


$CO + 3H_2 \rightarrow CH_4 + H_2O$
$CO_2 + 4\,H_2 \rightarrow CH_4 + 2H_2O$



Absorber Outlet

Component	% Conc.
Hydrogen	65.33
Nitrogen	31.80
Carbon Dioxide	0.08
Carbon Monoxide	0.48
Methane	1.81
Argon	0.41



Methanator Outlet

Component	% Conc.
Hydrogen	69.80
Nitrogen	28.00
Carbon Dioxide	< 5 ppm
Carbon Monoxide	< 5 ppm
Methane	1.70
Argon	0.30

Chapter five

Location ,Economic &Safety Consedration

5.1 Location and site considerations:

The site chosen for the plant chemical directly affect the success or failure of the manufacturing process. The basic principle is to have the factory site so that the cost of production and distribution of the product at the minimum , and this can only be achieved if it is taking a number of important factors into consideration during the design and selection process plant site.

The most important considerations to be followed in the selection of the plant site are:

1. Sources of raw materials
2. Consuming markets
3. energy resources
4. the climate
5. transport services
6. Water Resources
7. Waste disposal
8. Labor
9. Financial considerations

10. Prevention From Fire and floods

11. Proximity From Positions Industrial

5.1.1. Sourcing of raw materials

That one raw material processing of the most important factors that go into determining the plant site especially when large quantities required. The factory near the site of the raw materials are not allowed to reduce transportation expenses only, but to reduce the capital invested in storage services as well.

5.1.2. Common Markets

The viability of the delivery of the final product to the consumer quickly and the lowest possible cost no less important than site - built near the source of raw materials. So From the necessary that Be there is Comparison between Proximity From Items the source Crude and ease receipt Product to me The consumer. The choice Will depend On the quantity Required And the cost Transfer Articles Crude And the product. When Taking Marketing For the product Final The eyes consideration

Must that Calculates an account Sale And distribution Outputs Occasional Often What Be Factor at Profit Work.

5.1.3. Energy Sources

In most chemical factories need large fuel and energy be when coal and oil shale amounts required large site near the source of the processing is necessary , therefore the process is economical. When large amounts of electrical power is available , it is one of the very important factors to be taken into consideration, it is noted that in the case of the need for a large ownership of the energy it is better to be near the site of the hydroelectric generators.

5.1.4. Climate

It is possible that a large climate impact of the economic process. When Aldharov be harsh climate, the cost of personal protective buildings and services must be part of the total cost. In hot weather it may be dismissed in cooling towers, air - conditioning and possibly freeze coefficient of important considerations.

5.1.5. Transportation services

The choice of modes of transportation for both raw materials and the final product depends no doubt on the quality and quantity of the materials used , if possible, provide the three modes of transportation of any roads and population as rail and maritime transport. It is important that the plant site near either the railway station or docks.

5.1.6. Water Resources

Most industrial processes require large amounts of water not only for industrial process but also used for cooling processes, washing and steam generation processes. To make the process economical, the plant site or plant near a river or lake is an important and necessary.

5.1.7. Waste disposal

The water supply is linked to a large extent with the waste disposal problem, whether solid, liquid or gaseous where there are strict limitations on the issue of waste disposal and it changes according to industrial unit site.

It is necessary to study the stamina to be allowed for water, earth, air metastatic sites and in several cases the level of a test as a result of the center. The additional cost due to treatment outcomes emitted must be balanced with other appropriate site selection factors.

5.1.8. Employment

Although the trend is the increased use of automated equipment in the lab , a chemical addition to the angel of the assistant engineers , but there are several processes still need a large workforce especially when using a rotation system , it is important to choose the industrial unit site , taking into account the payment of wages prevailing rate specific region and industry competition as well as skilled and intelligent workers. In remote spokesman it should be calculated to provide housing and causes of social amenities for the workers expense.

5.1.9. Financial considerations

When trying to create a sophisticated industry in a particular area should provide the necessary and which have a significant impact in the selection of the plant site and financial incentives. It is also necessary to take the applicable tax system into consideration . When choosing the location of his family in the area of future expansion should his account , especially in the busy industrial areas.

5.1.10. Prevention of fires, floods

as such We mentioned Previously, for the purpose of Large amounts of cooling water for the chemical factories processing location is often near rivers and lakes .walhakiqh that this site will be vulnerable to flooding so it is necessary to know the historic flood that has happened in that region. It is important to provide adequate services to prevent n fire in the region, as well as knowledge of its impact on neighboring sites. The Earth and its terrain with a price tag and buildings should take into consideration.

5.1.11. Proximity to industrial centers

The plant site near the important industrial centers be useful in terms of getting large numbers of manpower of different means of transport. Perhaps there are other industries adjacent to bid for materials and services at low prices , and there are several cases where materials processing manufacturers have directly.

•Due to this reason we find that the desired position is basra because production of urea is used as raw material .

5.2 Safety and environmental consecrations:

Causing ammonia in the incidence of infections and irritabilities flogging and nose and eyes and throat part upper from the device respiratory and where that Ammonia the source the main element Nitrogen necessary growth the plants aqueous it possible that contribute Ammonia at a plus enrichment food surface Aqueous stagnant or a slow syriacs especially those spaces of the specific content of nitrogen addition to that is the average of ammonia toxin for objects water even right now did not prove scientifically that ammonia from articles casing cancer.

Returns Problems Environmental Which Cause In which Ammonia to me Feature Solubility water. The tendency Leak at Case Gaseous The Cause Ammonia at Happening Damage seriously life wilderness when Absorbed at Currents Water it causes Damage Fish To Fish water cold As well As for at Air Vtaathd Ammonia With Ions Sulfates And melt at waters Rain To return Quickly to me the soil The flats Aqueous

Safety Precautions for Ammonia:

- CORROSIVE, COMPRESSED GAS. May also be an EXPLOSION HAZARD, especially in confined spaces. Engineering controls, proper training, protective equipment requirements and personal hygiene measures are essential.
- Never work alone with this chemical.
- Unprotected persons should avoid all contact with this chemical, including contaminated equipment.
- Do not use with incompatible materials such as oxidizing agents (e.g., nitrogen oxide), halogens (e.g., chlorine, fluorine) and heavy metals (e.g., mercury, silver).
- Shut flow off at cylinder valve and not just at the regulator after use. Replace outlet caps or plugs and cylinder caps as soon as cylinder is disconnected from equipment.

  - Make sure cylinders are labeled clearly. Avoid damaging cylinders. Move cylinders by hand truck or cart designed for that purpose.

- Keep empty cylinders under slightly positive pressure. Do not use cylinders as rollers or for any other purpose than to contain the gas as supplied.
- Follow handling precautions on Material Safety Data Sheet. Have suitable emergency equipment for fires, spills and leaks readily available. Practice good housekeeping. Maintain handling equipment. Comply with applicable regulations.
-

5.3 Economic and feasibility considerations:

5.3.1. cost estimation of Absorption column:

Height = 5 m
Diameter = 0.87 m
Pressure = 30 bar
Material of construction : Carbon steel
Cost = 8000$
Pressure factor = 1.6
Material factor = 1
Cost = 8000*1.6*1 = 12800 $ in 2004
Cost in 2016 = (cost in 2004) *(cost index in 2015/cost index in 2004)
Cost index in 2004 = 480 $
Cost index in 2015 = 1081 $
Cost in 2016 = 12800 *(1081/480)
Cost in 2016 = 28826.66667 $
= 37186400 I.D

5.3.2 Cost estimation of Heat exchanger :

Area =210 m^2
Cost = 7500 $
Pressure factor = 1
Material factor = 1

Cost = 7500*1*1 = 7500 $ in 2004
Cost in 2016 = cost in 2004 *(cost index in 2015/cost index in 2004)
Cost index in 2004 = 480$
Cost index in 2015 = 1081$
Cost in 2016 = 7500*(1081/480)

Cost in 2016 = 16890.625 $

= 21788906.25 I.D

5.3.3 cost of estimation of reactor

D = 2.04 m

H = 8.24 m

P = 200 bar

Material factor = 1

Pressure factor = 7.8

Cost = 12200 $

Cost = 12200 *7.8*1 = 95160 $ in 2004

Cost in 2016 = cost in 2004 *(cost index in 2015/cost index in 2004)

Cost index in 2004 = 480 $

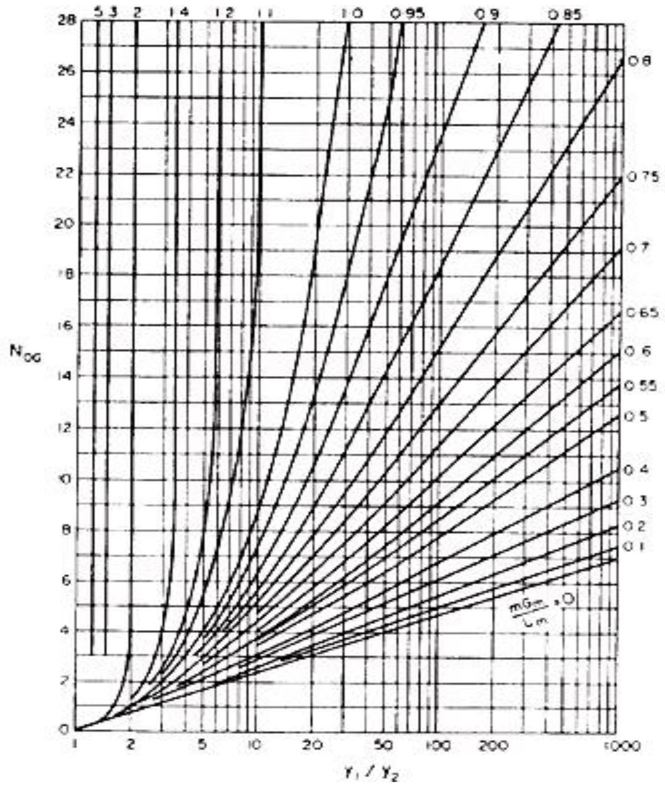
Cost index in 2015 = 1081 $

Cost in 2016 = 95160 *(1081/480)

= 214308.25$

= 276457642.5 I.D

Appendix A:



Figure(3.1) number of transfer unit NOG as a function y_1/y_2 with mG_m/L_m as parameter

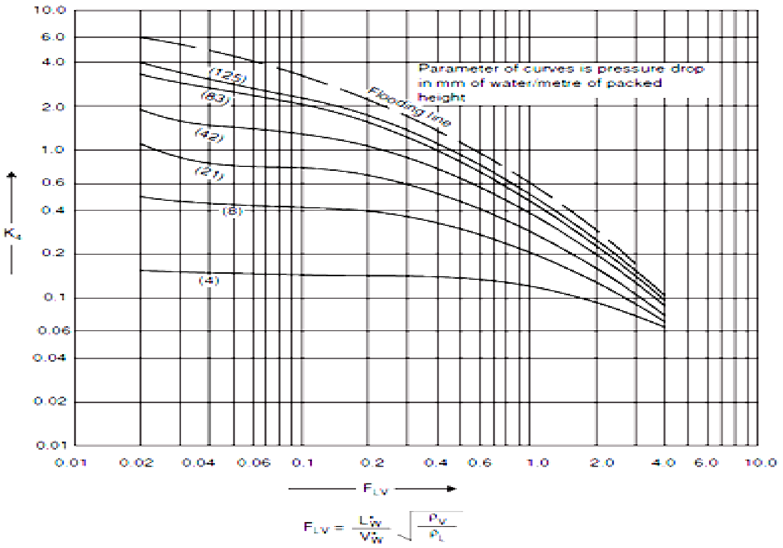


Figure (3.2) Generalized pressure drop correlation ,adapted from a figure by the Norton CO. with permission

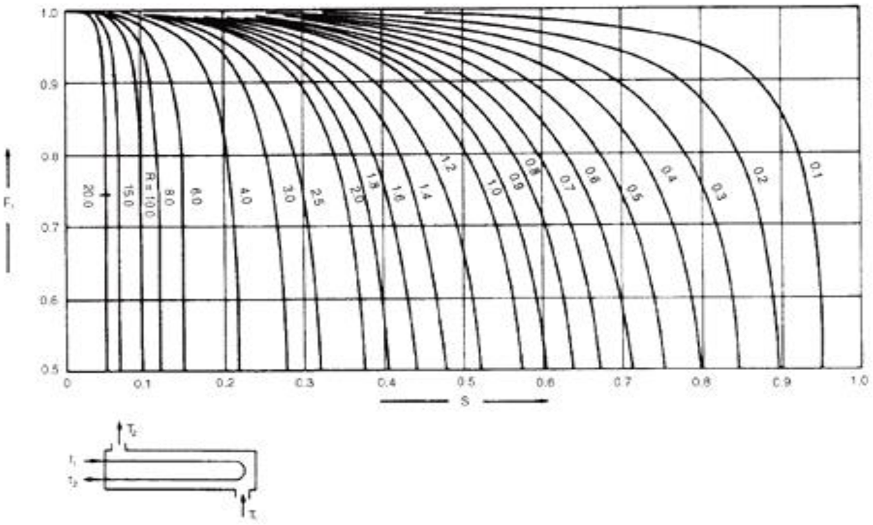


Figure (3.3)Temperature correction factor ,one shell pass ,tow or more even tube passes

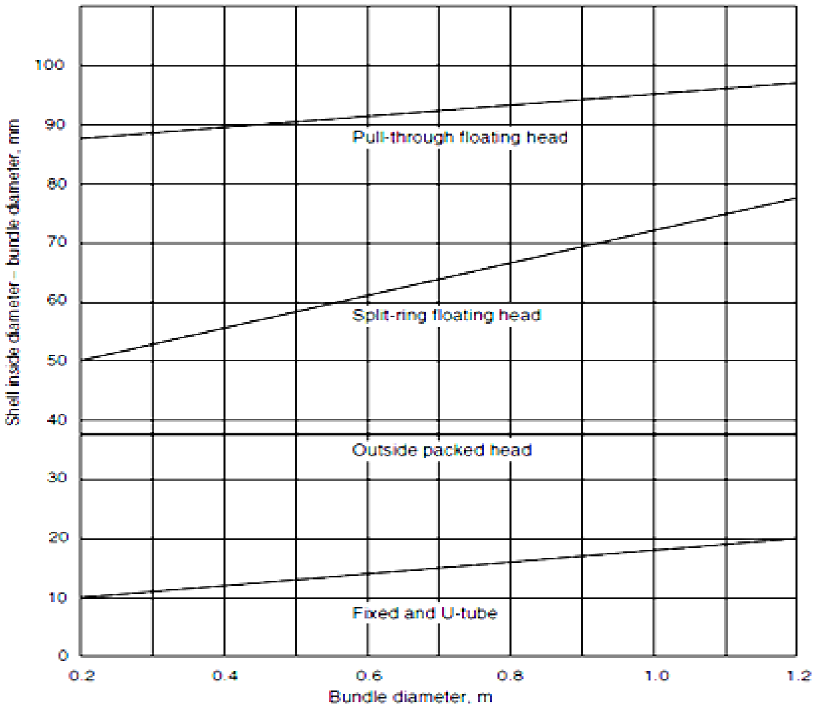


Figure.(3.4) shell bundle clearance

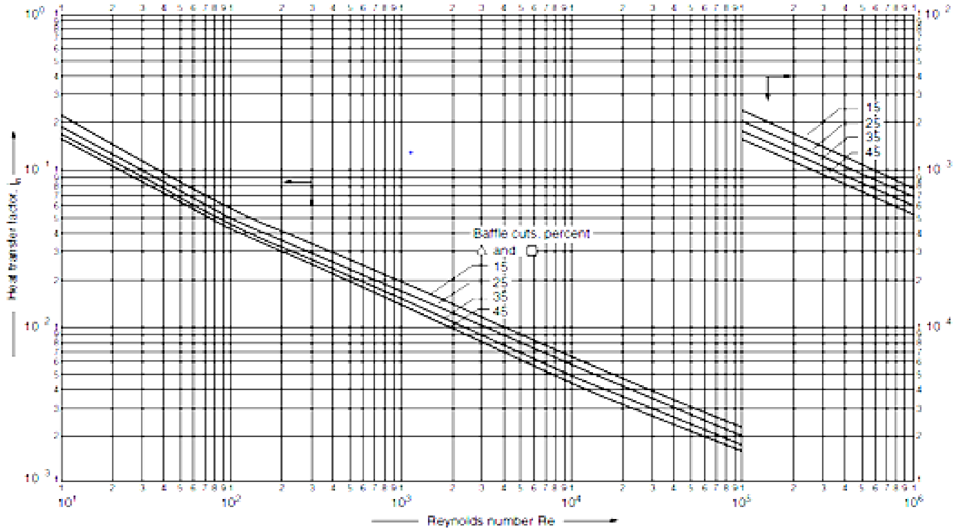


figure (3.5) Shell side heat transfer factor ,segmental baffles

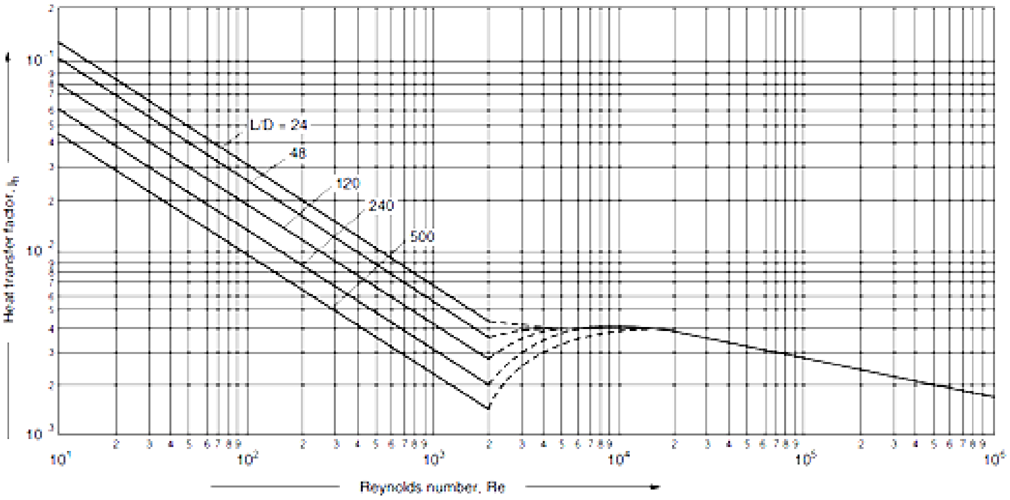
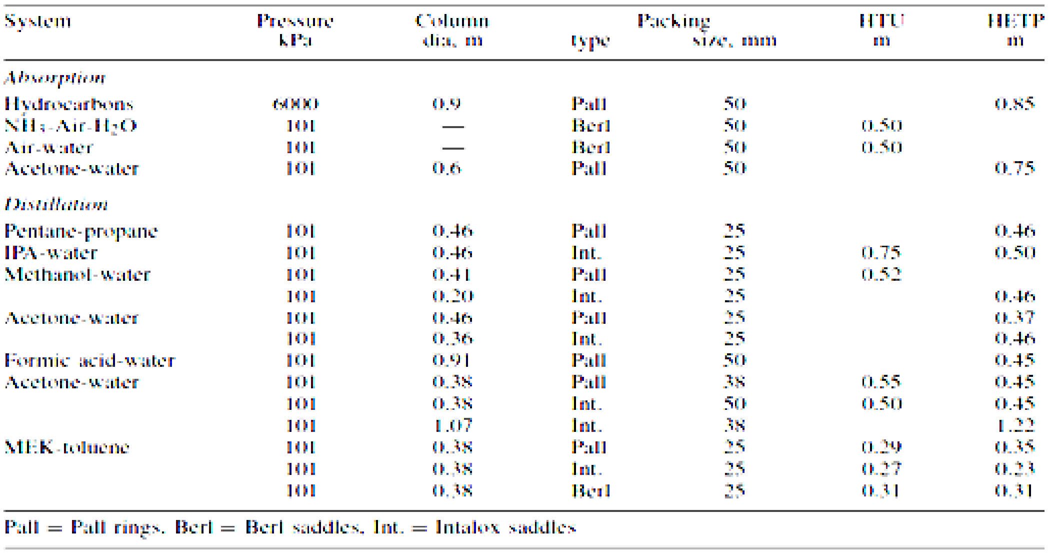


Figure (3.6)Tube side heat transfer factor

Appendix B:



System	Pressure kPa	Column dia, m	Packing type	Packing size, mm	HTU m	HETP m
Absorption						
Hydrocarbons	6000	0.9	Pall	50		0.85
NH_3-Air-H_2O	101	—	Berl	50	0.50	
Air-water	101	—	Berl	50	0.50	
Acetone-water	101	0.6	Pall	50		0.75
Distillation						
Pentane-propane	101	0.46	Pall	25		0.46
IPA-water	101	0.46	Int.	25	0.75	0.50
Methanol-water	101	0.41	Pall	25	0.52	
	101	0.20	Int.	25		0.46
Acetone-water	101	0.46	Pall	25		0.37
	101	0.36	Int.	25		0.46
Formic acid-water	101	0.91	Pall	50		0.45
Acetone-water	101	0.38	Pall	38	0.55	0.45
	101	0.38	Int.	50	0.50	0.45
	101	1.07	Int.	38		1.22
MEK-toluene	101	0.38	Pall	25	0.29	0.35
	101	0.38	Int.	25	0.27	0.23
	101	0.38	Berl	25	0.31	0.31

Pall = Pall rings. Berl = Berl saddles. Int. = Intalox saddles

Table (3.1) typical packing efficiencies

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