

BANGLADESH UNIVERSITY OF ENGINEERING & TECHNOLOGY



DESIGN OF A 1200 TPD AMMONIA PRODUCTION PLANT

**BACHELOR OF SCIENCE IN ENGINEERING
(CHEMICAL)**

SUBMITTED BY

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May, 2023

Statement of Academic Integrity

The design project titled “Design of a 1200 TPD Ammonia Production Plant” for the course CHE-408 is completed by Md. Mostafizur Rahman, Zulkar Nyene Bhuyan, Sadman Islam and Tasfia Ferdoush under the supervision of Dr. Nahid Sanzida. This project is our original work and the help we have taken from external sources is provided clearly with due reference to the literature.

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Acknowledgement

This design project was done under the supervision of Dr. Nahid Sanzida, Professor, Department of Chemical Engineering, BUET. The authors would like to express heartfelt gratitude to Dr. Nahid Sanzida for her guidance, generous help and timely suggestions at every stage of this design project. The authors would also like to express gratitude to Karnafuli Fertilizer Company for arranging visit and providing necessary data.

The writers also acknowledge and appreciate the opportunity of conducting this activity and express their gratitude towards Department of Chemical Engineering, BUET

Abstract

An ammonia plant having 1200 TPD capacity have been designed under the course CHE 408. The location of the plant is selected at Rangadia, under Anowara Upazila, Chittagong, Bangladesh, beside the river Karnaphuli. Preliminary market study was done before choosing the plant capacity for the design project as ammonia has higher demand nowadays. Haldor Topsoe process is selected for producing ammonia based on comparison with other processes. The process is defined as stated in Chapter 2. Material and Energy balances are done based on some assumptions and plant data obtained from KAFCO. 4 major equipment such as Primary Reformer, Low Temperature Shift Reactor, CO₂ absorber and Ammonia converter are designed. After doing economic analysis, it can be said that total capital investment of this plant will be 506.68 million dollars, total production cost will be 254.65 million dollars per year and net profit will be 53.37 million dollars per year. To check the feasibility of the plant IRR, ERR are calculated and a sensitivity analysis is shown in the economic analysis section. IRR will be 14% and ERR will be 10.85% for this plant. Equipment layout and plot plan of the plant are given for this plant.

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Chapter 1

INTRODUCTION

Ammonia is an ingredient used as an intermediary in the production of nitrogenous fertilizers. It can either be applied directly to the soil or mixed with water-based solutions of other nitrogenous fertilizers, such as urea and/or ammonium nitrate. In addition to these, ammonia is used to make nitric acid, soda ash, cleaning products, leather tanning, oil refining, pulp & paper, textiles, refrigeration, rubber & synthetic resin industries, explosives, and food & beverages. Nowadays demand of ammonia increases as demand of nitrogenous fertilizer, herbicides and insecticides increase for higher food demand. Physical and chemical properties of ammonia, raw materials of ammonia production and market demand analysis are given below-

1.1 Physical and Chemical Properties of Ammonia

1.1.1 Physical Properties of Anhydrous Ammonia

Table 1.1.1: Physical Properties of Ammonia (Widiawati et al., 2020)

Property	Detail
Molecular Mass	17.03 g/mol
Colour	Colourless
Odour	Sharp, Irritating
Physical State	Gas (at room temperature)
Melting Point	-77.7°C
Boiling Point	-33.35 °C
Flash Point	11 °C
Decomposition Point	500 °C
Density (Gas)	0.7710 g/L
Density (Liquid)	0.6818 g/L
Vapour Density	0.5697
Critical Temperature	132.4 °C
Critical Pressure	111.3atm
Heat of Fusion	58.1 kJ/mol
Heat of Vaporization	23.3 kJ/mol
Heat of Combustion	-316 kJ/mol

1.1.2 Chemical Properties of Anhydrous Ammonia

Table 1.1.2: Chemical Properties of Ammonia

Property	Details
Chemical Formula	NH ₃
Type of Base	Weak
Affinity (Water)	High
Corrosiveness	Corrosive to Some Metals
Oxidation Power	Strong Reducing Agent
Reactivity	Quite Reactive
Volatility	Increases with increase in pH

1.2 Raw Materials of Ammonia Production

Raw materials used for ammonia production are methane, steam and air. Methane and steam are used to produce syngas. Air is the source of N₂ and syngas is the source of H₂. Sources of methane are natural gas, naphtha, gasification coal etc. However, natural gas is widely used in our country. Steam is produced from natural sources of water after proper treatment.

1.3 Market Demand

The ammonia industries represent significant and vital components of the chemical economy. They serve as significant sources of revenue and employment, in particular within countries with significant availability of feed stocks such as natural gas and coal. The principal application of ammonia is within the fertilizer sector, which is critical to agricultural production and the wider world economy. Apart from that ammonia also used as neutralizing acid constituents of crude oil during the refining process, purification of smoke gas in coal fired or oil fired power plants, as a source of hydrogen for hydrogenation of oils and fats, for extraction of copper, nickel and molybdenum from their ores, processing agent in the manufacturing of pharmaceuticals, substitute for calcium in the pulping of wood. In Bangladesh there are mainly seven ammonia industries from which six in the public sector and one in private sector as a joint venture between the government of Bangladesh and foreign companies. These industries have the total capacity of almost 5.89 MT/day where the demand is 21.66 MT/day. There are 148 active Ammonia Buyers in Bangladesh. Top 3 Ammonia Buyers are Star Multipurpose Cold Storage, Transcom Distribution Co and Bangladesh Krishi Bank Sharsha BR. These three industries are also the top importers of ammonia in Bangladesh.

Global ammonia consumption in 2020 is estimated at around 187 million tons. Global demand growth over 2019 was modest given strong demand for fertilizers against subdued industrial consumption at the beginning of the year due to disruption from the COVID-19 pandemic. Despite interruptions to global trade, governments across the world have prioritized food security, hence little impact has been seen in overall fertilizer consumption of ammonia and urea globally. Total global consumption of ammonia was 184 million tons in 2019, an increase of around ten million tons over 2018. The top three consuming regions are Asia Pacific (with China and India accounting for the majority of the demand), North America and Europe.

Chapter 2

PROJECT DEFINITION

An ammonia plant, licensed by Haldor Topsoe process, is to be set up at Rangadia, under Anowara Upazila, Chittagong, Bangladesh, beside the river Karnaphuli, with a capacity of 1200 tonne/day NH_3 production and 1373 tonne/day CO_2 as by product, corresponding to 396000 tonnes of NH_3 and 453090 tonnes of CO_2 production per year, including all offsite, auxiliaries, utilities and supporting facilities within 75 acres area, using natural gas (97.73% CH_4) as feedstock and fuel taken from Karnaphuli Gas Distribution Company, underground water and ambient air as raw material.

Chapter 3
SELECTION OF PROCESS AND PROCESS DESCRIPTION

3.1 Selection of Process

There are 4 major process licensors available for ammonia production. They are- Haldor Topsoe, Kellogg, Udhe and Casale. Mostly used process licensors in Bangladesh are Kellogg and Haldor Topsoe. Factors that differentiate these 2 processes are- types of ammonia converter, types of primary reformer used based on firing system, refrigeration system. Comparison between these two processes are given below-

- 1. Reaction conditions:** The Kellogg process uses a lower pressure of around 150-250 atmospheres and a higher temperature of around 800-900°C. On the other hand, the Haldor Topsoe process uses a higher pressure of around 200-300 atmospheres and a lower temperature of around 450-550°C.
- 2. Catalyst:** In the Kellogg process, the catalyst used is made of iron oxide and alumina. In contrast, the Haldor Topsoe process uses a ruthenium-based catalyst.
- 3. Efficiency:** The Haldor Topsoe process is considered to be more efficient than the Kellogg process. The Haldor Topsoe process has a higher conversion rate of hydrogen and nitrogen to ammonia, which means that more ammonia can be produced using the same amount of feedstock. Additionally, the Haldor Topsoe process requires less energy to operate, which results in lower production costs.
- 4. Equipment:** The equipment used in the two processes is different. The Kellogg process requires a multi-stage reactor system, while the Haldor Topsoe process uses a single-stage reactor system.

Based on the comparison, it is selected that this plant will be based on Haldor Topsoe process. TOPSOE™ Low Energy Ammonia Process is energy efficient and can be applicable to both small and large-scale ammonia production. Advantages and disadvantages of this process are listed below-

Advantages

- Lowest possible energy consumption and lowest emission.
- Reforming is an important step because efficient and reliable conversion of feed to syngas has great impact on the economics of the plant. TOPSOE side fired reforming furnace, specialized secondary reformer burners and high active reforming catalysts contribute to improve the operation of the plant.
- CO shift conversion is an important factor in determining operating cost of plant. TOPSOE shift reactors maximize the conversion for an extended period of time.

- TOPSOE radial flow converters (S-200, S-300, S-50) and KM catalyst ensures an optimal and stable performance of ammonia synthesis. These converters have simple design and provide efficient conversion.
- CO₂ recovery technologies are designed in most optimal way to ensure reliable operation. H₂ and N₂ gases are also recovered from purge gas to increase the percentage of NH₃ in product stream. About 20% recovery is possible in Haldor Topsoe process.

Disadvantages

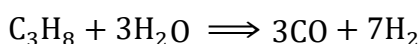
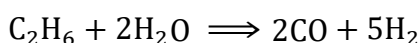
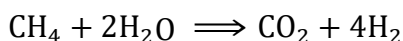
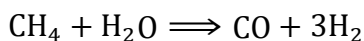
- Shorter life span of converters.
- 20% efficiency loss in making up gas which is unconvertable.
- Regeneration of Benfield solution consumes energy and this process requires time.

3.2 Process Description

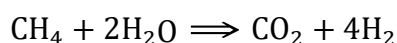
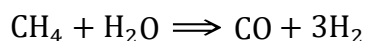
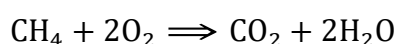
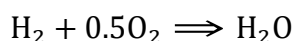
Haldor Topsoe process has the following steps [Appl. M (1999). "Ammonia- Principles and Industrial Practices", WILEY-VCH] -

3.2.1 Preparation of Raw Synthesis Gas

- 1. Process natural gas purification-** Process natural gas should be purified before reforming. Ni-containing catalysts that are used in reformers can be poisoned by sulfur. So desulfurization of process natural gas is important. Co-Mo and ZnO catalysts are used for desulfurization. Temperature range for this operation is usually 390-400°C. In our country, sulfur is not present in natural gas. So, this step is not required.
- 2. Primary reforming-** In primary reforming, process natural gas and steam are fed to the primary reformer at high temperature. Steam reforming reaction occurs inside the reformer tubes which are filled with Ni-containing catalyst. As reforming reaction is endothermic, heat is supplied to the tubes by combustion of fuel natural gas in furnace. Temperature range for primary reforming is 520-750°C. High steam to carbon ratio (S/C) is maintained to avoid coke formation on catalyst. Primary reformer furnace has different configurations- top fired, bottom fired, side fired and terrace fired. Reactions that occur in reformer tubes are-



- 3. Secondary reforming-** It is a refractory-lined vessel with Ni catalyst. In secondary reformer air and outlet stream of primary reformer are fed. Combustion reaction occurs in upper portion which produces heat for endothermic combustion reaction. In lower portion, residual methane reforms. Temperature range for secondary reforming is 550-950°C and 32.3 kg/cm² (gage). Air is compressed to this pressure and then injected into primary reformer effluent. Reactions that occur in secondary reformer are-



3.2.2 Raw Synthesis Gas Purification

1. Shift reaction- Water gas shift reaction is important for converting CO to CO₂. There are 3 types of shift conversion- high temperature shift conversion, medium temperature shift conversion and low temperature shift conversion. Among them, high and low temperature shift conversions are widely used in designing ammonia plant.

Operating condition of High Temperature Shift (HTS) conversion:

Temperature- 350-420°C

Pressure- 32.2 kg/cm² (gage)

Catalyst- Fe oxide

Reaction- $\text{CO} + \text{H}_2\text{O} \rightleftharpoons \text{H}_2 + \text{CO}_2$

Operating condition of Low Temperature Shift (LTS) conversion:

Temperature- 220-260°C

Pressure- 31.2 kg/cm² (gage)

Catalyst- CuO + ZnO

Reaction- $\text{CO} + \text{H}_2\text{O} \rightleftharpoons \text{H}_2 + \text{CO}_2$

At first, secondary reformer effluent is passed through HTS converter and then through LTS converter. There is a waste heat boiler between these 2 converters.

2. CO₂ absorption- CO₂ is poisonous for ammonia synthesis catalyst. So, CO₂ should be removed from syngas. CO₂ that is removed from syngas is used as raw material for urea synthesis. It is a by-product of ammonia plant.

Most common way of removing CO₂ is absorption of CO₂ using Benfield solution. Benfield solution is a solution of 29% K₂CO₃, 2.9% DEA and 0.9% V₂O₅ with balanced water. CO₂ is absorbed from raw syngas to benfield solution in absorption column. Then benfield solution is regenerated in regeneration column at higher temperature. CO₂ is sent to urea plant or for other uses.

Reactions-

Absorption: $\text{CO}_2 + \text{K}_2\text{CO}_3 + \text{H}_2\text{O} \Rightarrow 2\text{KHCO}_3$

Regeneration: $2\text{KHCO}_3 \Rightarrow \text{CO}_2 + \text{K}_2\text{CO}_3 + \text{H}_2\text{O}$

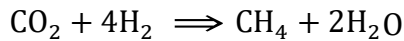
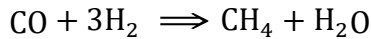
3.Methanation- outlet gas stream of absorption column enters methanator where residual CO and CO₂ are converted to CH₄ by methanation reaction. Operating condition in methanator is-

Temperature- 363-392°C

Pressure- 26 kg/cm² (gage)

Catalyst- Ni

Reactions-



After methanation reaction, ratio of H₂ and N₂ in syngas should be 3:1, which is the feed ratio of ammonia converter.

3.2.3 Ammonia Synthesis and Refrigeration

1. Ammonia synthesis- Purified syngas is fed to ammonia converter after passing series of heat exchangers. Ammonia synthesis reaction occurs in converter.

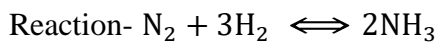
Operating condition of ammonia converter:

Temperature- 400-550°C

Pressure- 150-350 atm

Catalyst- Fe

Catalyst promoters- Al₂O₃, K₂O, MgO, CaO



In Haldor Topsoe process, radial flow converter is used. Mostly used Topsoe converters are- S-50, S-200, S-300, S-350. Maximum conversion that can be achieved in ammonia converter is 25-35%. Unconverted synthesis gas is recycled to the inlet of converter. To avoid inert build up in synthesis loop, a small continuous purge stream is withdrawn.

2. Refrigeration- Outlet of ammonia converter passes a series of heat exchangers, cools down and is flashed in a separator, liquid NH₃ is the bottom product of separator, and it is pump to a let-down vessel and then to the cooler. Then liquid ammonia is stored to sale in cylindrical storage tanks. There are 3 types of ammonia storage. In this plant, atmospheric storage at -33°C will be used.

3. Purge gas recovery unit- Purge gas contains NH₃ and H₂ which can be recover to increase product yield. NH₃ can be recovered by water wash at absorption column and H₂ can be recovered by cryogenic separation, membrane separation or pressure swing adsorption. Purge gas recovery increases NH₃ production to 4-10%.

Chapter 4

DESIGN BASIS

4.1 Geological Data

Geological data including site location, soil characteristics, meteorological information is given below-

4.1.1 Site location

Longitude- 91°50'43.0" E

Latitude- 22°13'39.2" N

Elevation- 7 m

4.1.2 Soil characteristics

Corrosion tendency- Corrosive

Earthquake tendency- zone 3: moderate risk

4.1.3 Meteorological information

Climate type- Tropical monsoon (G. W. Hydrology, 2003)

Maximum precipitation = 650 mm, in the month of July (G. W. Hydrology, 2003)

Average annual rainfall = 2960 mm (G. W. Hydrology, 2003)

Maximum average temperature = 29.4°C

Minimum average temperature = 18.9°C

Table 4.1.1: Meteorological information of Rangadia, Chittagong

Meteorological Parameters	Pre-monsoon (March- May)	Monsoon (June- September)	Post-monsoon (October- November)	Winter (December- February)
Wind Speed (m/s)	6.0 (+1.2%)	5.7 (-3.3%)	2.8 (+0.0%)	2.5 (-0.6%)
Wind direction	SW (29%)	S (33%)	NE (19%)	N (21%)
Air temperature (°C)	29.4 (+4.4%)	28.6 (+4.1%)	24.3 (+7.1%)	18.9 (+10.4%)
Relative humidity (%)	80.4 (-0.4%)	87.8 (-0.6%)	81.4 (-1.1%)	78.0 (+0.6%)
Precipitation (mm/day)	2.73 (-24.8%)	8.43 (+9.1%)	1.79 (+53.7%)	0.38 (-15.0%)

[**Reference:** “Study on Residual Flow in the Bay of Bengal considering Future Climate Change Induced Hydro- meteorological Scenarios.”]

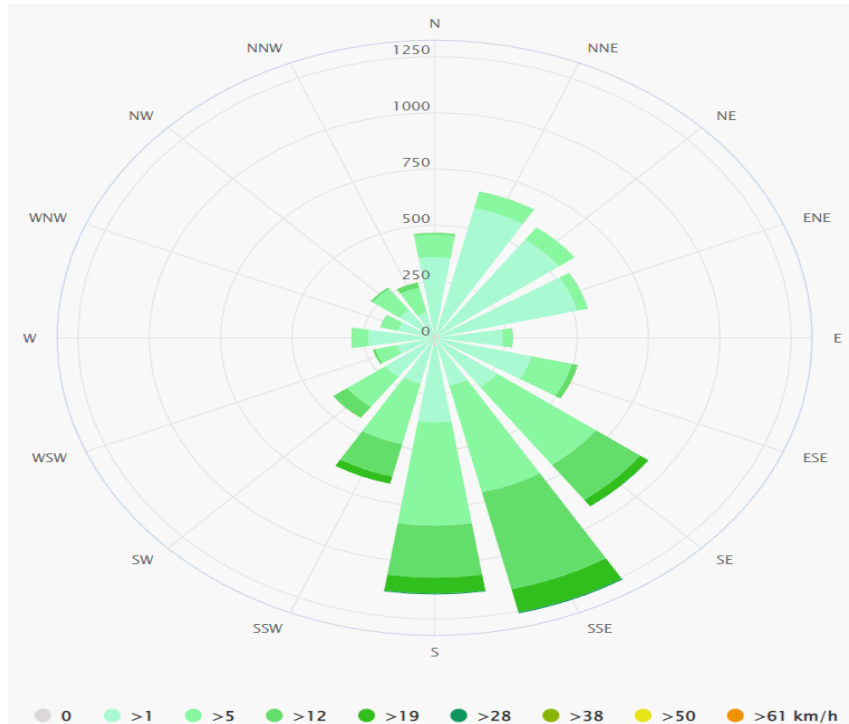


Figure 4.1.1: Wind rose diagram of Chittagong

[Reference:

https://www.meteoblue.com/en/weather/historyclimate/climatemodelled/chittagong_bangladesh_1205733]

4.2 Utility condition

Steam

High pressure steam- 520°C, 412 psig

Medium pressure steam- 354°C, 150 psig

Low pressure steam- 220°C, 45 psig

Air

Ambient air at ambient temperature and 1 atm pressure

Air composition- 78.08% N₂, 20.95% O₂, 0.03% CO₂ and 0.94% Ar

Fuel

Natural gas taken from Karnaphuli Gas Distribution Company

Composition: 97.86% methane, 1.73% ethane, 0.41% propane

Water

Process water- underground water at 22°C (average)

Table 4.2.1: Underground water parameters of Chittagong (M. J. Ahmed *et al.*, 2010)

Parameters	Values	BSTI Standard
pH	7.38	6.4-7.4
Electrical Conductivity	2822.13 $\mu\text{S}/\text{cm}$	---
TDS	787 mg/L	Max. 500 mg/L
DO	1.75 mg/L	Max. 6 mg/L
%NaCl	2.54	---
Acidity	40.55 mg/L	---
Total alkalinity	355.83 mg/L	---
Total hardness	224.59 mg/L	Max. 500 mg/L
Chloride	378.28 mg/L	Max. 600 mg/L
Lead	0.049 mg/L	Max. 0.05 mg/L
Arsenic	0.413 mg/L	Max. 0.05 mg/L
Iron	1.67 mg/L	0.3-1.0 mg/L

Cooling water- from Karnaphuli river, at 25°C (average)

Table 4.2.2: Water parameters of Karnaphuli river (M. J. Ahmed *et al.*, 2010)

Parameters	Values
pH	6.36-9.86
Electrical Conductivity	90-45600 $\mu\text{S}/\text{cm}$
TDS	45-20000 mg/L
DO	0-7.91 mg/L
Transparency	1.5-150 cm
Acidity	1.60-52.25 mg/L
Total alkalinity	5.64-121.00 mg/L
Total hardness	10-4500 mg/L
BOD	0.21-9.17 mg/L
COD	11.39-179.87 mg/L
Chloride	2.09-13147.7 mg/L

Raw materials

1. Natural gas taken from Karnaphuli Gas Distribution Company. Composition of this gas is 97.86% methane, 1.73% ethane, 0.41% propane.
2. Steam generated at process utility using underground water

Power Generation

1110 V -3 phase, 60 Hz

440 V -3 Phase, 60 Hz

220 V -1 Phase, 60 Hz

Source: Steam turbine

Emergency source: Diesel generator

4.3 List of Auxiliaries, Offsites and Package units

1. Water treatment plant
2. Effluent treatment plant
3. Cooling tower
4. Flare
5. Refrigeration unit
6. Boiler
7. Power generation unit
8. Instrument air unit
9. Inert generation unit
10. Fire control unit consisting of fire water pond, fire water pump generator, and others.

Others

- Laboratory
- Maintenance shop

Chapter 5
PROCESS BLOCK DIAGRAM (PBD)
and
PROCESS FLOW DIAGRAM (PFD)

5.1 Process Block Diagram (PBD)

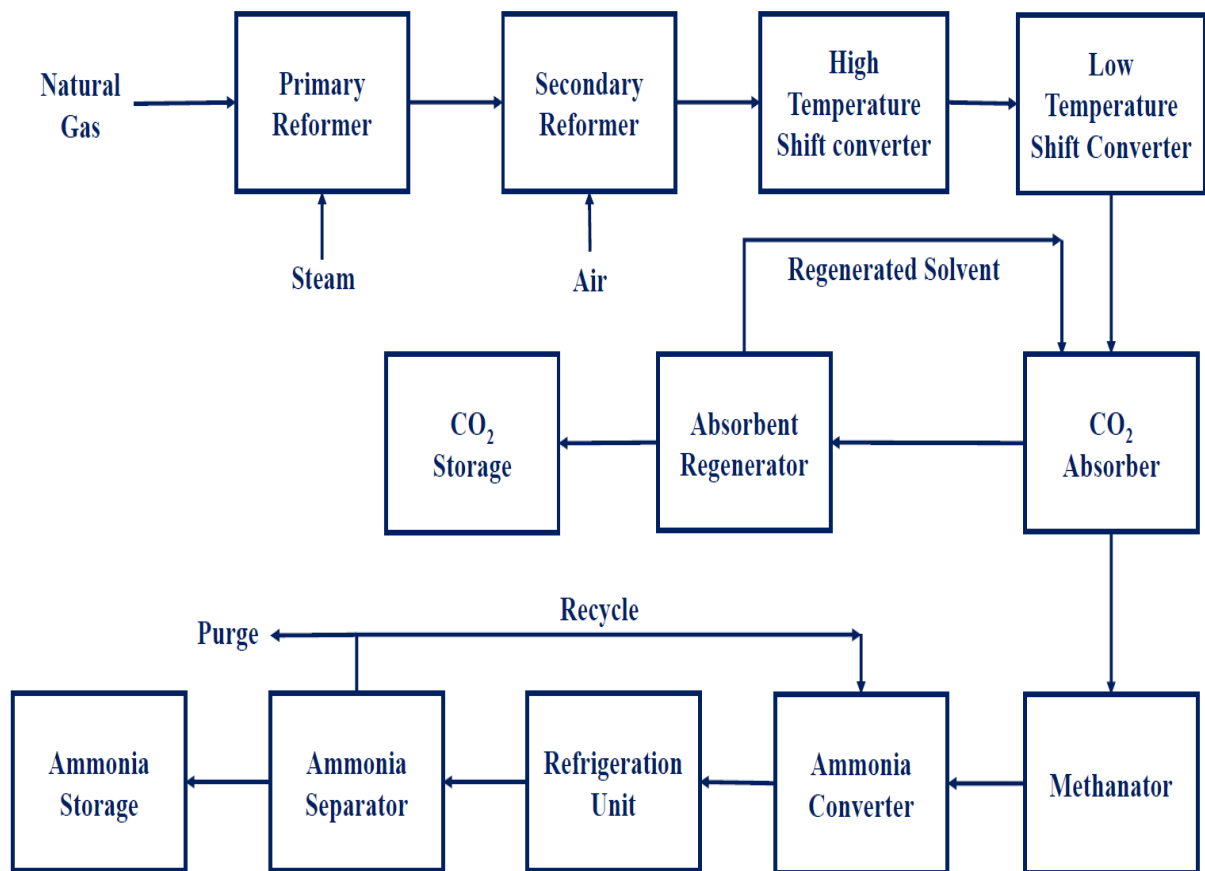


Figure 5.1.1: Process block diagram (PBD) of ammonia plant

5.2 Process Flow Diagram (PFD)

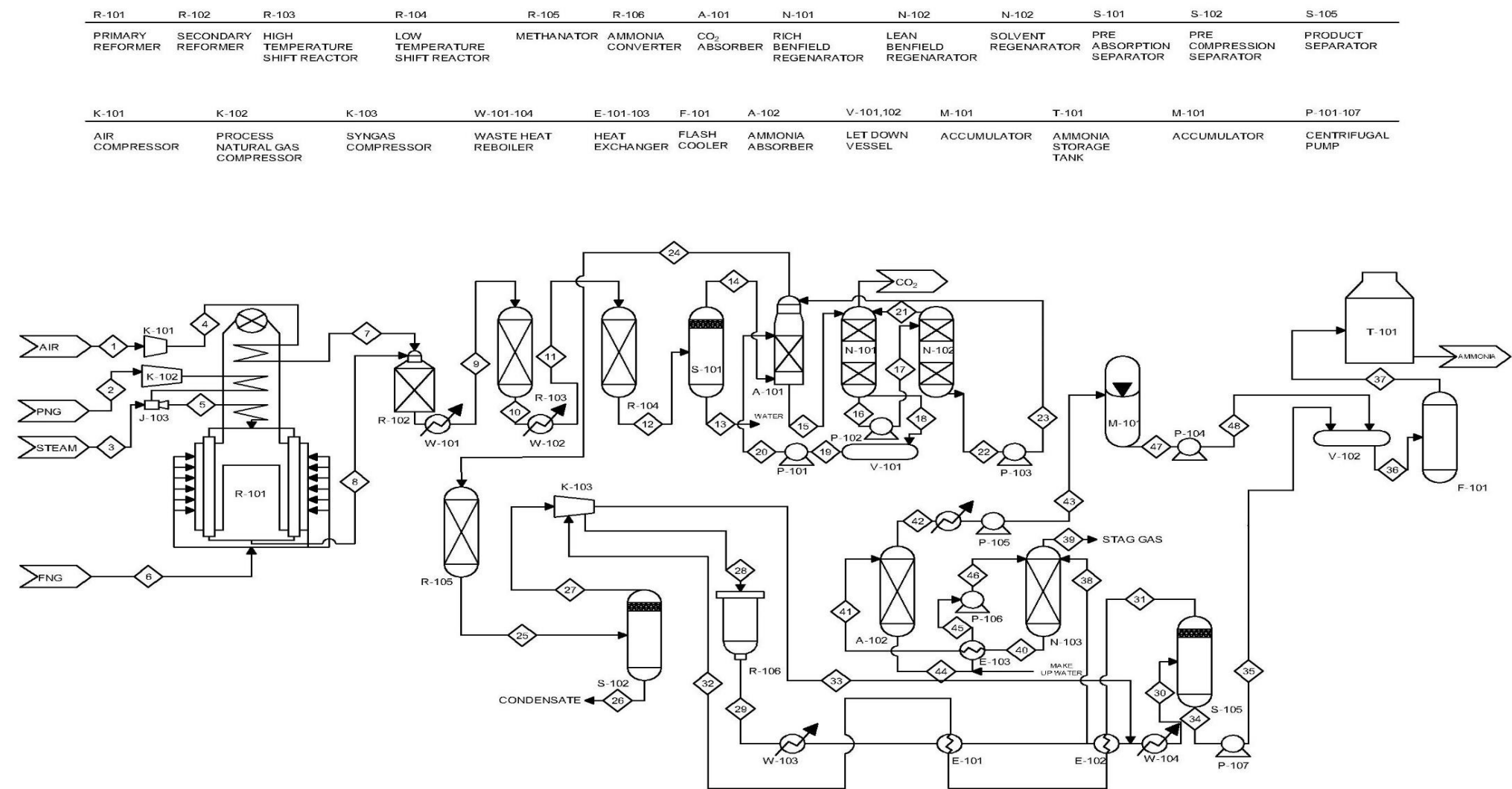
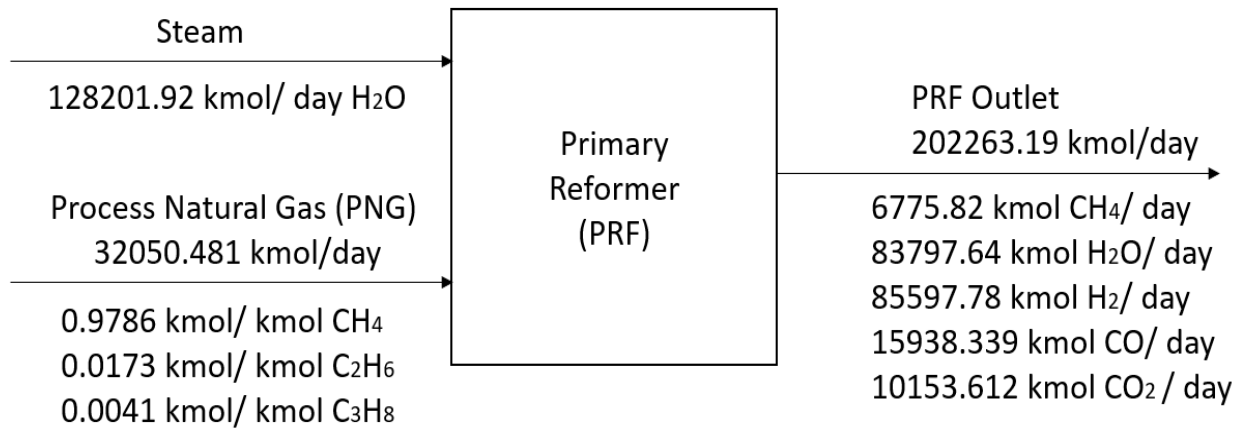


Figure 5.2.1: Process flow diagram (PFD) of ammonia plant

Chapter 6
MATERIAL BALANCE

6.1 Primary Reformer (PRF)



Basis- 32050.48 kmol/day PNG fed to PRF

Reactions-

1. $\text{CH}_4 + \text{H}_2\text{O} \Rightarrow \text{CO} + 3\text{H}_2$
2. $\text{CH}_4 + 2\text{H}_2\text{O} \Rightarrow \text{CO}_2 + 4\text{H}_2$
3. $\text{C}_2\text{H}_6 + 2\text{H}_2\text{O} \Rightarrow 2\text{CO} + 5\text{H}_2$
4. $\text{C}_3\text{H}_8 + 3\text{H}_2\text{O} \Rightarrow 3\text{CO} + 7\text{H}_2$

Assumptions-

- NG:Steam ratio = 1:4
- 100% conversion of hydrocarbons (C₂H₆, C₃H₈)
- PRF outlet composition is 3.35% CH₄, 42.32% H₂, 41.43% H₂O, 5.02% CO₂, 7.88% CO

[**Reference:** Technology and Manufacture of Ammonia by Samuel Strelzoff, Table 2.4, Page: 107]

Material Balance-

Let, PRF outlet flow be X kmol/day

Atomic carbon balance on PRF:

$$32050.481 (0.9786 \times 1 + 0.0173 \times 2 + 0.0041 \times 3) = X \times (0.0335 + 0.0788 + 0.0502)$$

So, X = 202263.19 kmol/day

PRF outlet mole flow-

$$(202263 \times 0.0335) \text{ kmol/day CH}_4 = 6775.82 \text{ kmol/day CH}_4$$

$$(202263 \times 0.4232) \text{ kmol/day H}_2 = 85597.78 \text{ kmol/day H}_2$$

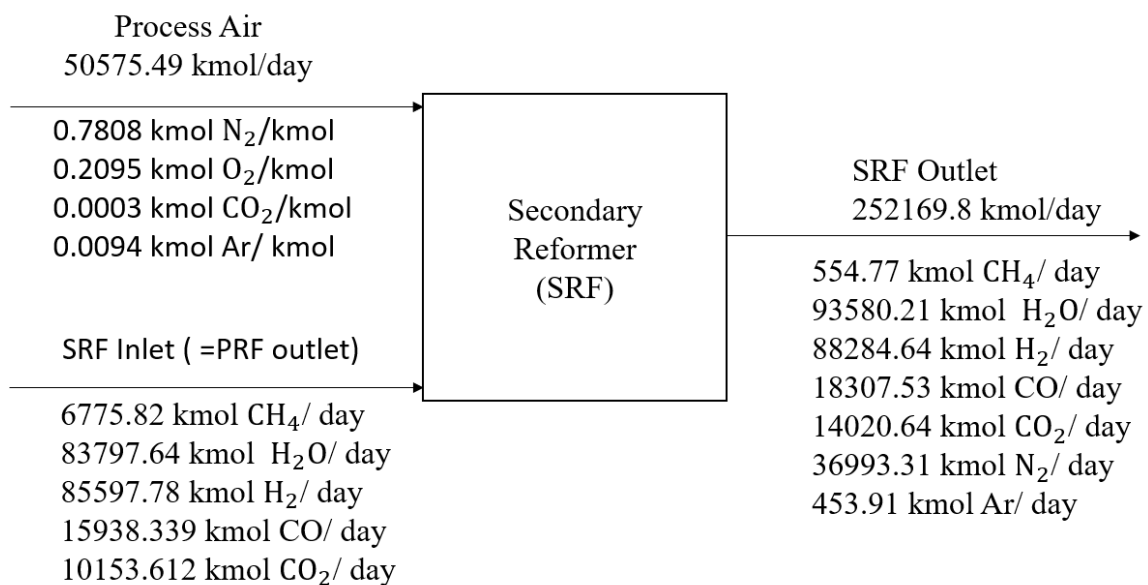
$$(202263 \times 0.4143) \text{ kmol/day H}_2\text{O} = 83797.64 \text{ kmol/day H}_2\text{O}$$

$$(202263 \times 0.0502) \text{ kmol/day CO}_2 = 10153.61 \text{ kmol/day CO}_2$$

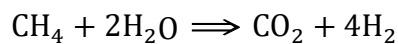
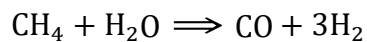
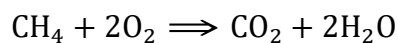
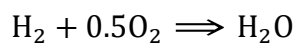
$$(202263 \times 0.0788) \text{ kmol/day CO} = 15938.34 \text{ kmol/day CO}$$

$$\begin{aligned} \therefore \text{CH}_4 \text{ conversion in PRF} &= \frac{32050.48 \times 0.9786 - 6775.82}{32050.48 \times 0.9786} \times 100\% \\ &= 78.38\% \end{aligned}$$

6.2 Secondary Reformer (SRF)



Reactions-



Assumptions-

SRF outlet composition is 0.22% CH₄, 37.11% H₂O, 14.67% N₂, 35.01% H₂, 7.26% CO, 5.56% CO₂, 0.18% Ar

[**Reference:** Modeling and simulation of an industrial secondary reformer reactor in the fertilizer plants, International Journal of Industrial Chemistry, A Springer Open Journal, 2012]

Material Balance-

Let, SRF outlet flow be X kmol/day

Atomic carbon balance on SRF:

$$6775.82 + 15938.339 + 10153.612 = X \times (0.0022 + 0.0726 + 0.0556)$$

$$\therefore X = 252169.8 \text{ kmol/day}$$

SRF outlet mole flow-

$$(252169 \times 0.0022) \text{ kmol/day CH}_4 = 554.78 \text{ kmol/day CH}_4$$

$$(252169 \times 0.3711) \text{ kmol/day H}_2\text{O} = 93580.21 \text{ kmol/day H}_2\text{O}$$

$$(252169 \times 0.1467) \text{ kmol/day N}_2 = 36993.31 \text{ kmol/day N}_2$$

$$(252169 \times 0.3501) \text{ kmol/day H}_2 = 88284.64 \text{ kmol/day H}_2$$

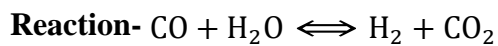
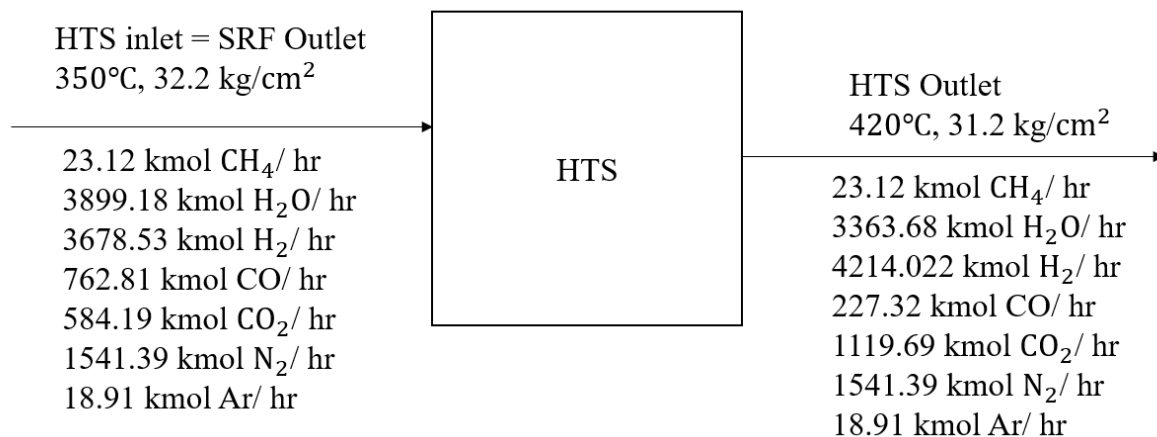
$$(252169 \times 0.0556) \text{ kmol/day CO}_2 = 14020.64 \text{ kmol/day CO}_2$$

$$(252169 \times 0.0726) \text{ kmol/day CO} = 18307.53 \text{ kmol/day CO}$$

$$(252169 \times 0.0018) \text{ kmol/day Ar} = 453.91 \text{ kmol/day Ar}$$

$$\begin{aligned} \therefore \text{CH}_4 \text{ conversion in SRF} &= \frac{6775.82 - 554.78}{6775.82} \times 100\% \\ &= 91.81\% \end{aligned}$$

6.3 High Temperature Shift Reactor (HTS)



Assumptions-

1. 70.2% single pass conversion of CO in the reactor (KAFCO)
2. No side reaction

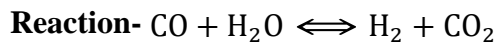
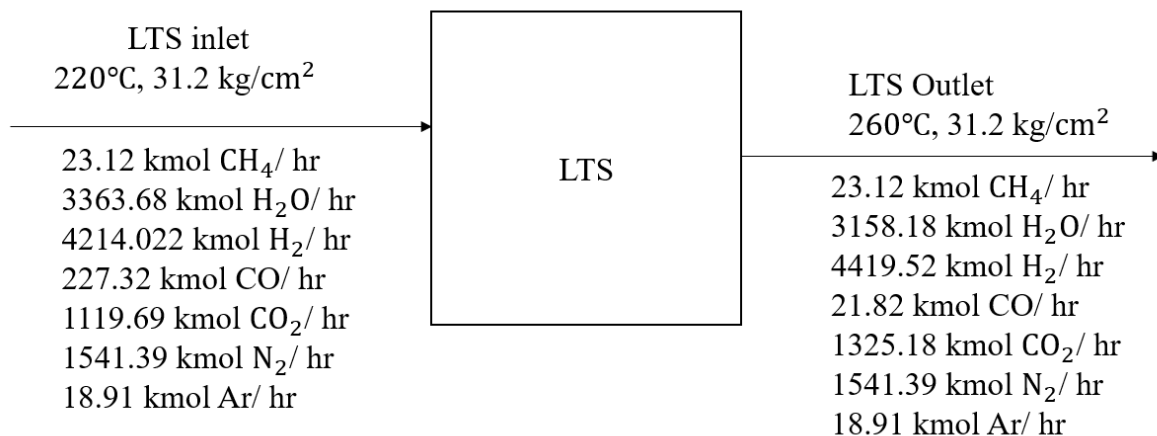
Material Balance-

From material balance around HTS reactor methane, nitrogen and argon is remained same as SRF outlet.

Balance Equations-

1. $X_1 = 584.19 + (0.702 \times 762.81) = 1119.69 \text{ kmol/h}$
2. $X_2 = 762.81 - (0.702 \times 762.81) = 227.32 \text{ kmol/h}$
3. $X_3 = 3678.53 + (.702 \times 762.81) = 4214.022 \text{ kmol/h}$
4. $X_4 = 3899.18 - (0.702 \times 762.81) = 3363.68 \text{ kmol/h}$

6.4 Low Temperature Shift Reactor (LTS)



Assumptions-

1. 90.4 % single pass conversion of CO in the reactor.
2. No side reaction

Material Balance-

From material balance around LTS reactor methane, nitrogen and argon is remained same as LTS outlet.

Balance Equations-

1. $X_1 = 1119.69 + (0.904 \times 227.32) = 1325.184 \text{ kmol/h}$
2. $X_2 = 227.32 - (0.904 \times 227.32) = 21.82 \text{ kmol/h}$
3. $X_3 = 4214.022 + (0.904 \times 227.32) = 4419.518 \text{ kmol/h}$
4. $X_4 = 3363.68 - (0.904 \times 227.32) = 3158.184 \text{ kmol/h}$

6.5 Process Condensate Separator

Condition- 27.7 kg/ cm², 101°C (similar to KAFCO)

Assumption- At process condition all water is separated in the process condensate liquid stream. Because, the boiling point of water at the given pressure is 228°C. (Smith, 1950)

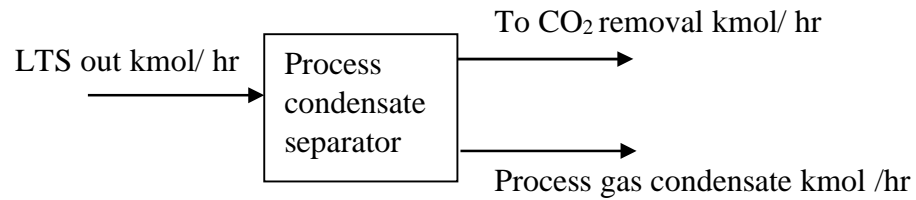
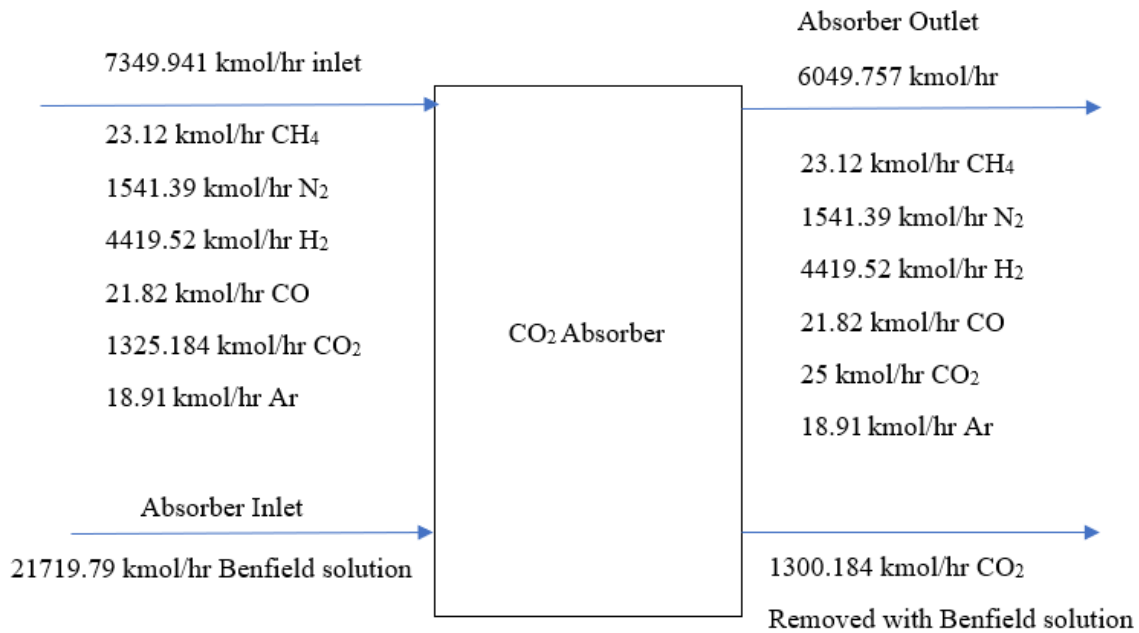


Table 6.5.1: Material balance around process condensate separator

Stream Compositions			
Components	LTS out (kmol/ hr)	To CO ₂ removal (kmol/ hr)	Process gas condensate (kmol/ hr)
CO ₂	1325.184	1325.184	-
CO	21.82	21.82	-
H ₂	4419.52	4419.52	-
CH ₄	23.12	23.12	-
N ₂	1541.39	1541.39	-
Ar	18.91	18.91	-
H ₂ O	3158.18	-	3158.18

6.6 Absorber



Material Balance-

Benfield solution = 30% K₂CO₃

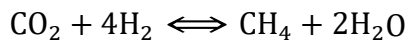
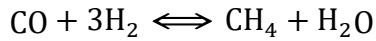
Solubility of CO₂ in Benfield solution taken = 2.81 mol/L

[**Reference:** Technology and Manufacture of Ammonia by Samuel Strelzoff, Table 2.4, Page: 220]

Thus, to dissolve 1300.18 kmol/hr of CO₂, volumetric flow of Benfield solution required = 21719.79 kmol/hr

6.7 Methanator

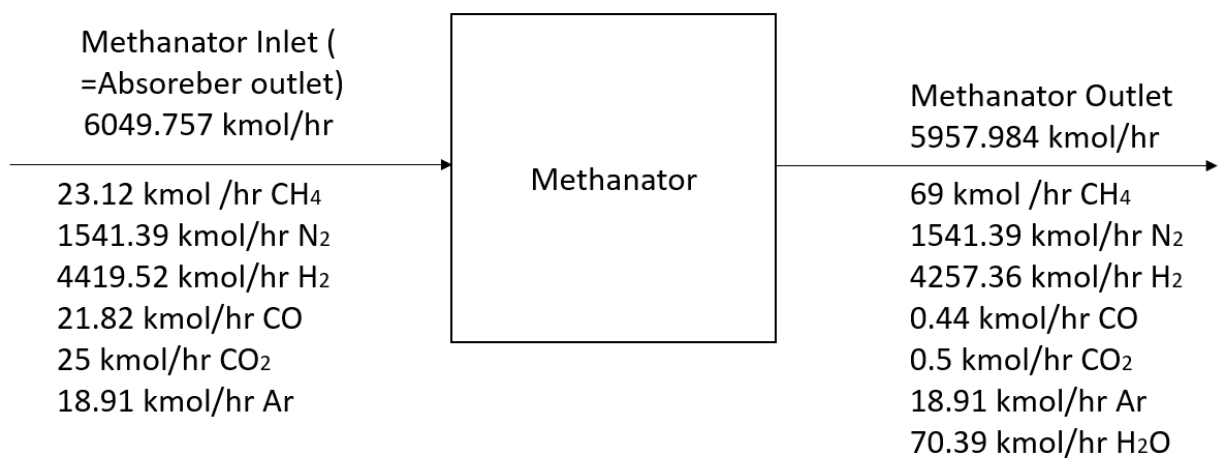
Reactions-



Assumptions-

1. Both reactions have 98 % conversion at vessel condition.
2. Reversible reactions.

Reactor Temperature and Pressure- 300°C and 26 kg/cm² (gage)



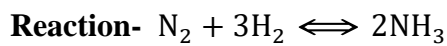
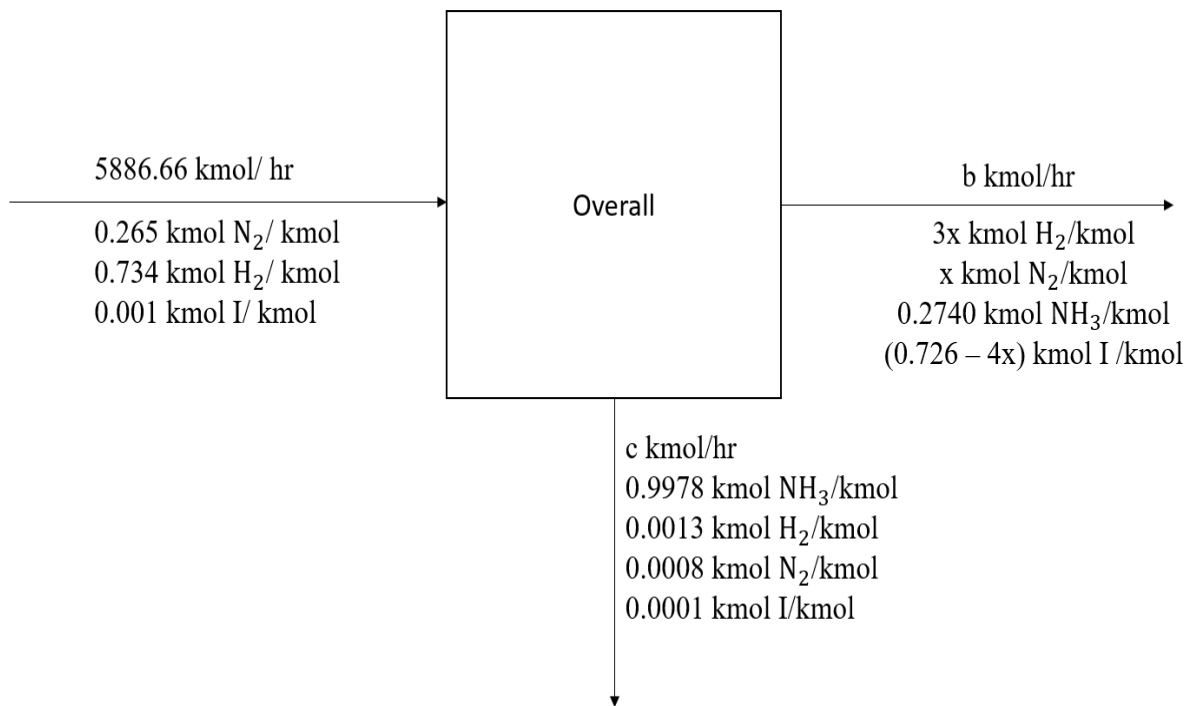
Material Balance-

$$\text{CH}_4 = 23.12 + 21.82 \times 0.98 + 25 \times 0.98 = 69 \text{ kmol/h}$$

$$\text{H}_2 = 4419.52 - (21.82 \times 0.98 \times 3 + 25 \times 0.98 \times 4) = 4257.36 \text{ kmol/h}$$

$$\text{H}_2\text{O} = 21.82 \times 0.98 + 25 \times 2 \times 0.98 = 70.38 \text{ kmol/h}$$

6.8 Overall Balance



98% N_2 conversion:

$$0.265 \times 5886.66 \times 0.02 = b \times x + c \times 0.0008$$

N_2 balance:

$$2 \times b \times x + 0.274 \times b + 0.9994 \times c = 5886.66 \times 0.265 \times 2$$

Inert Balance

$$5886.66 \times 0.001 = b \times (0.726 - 4x) + c \times 0.0001$$

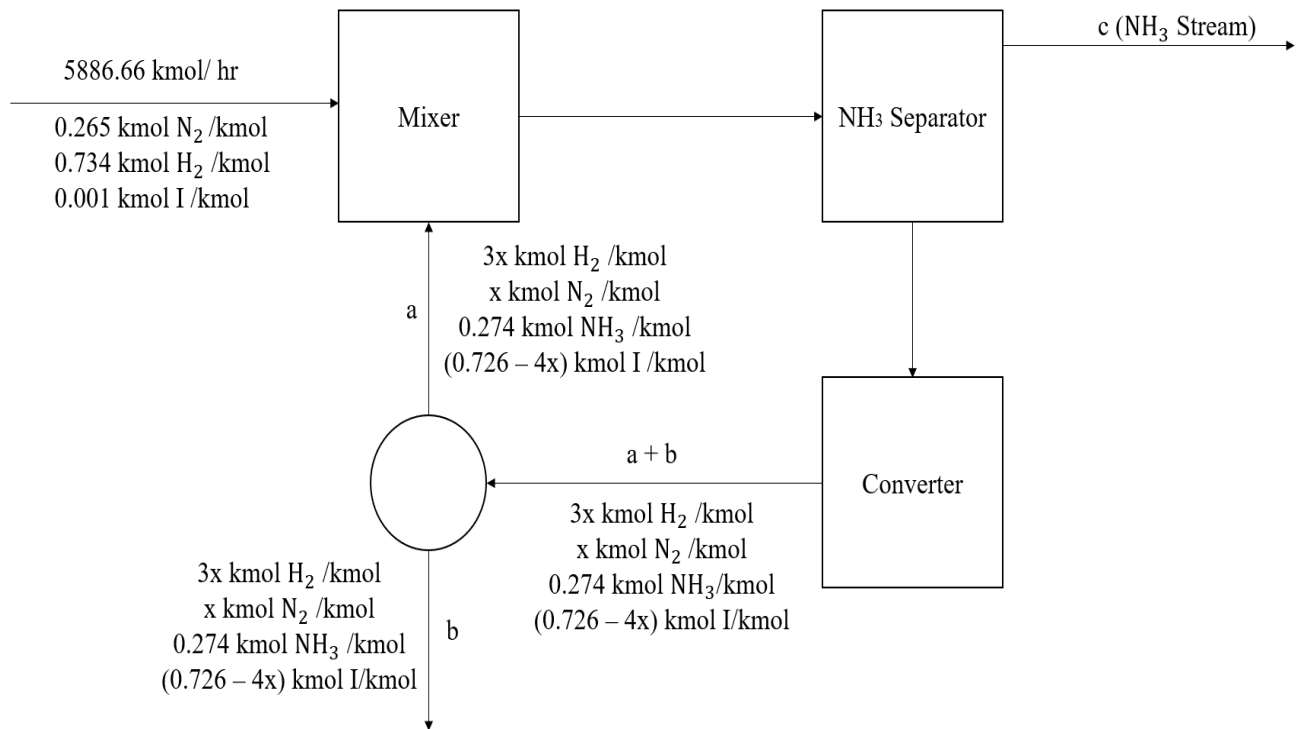
Solving,

$$b = 190.47 \text{ kmol/hr}$$

$$c = 2941.4 \text{ kmol/hr}$$

$$x = 0.1491$$

6.9 Synthesis Loop balance



Concentration of Ammonia (%) in equilibrium with a mixture of 3:1 $H_2/N_2 = 27.4\%$

[**Reference:** Technology and Manufacture of Ammonia by Samuel Strelzoff, Table 2.4, Page: 69, 450°C and 200 atm]

25% once through conversion is assumed

N_2 Conversion:

$$5886.66 \times 0.265 \times 0.75 + 0.1491 \times 0.75 \times a = (a + 190.47) \times 0.1491$$

So, $a = 30625.74 \text{ kmol/hr}$

Scaling up:

Required production rate of ammonia = 1200 tons/ day

$$= (1200 \times 1000) / 17 \text{ kmol/day}$$

$$= 70588 \text{ kmol/day}$$

Current production rate = 2941.4 kmol/hr

$$= (2941.4 \times 24) \text{ kmol/day}$$

$$= 70593 \text{ kmol/day; which is satisfactory}$$

So, there is no need to scale up the production rate.

Chapter 7
ENERGY BALANCE

7.1 Primary Reformer (PRF)

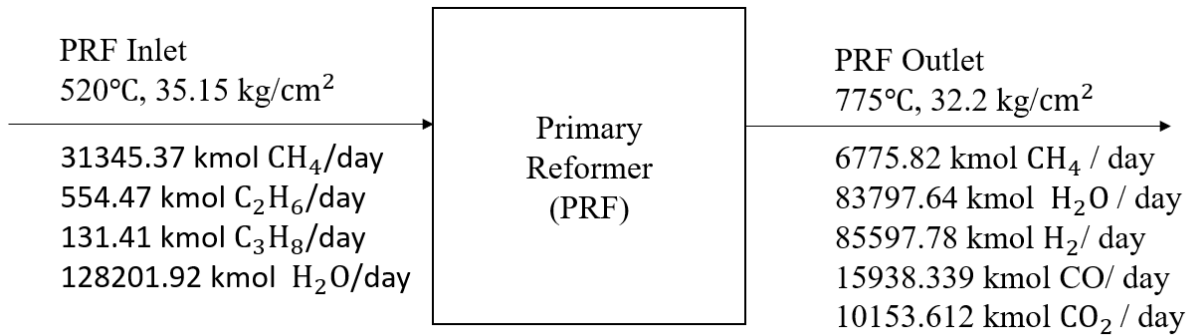


Table 7.1.1: Energy balance around PRF

Reference: C(s), H₂(g), O₂(g), N₂(g) at 25°C and 1 atm

Components	n_{in} (kmol/day)	H_{in} (kJ/kmol)	n_{out} (kmol/day)	H_{out} (kJ/kmol)
CH ₄	31345.37	H ₁ = -50520	6775.82	H ₅ = -33130
C ₂ H ₆	554.47	H ₂ = -44050	-	-
C ₃ H ₈	131.41	H ₃ = -45217	-	-
H ₂ O	128201.92	H ₄ = -224059	83797.64	H ₆ = -213815
H ₂	-	-	85597.78	H ₇ = 22085
CO	-	-	15938.339	H ₈ = -87224
CO ₂	-	-	10153.612	H ₉ = -357025
	$\sum n_{in}H_{in} = -3.034 \times 10^{10}$ kJ/day		$\sum n_{out}H_{out} = -2.127 \times 10^{10}$ kJ/day	

$$\Delta H = \sum n_{out}H_{out} - \sum n_{in}H_{in}$$

$$= 9.07 \times 10^9 \text{ kJ/day}$$

7.2 Secondary Reformer (SRF)

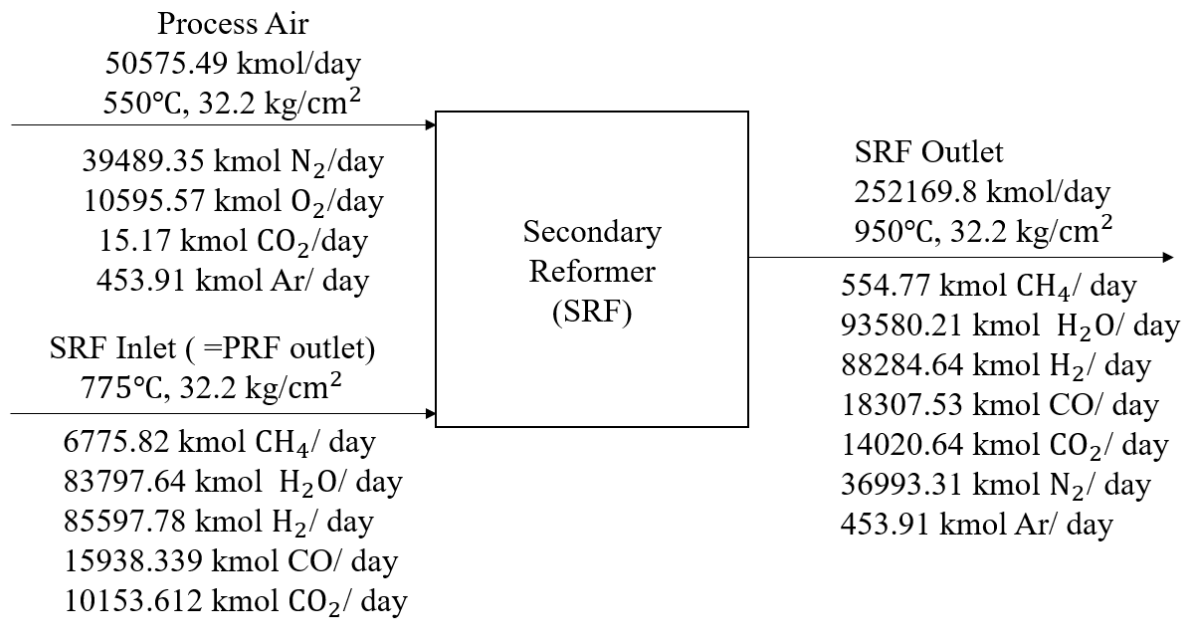


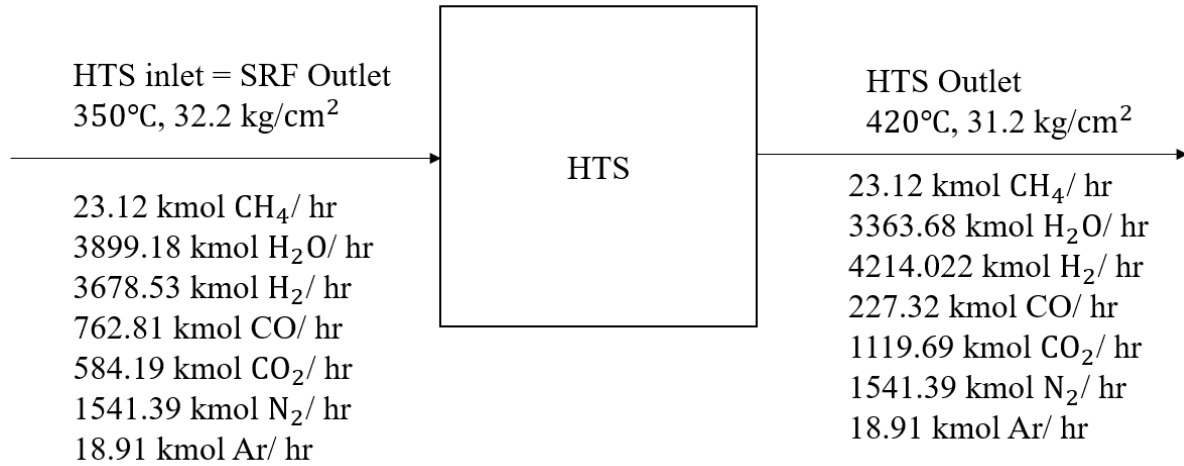
Table 7.2.1: Energy balance around SRF

Reference: C(s), H₂(g), O₂(g), N₂(g), Ar(g) at 25°C and 1 atm

Components	n _{in} (kmol/day)	H _{in} (kJ/kmol)	n _{out} (kmol/day)	H _{out} (kJ/kmol)
CH ₄	6775.82	H ₁ = -33130	554.77	H ₁₀ = -19645
H ₂ O	83797.64	H ₂ = -213,815	93580.21	H ₁₁ = -206,636
H ₂	85597.78	H ₃ = 22085	88284.64	H ₁₂ = 27474
CO	15938.34	H ₄ = -87224	18307.53	H ₁₃ = -81326
CO ₂	10153.61	H ₅ = -357025	14020.64	H ₁₄ = -347,741
O ₂	10595.57	H ₆ = 16718	-	-
N ₂	39489.35	H ₇ = 15802	36993.31	H ₁₅ = 28868
CO ₂	15.17	H ₈ = -369,583	-	-
Ar	453.91	H ₉ = 10926	453.91	H ₁₆ = 19240
	Σ n _{in} H _{in} = -2.047 × 10 ¹⁰ kJ/day		Σ n _{out} H _{out} = -2.221 × 10 ¹⁰ kJ/day	

$$\Delta H = \sum n_{\text{out}} H_{\text{out}} - \sum n_{\text{in}} H_{\text{in}} = -1.74 \times 10^9 \text{ kJ/day}$$

7.3 High Temperature Shift Reactor (HTS)



Reaction-

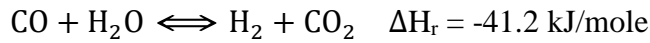


Table 7.3.1: Energy balance around HTS reactor

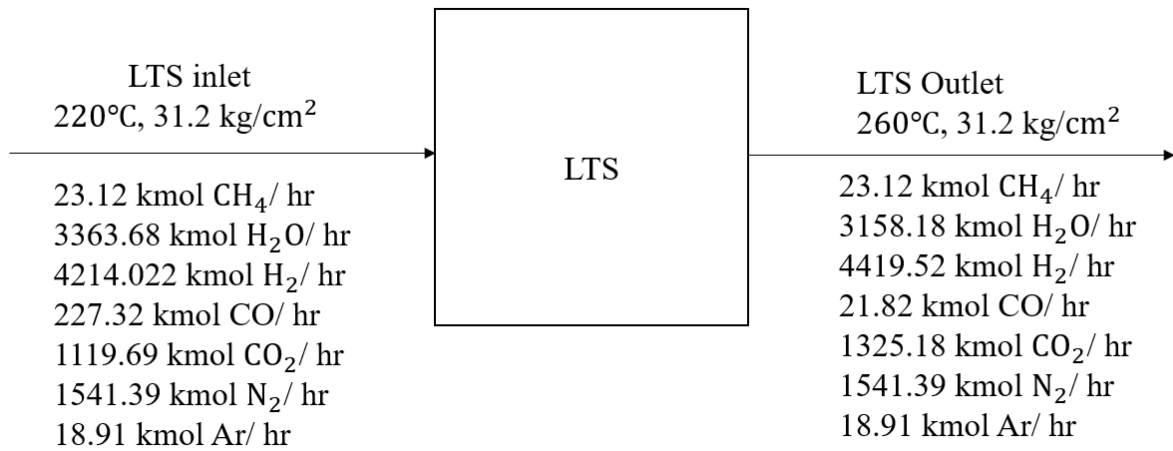
Reference: C(s), H₂(g), O₂(g), N₂(g), Ar(g) at 25°C and 1 atm

Components	n_{in} (kmol/hr)	H_{in} (kJ/kmol)	n_{out} (kmol/hr)	H_{out} (kJ/kmol)
CH ₄	23.12	$H_1 = 14494$	23.12	$H_8 = 18363$
H ₂ O	3899.18	$H_2 = 11389$	3363.68	$H_9 = 13981$
H ₂	3678.53	$H_3 = 9421.37$	4214.022	$H_{10} = 11473$
CO	762.81	$H_4 = 9701$	227.32	$H_{11} = 11867$
CO ₂	584.19	$H_5 = 13931$	1119.69	$H_{12} = 17329$
N ₂	1541.39	$H_6 = 9630$	1541.39	$H_{13} = 11767$
Ar	18.91	$H_7 = 6757$	18.91	$H_{14} = 8212$
	$\sum n_{\text{in}} H_{\text{in}} = 1.099 \times 10^8 \text{ kJ/hr}$		$\sum n_{\text{out}} H_{\text{out}} = 1.362 \times 10^8 \text{ kJ/hr}$	

$$\text{Extent of reaction, } \xi = \frac{762.81 - 227.32}{1} = 535.49$$

$$\text{So, } \Delta H = \xi \times \Delta H_r + \sum n_{\text{out}} H_{\text{out}} - \sum n_{\text{in}} H_{\text{in}} = 4.238 \times 10^6 \text{ kJ/h}$$

7.4 Low Temperature Shift Reactor (LTS)



Reaction-



Table 7.4.1: Energy balance around LTS reactor

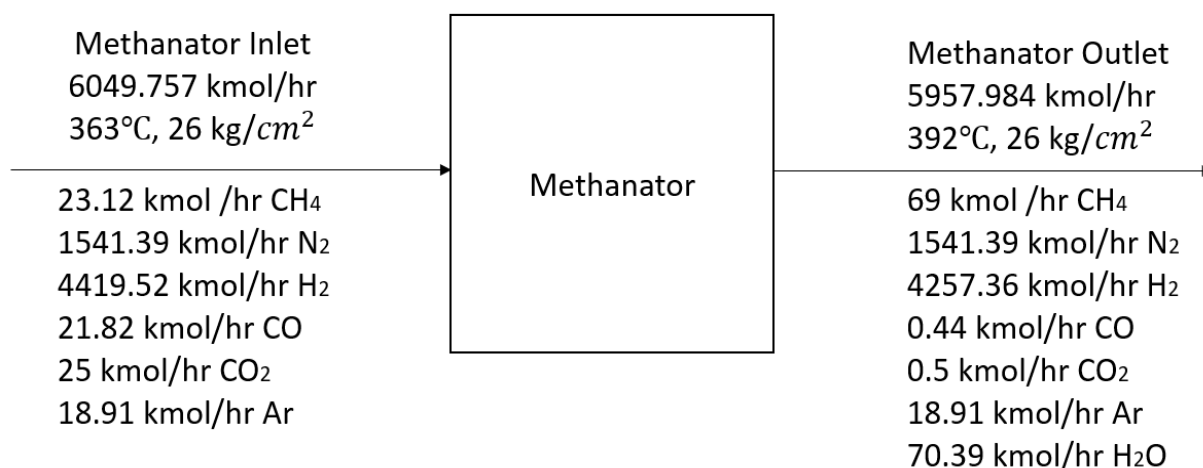
Reference: C(s), H₂(g), O₂(g), N₂(g), Ar(g) at 25°C and 1 atm

Components	\dot{n}_{in} (kmol/hr)	\dot{H}_{in} (kJ/kmol)	\dot{n}_{out} (kmol/hr)	\dot{H}_{out} (kJ/kmol)
CH ₄	23.12	$H_1 = 8003$	23.12	$H_8 = 9903$
H ₂ O	3363.68	$H_2 = 6713$	3158.18	$H_9 = 8134$
H ₂	4214.02	$H_3 = 5636$	4419.52	$H_{10} = 6798$
CO	227.32	$H_4 = 5754.7$	21.82	$H_{11} = 6959$
CO ₂	1119.69	$H_5 = 7955$	1325.18	$H_{12} = 9743$
N ₂	1541.39	$H_6 = 5726$	1541.39	$H_{13} = 6918$
Ar	18.91	$H_7 = 4054$	18.91	$H_{14} = 4886$
	$\sum \dot{n}_{\text{in}} \dot{H}_{\text{in}} = 6.563 \times 10^7 \text{ kJ/hr}$		$\sum \dot{n}_{\text{out}} \dot{H}_{\text{out}} = 7.978 \times 10^7 \text{ kJ/hr}$	

$$\text{Extent of reaction, } \xi = \frac{227.32 - 21.82}{1} = 205.5$$

$$\text{So, } \Delta H = \xi \times \Delta H_r + \sum \dot{n}_{\text{out}} \dot{H}_{\text{out}} - \sum \dot{n}_{\text{in}} \dot{H}_{\text{in}} = 5.683 \times 10^6 \text{ kJ/hr}$$

7.5 Methanator



Reactions-

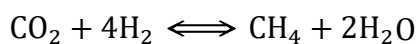
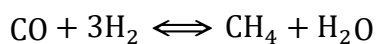


Table 7.5.1: Energy balance around methanator

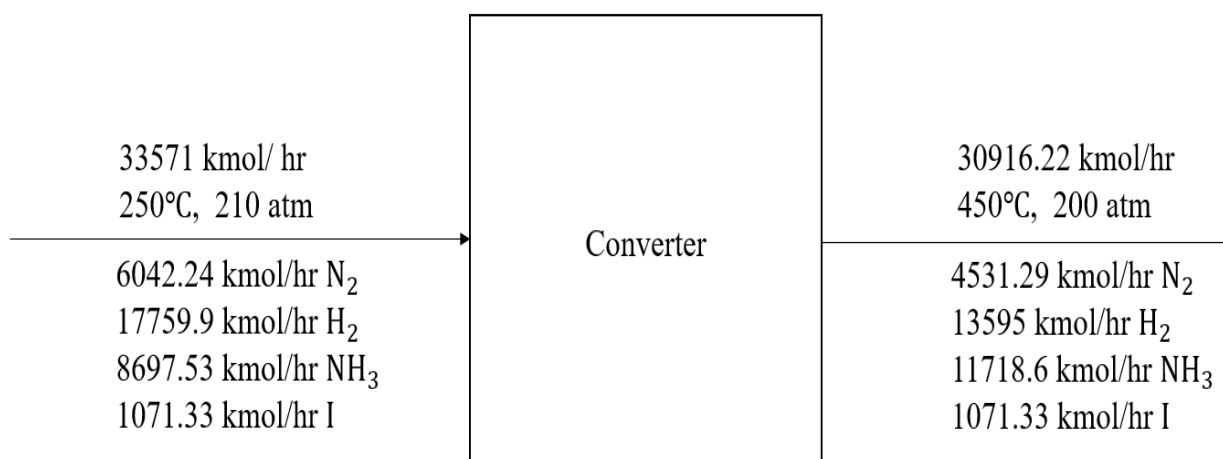
Reference: C(s), H₂(g), O₂(g), N₂(g), Ar(g) at 25°C and 1 atm

Components	n _{in} (kmol/hr)	H _{in} (kJ/kmol)	n _{out} (kmol/hr)	H _{out} (kJ/kmol)
CH ₄	23.12	H ₁ = -59656	69	H ₇ = -58064
H ₂ O	-	-	70.39	H ₈ = -228892
H ₂	4419.52	H ₂ = 9801	4257.36	H ₉ = 10651
CO	21.82	H ₃ = -100418	0.44	H ₁₀ = -99522
CO ₂	25	H ₄ = -378947	0.5	H ₁₁ = -377544
N ₂	1541.39	H ₅ = 10024	1541.39	H ₁₂ = 10909
Ar	18.91	H ₆ = 7027	18.91	H ₁₃ = 7630
	Σ n _{in} H _{in} = 4.586 × 10 ⁷ kJ/hr		Σ n _{out} H _{out} = 4.195 × 10 ⁷ kJ/hr	

$$\Delta H = \sum n_{\text{out}} H_{\text{out}} - \sum n_{\text{in}} H_{\text{in}}$$

$$= -3.91 \times 10^6 \text{ kJ/hr}$$

7.6 Ammonia Converter



Reaction-



Table 7.6.1: Energy balance around ammonia converter

Reference: C(s), H₂(g), O₂(g), N₂(g), Ar(g) at 25°C and 1 atm

Components	n_{in} (kmol/hr)	H_{in} (kJ/kmol)	n_{out} (kmol/hr)	H_{out} (kJ/kmol)
N ₂	6042.24	H ₁ = 6620	4531.29	H ₅ = 12691
H ₂	17759.9	H ₂ = 6507	13595	H ₆ = 12355
NH ₃	8697.53	H ₃ = 8839	11718.6	H ₇ = 17986
I	1071.33	H ₄ = 9420	1071.33	H ₈ = 20100
	$\sum n_{\text{in}} H_{\text{in}} = 2.425 \times 10^8 \text{ kJ/hr}$		$\sum n_{\text{out}} H_{\text{out}} = 4.578 \times 10^8 \text{ kJ/hr}$	

$$\text{Extent of reaction, } \xi = \frac{6042.24 - 4531.68}{1} = 1510.56$$

$$\begin{aligned} \text{So, } \Delta H &= \xi \times \Delta H_r + \sum n_{\text{out}} H_{\text{out}} - \sum n_{\text{in}} H_{\text{in}} \\ &= 7.57 \times 10^7 \text{ kJ/hr} \end{aligned}$$

Chapter 8

UTILITY BALANCE

8.1 Water requirement

Water is required in waste heat boilers to produce steam. These required amounts can be calculated using Aspen hysys V10 after simulating those boilers. The volumetric flow rate of water for 4 waste heat boilers are 871.8 m³/h, 97.21 m³/h, 1237 m³/h and 25.44 m³/h.

$$\therefore \text{Water required for the process} = (871.8 + 97.21 + 1237 + 25.44) = 2231.45 \text{ m}^3/\text{h} \\ = 53554.8 \text{ m}^3/\text{day}$$

8.2 Fuel Natural Gas (FNG) requirement

From energy balance around primary reformer, heat duty = 9.07×10^9 kJ/day

In a conventional PRF, 60% of total heat is used for reactions.

[Reference: Appl. M. 'Ammonia- Principles and Industrial Practice', WILEY-VCH]

$$\therefore \text{Total heat duty} = \frac{9.07 \times 10^9}{0.6} = 1.5 \times 10^{10} \text{ kJ/day}$$

Composition of fuel natural gas = 97.86% methane, 1.73% ethane, 0.41% propane

Using NIST webbook,

Lower heating Value of CH₄ = 802.3 kJ/mol

Lower Heating Value of C₂H₆ = 1425.3 kJ/mol

Lower Heating Value of C₃H₈ = 2039.5 kJ/mol

$$\text{Heating value of fuel} = (802.3 \times 0.9786) + (1425.3 \times 0.0173) + (2039.5 \times 0.0041) \\ = 818.15 \text{ kJ/mol}$$

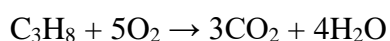
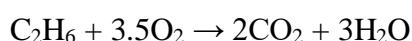
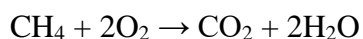
$$\therefore \text{Required amount of fuel} = \frac{1.5 \times 10^{10}}{818.15 \times 1000} = 18334 \text{ kmol/day}$$

$$\therefore \text{Total requirement of natural gas} = \text{PNG requirement in primary reformer} \\ + \text{FNG requirement in primary reformer} \\ = 32050.48 + 18334 \text{ kmol/day} = 50384.48 \text{ kmol/day}$$

8.3 Air requirement

Air is required in primary reformer furnace for combustion and in secondary reformer as raw material.

Complete combustion reactions involved in primary reformer furnace are,



$$\therefore \text{Theoretical air requirement} = \frac{18334 \times (0.9786 \times 2 + 0.0173 \times 3.5 + 0.0041 \times 5)}{0.21} = 177949 \text{ kmol/day}$$

Assuming 40% excess air [Kern 1997],

$$\text{Air consumed in furnace} = 1.4 \times 177949 = 249128 \text{ kmol/day}$$

$$\begin{aligned}
 \therefore \text{Total air consumption in process} &= \text{air consumed in primary reformer furnace} \\
 &\quad + \text{air consumed in secondary reformer} \\
 &= 249128 + 50575.49 = 299703.49 \text{ kmol/day}
 \end{aligned}$$

Flow rate of flue gas leaving primary reformer stack will be equal to the sum of flow rate of FNG and air required for combustion.

$$\therefore \text{Flow rate of flue gas} = 249128 + 18334 = 267462 \text{ kmol/day}$$

Chapter 9
ENVIRONMENTAL MANAGEMENT PLAN

9.1 Hazard Analysis

Chemicals in ammonia plant-

Raw materials- CH₄, air, steam, C₂H₆ (In natural gas)

Intermediate products- CO, CO₂, H₂

Product- NH₃

Solution- Benfield solution (K₂CO₃, V₂O₅, DEA)

Metal catalysts- Ni, Fe oxides, CuO, ZnO

Table 9.1.1: Chemical components and their TLV, LD50/LC50, potential of fire and explosion hazard

Components	PEL/TLV	LD50/LC50	Potential of fire and explosion hazard
CH ₄	1000 ppm	Not classified	Fire and explosion
C ₂ H ₆	1000 ppm	LC50 Inhalation: 658 mg/l/4h (Rat)	Fire and explosion
Steam	Not applicable	LC50 Inhalation: 820000 ppm/ 4 hr (Rat)	Explosion
CO	25 ppm	LC50 Inhalation: 1880 ppm/ 4 hr (Rat)	Fire and explosion
CO ₂	5000 ppm		Explosion
H ₂	Asphyxiant		Fire and explosion
NH ₃	25 ppm	LC50 Inhalation: 7338 ppm/hr (Rat)	Fire and explosion
K ₂ CO ₃	2 mg/m ³	LD50 Oral: 1870 mg/kg (Rat) LD50 Dermal: >2000 mg/kg (Rabbit) LC50 Inhalation: >4.96 mg/L/4.5 hr (Rat)	–
V ₂ O ₅	0.05 mg/m ³	LD50 Oral: 474.2 mg/kg (Rat) LD50 Dermal: >2500 mg/kg (Rabbit) LC50 Inhalation: >11.09 mg/L/4 hr (Rat)	–
DEA	1 mg/m ³	LC50: >100 mg/L (Fish)	–

Ni	1.5 mg/m ³	LC50: >100 mg/L (Fish)	–
Fe oxides	5 mg/m ³		
CuO	1 mg/m ³	Not classified	–
ZnO	5 mg/m ³	LD50 Oral: 7950 mg/kg (Mouse)	–

9.2 Water pollution and its mitigation plan

As a basic upstream product for chemical fertilizers, synthetic ammonia production is a very important industry which is energy intensive that creates considerable water pollution. The water pollution in ammonia industry can be controlled by listed technology in the table 9.2.1.

Table 9.2.1: Process-integrated water pollution reduction technologies

Technology	Major Effect for water pollution control
Methanolization - Methanation purification technology and Methanolization-Hydrocarbylation purification technology.	Eliminate dilute ammonia liquor discharged in the purification stage of about 3-7 kg NH ₄ -N/tNH ₃ compared with traditional copper washing technology.
Recovering liquid ammonia from dilute ammonia liquor by distillation.	Recover more than 99.6% ammonia in dilute ammonia liquor; all residual wastewater can be reused.
Dilute ammonia liquor concentration and reuse.	Reduce ammonia loss in wastewater and condensate, and increased nitrogen fertilizer yield rate.
Near-zero emission of circulating cooling water (using desalted water as replacement for the circulating cooling system).	Reduce cooling water discharge by 5-25 t/tNH ₃ due to the increased degree of concentration.
Oil/water separation and oil recovery	Remove most oil from oily wastewater.
Alkaline solution-based desulfurization for feed gas and shift gas.	Reduce dilute ammonia liquor discharge by approximately 2-4 kg NH ₄ -N/tNH ₃ compared with ammonia solution-based desulfurization.
Reduce dilute ammonia liquor discharge by approximately 2-4 kg NH ₄ -N/tNH ₃ compared with ammonia solution-based desulfurization.	Reduce generation loads of wastewater by 5-10 t/tNH ₃ COD, by 0.4-1.2 kg/tNH ₃ NH ₄ -N, by 0.15-0.2 kg/tNH ₃ phenol, by 1.5-7.5 g/tNH ₃ and by 5-35 g/tNH ₃ cyanide.

9.3 Air pollution and its mitigation plan

The production of ammonia using natural gas as feedstock generates few environmental problems. Most air emission sources at synthetic ammonia plant comply with environmental regulations without the use of add-on control equipment. The mitigation plan for air pollution in ammonia plant are discussed below.

Primary and secondary reformer: The primary reformer emits the combustion product of nitrogen oxides, carbon monoxide and particulate matter, with nitrogen oxides being emitted in the greatest quantity. In addition, the stripper overhead generated later in the ammonia production process during carbon monoxide shifts is reinjected into the primary reformer stack. Stripper overhead contains the pollutant collected from the gas stream being treated and exits the stripper tower as the waste stream. This stream contains ammonia and methanol. Currently, control devices are not required to control air emission from the primary reformer. However, various methods are available to control nitrogen oxide emissions. These methods are:

Ammonia and Hydrogen Recovery: Hydrogen present in the fuel gas stream increases the flame temperature, which in turn increases the production of thermal nitrogen oxides. A cryogenic or permeation system may be used to recover the hydrogen from the purge gas. While water scrubbing is used to remove ammonia. This method removes a great source of nitrogen oxide emissions from the furnace and increases production.

Post combustion Nitrogen-Oxide control: Two methods available for post combustion nitrogen oxide control include selective catalytic reduction (SCR) and selective non-catalytic reduction (SNCR).

Another air pollutant in the ammonia industry is Carbon dioxide which emission can be controlled by a scrubbing solution either monoethanolamine or hot potassium carbonate solution. The gas vented from the carbon dioxide absorption system is predominantly carbon dioxide (99%) with trace amount of ammonia, carbon monoxide, water and methanol.

Ammonia leaks often occur when the product is being handled. Such leaks happen when the product is being loaded into trucks or railroad cars. Since these leaks occur in identified location the emission can be easily controlled through capture and treatment by wet scrubbers.

9.4 Noise pollution and its mitigation plan

Industrial noise control generally involves the replacement of noise-producing machinery or equipment with quieter alternatives. For example, the noise from an air fan may be reduced by increasing the number of blades or their pitch and decreasing the rotational speed, thus obtaining the same air flow. Industrial noise may also be decreased by interrupting its path; for example, a noisy motor may be covered with insulating material. A method of noise control often used in ammonia industry is providing workers with hearing protection devices. These devices must have enough noise attenuation to protect against the anticipated exposures, but must not interfere with the ability to hear human speech and warning signals in the workplace.

9.5 Solid waste pollution and its mitigation plan

Waste materials generated in ammonia industry are recycled where possible, with a comprehensive recycling program implemented onsite. Due to residual contamination contained in certain materials recycling may not always be feasible. Waste streams generated as a result of the Ammonia Plant uprate have been divided into three categories being:

- 1.** Construction wastes generated during plant improvement and modification activities and include scrap metals, concrete, timber and packaging wastes.
 - 2.** Maintenance wastes resulting from maintenance activities undertaken on ammonia plant and equipment. Maintenance wastes are typically generated periodically, corresponding to planned maintenance periods. Maintenance wastes include spent oils, greases, oil rags, absorbents and process residues include wash water and cooling tower precipitates.
 - 3.** Operational wastes that are generated during normal operation of the ammonia plant and supporting activities. Operational wastes include spent catalyst, effluent and general refuse.
- The Management strategy for these solid wastes are listed in table 9.5.1.

Table 9.5.1: Ammonia plant solid waste management strategy

Description	Mitigation Strategy
Waste Generated During Construction Activities	
Scrap metal	Waste scrap metals are segregated into different disposal bins including for ferrous, copper and aluminium and recycled through a scrap metal merchant.
Timber	Pallets are reused onsite where possible. Excess untreated pallets are transferred to an external party to be recycled as compost.
Concrete	Concrete is tested for the presence of contaminants and if suitable recycled through an external provider.
Asbestos	Disposed of at an appropriately licensed waste facility.
Waste generated during maintenance activities	
Absorbents	Thorough assessment of risks in maintenance work to minimize the likelihood of any loss of containment requiring the use of absorbent materials.
Process precipitates	Process precipitates are assessed in accordance with the Waste Classification Guidelines (DECC, 2008) and disposed at an appropriately licensed facility.
Operation Generated Waste	
Ammonia Plant Catalyst	Recycled or stored onsite.
General refuse	Segregation of recyclable materials to minimize volume of waste disposed to landfill.
Paper and cardboard	Paper and Cardboard is segregated in designated collection bins for recycling.

Chapter 10
EQUIPMENT LIST AND SIZING OF EQUIPMENT

10.1 List of Equipment and Sizing

List of major equipment for this plant and their sizes are given in Table 10.1.1

Table 10.1.1: List of equipment and their sizing

Name of the Equipment	Identification No.	No. of items	Size
Secondary Reformer	R-102	1	440.09 m ³
High Temperature Shift Reactor (HTS)	R-103	1	24.75 m ³
Methanator	R-105	1	11.85 m ³
Centrifugal Pump	P-101-107	7	19.5 kW
Ammonia Storage Tank	T-101	1	3781.82 m ³
Heat Exchanger	W-101-107	7	60.32 m ²
Compressor	K-103-104	2	7731 kW
Pre-absorption Separator	S-101	1	4.81 m ³
Pre-compression Separator	S-102	1	6.2 m ³

Chapter 11
INDIVIDUAL EQUIPMENT SIZING

11.1 Primary reformer design

11.1.1 Calculation of Major Parameters

Assumptions:

1. Space velocity is equal to the calculated space velocity using KAFCO data.
2. Reformer tube diameter and length are equal to KAFCO specification.

KAFCO info:

$$\text{PNG inlet flow rate} = 30827.5 \text{ Nm}^3/\text{h} = \frac{30827.5 \times (273 + 375.38)}{273 \times 34.23} = 2138.94 \text{ m}^3/\text{h}$$

$$\text{Steam inlet flow rate} = 89.49 \text{ t/h} = 4510.22 \text{ kmol/h}$$

$$\text{Steam inlet temperature} = 380^\circ\text{C}$$

$$\text{Steam inlet pressure} = 38 \text{ kg/cm}^2 \text{ G}$$

$$\text{No. of tubes} = 200$$

$$\text{Tube inner diameter} = 5 \text{ inch} = 0.13 \text{ m}$$

$$\text{Tube length} = 40 \text{ ft} = 12.2 \text{ m}$$

$$\text{No. of burners} = 480$$

Calculation of space velocity:

$$\text{Volumetric flow rate of steam} = \frac{4510.22 \times 10^3 \times 8.314 \times (273 + 380)}{39 \times 10^5} = 6278.5 \text{ m}^3/\text{h}$$

$$\text{Total inlet flow rate} = 2138.94 + 6278.5 = 8417.44 \text{ m}^3/\text{h}$$

$$\text{Volume of reactor tubes} = \frac{200 \times 3.14 \times (0.13)^2 \times 12.2}{4} = 32.37 \text{ m}^3$$

$$\therefore \text{Space velocity} = \frac{\text{Volumetric flow rate of PRF inlet}}{\text{Volume of reactor tubes}} = \frac{8417.44}{32.37} = 260 \text{ h}^{-1}$$

Calculation of no. of tubes of PRF:

$$\text{Using the assumption, retention time} = 1/260 = 0.0038 \text{ h}$$

$$\text{Total molar inlet flow rate} = 160233.17 \text{ kmol/day} = 6676.38 \text{ kmol/h}$$

$$\text{Inlet temperature} = 520^\circ\text{C}$$

$$\text{Inlet pressure} = 35.15 \text{ kg/cm}^2 \text{ G}$$

$$\therefore \text{Volumetric flow rate} = \frac{6676.38 \times 1000 \times 8.314 \times (520 + 273)}{36.15 \times 10^5} = 12176.32 \text{ m}^3/\text{h}$$

$$\text{Total volume of reformer tubes} = 12176.32 \times 0.0038 = 46.27 \text{ m}^3$$

$$\text{Volume of one tube} = \frac{3.14 \times (0.13)^2 \times 12.2}{4} = 0.162 \text{ m}^3$$

$$\therefore \text{No. of tubes} = \frac{46.27}{0.162} = 286$$

Catalyst specification:

Catalyst- Ni (around 14%-16%)

Base- CaAl_2O_4

Shape- Ring

Dimension (OD×ID×H)- 17 mm× 16 mm× 16 mm

(Reference: <https://www.thermaldynamix.com/nickel-catalysts>)

Bulk density – 40 lb/ft³

Catalyst Charged- 7 m³ Ni and 39 m³ CaAl_2O_4

[Reference: Fertilizer Manual]

Calculation of pressure drop:

According to Ergun equation,

$$\frac{\Delta P}{L} = \frac{150\mu}{D_e^2} \frac{(1-\varepsilon)^2}{\varepsilon^3} v_s + \frac{1.75\rho v_s^2}{D_e} \frac{(1-\varepsilon)}{\varepsilon^3}$$

Where,

ΔP = Pressure drop across packed tube

L = Height of tube

D_e = Equivalent spherical diameter of particle

ρ = Density of fluid

μ = Dynamic viscosity of fluid

v_s = Superficial velocity

ε = Void fraction

$$\begin{aligned} \text{Surface area of catalyst} &= 2\pi \times \left(\frac{17}{2}\right) \times 16 + 2\pi \times \left(\frac{16}{2}\right) \times 16 + 2\pi \times ((17)^2 - (16)^2) \\ &= 1865.16 \text{ mm}^2 \end{aligned}$$

For calculating equivalent diameter,

$$4\pi \times \frac{D_e^2}{4} = 1865.16$$

$$\text{or, } D_e = 24.37 \text{ mm} = 0.024 \text{ m}$$

For PRF inlet, mass density, $\rho = 9.708 \text{ kg/m}^3$, dynamic viscosity, $\mu = 2.373 \times 10^{-5} \text{ Pa.s}$

[Using Aspen Hysys V10 for PRF]

$$\text{Superficial velocity, } v_s = \frac{12176.32 \times 4}{\pi \times (0.13)^2 \times 286} = 3209 \text{ m/h} = 0.89 \text{ m/s}$$

$$\therefore \frac{\Delta P}{L} = \frac{150\mu}{D_e^2} \frac{(1-\varepsilon)^2}{\varepsilon^3} v_s + \frac{1.75\rho v_s^2}{D_e} \frac{(1-\varepsilon)}{\varepsilon^3} = \frac{150 \times 2.373 \times 10^{-5} \times (1-0.4)^2 \times 0.89}{(0.024)^2 \times (0.4)^3} + \frac{1.75 \times 9.708 \times (0.89)^2 \times (1-0.4)}{0.024 \times (0.4)^3}$$

Or, $\Delta P = 5287.57 \times 12.2 = 64508.35 \text{ Pa} = 0.63 \text{ atm}$

Calculation of fuel and air consumption rate:

From energy balance around primary reformer, heat duty = $9.07 \times 10^9 \text{ kJ/day}$

In a conventional PRF, 60% of total heat is used for reactions.

[Reference: Appl. M. 'Ammonia- Principles and Industrial Practice', WILEY-VCH]

$$\therefore \text{Total heat duty} = \frac{9.07 \times 10^9}{0.6} = 1.5 \times 10^{10} \text{ kJ/day}$$

Composition of fuel natural gas = 97.86% methane, 1.73% ethane, 0.41% propane

Using NIST webbook,

Lower heating Value of $\text{CH}_4 = 802.3 \text{ kJ/mol}$

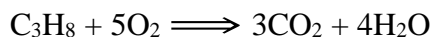
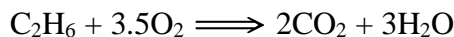
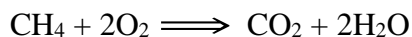
Lower Heating Value of $\text{C}_2\text{H}_6 = 1425.3 \text{ kJ/mol}$

Lower Heating Value of $\text{C}_3\text{H}_8 = 2039.5 \text{ kJ/mol}$

$$\begin{aligned} \text{Heating value of fuel} &= (802.3 \times 0.9786) + (1425.3 \times 0.0173) + (2039.5 \times 0.0041) \\ &= 818.15 \text{ kJ/mol} \end{aligned}$$

$$\text{Required amount of fuel} = \frac{1.5 \times 10^{10}}{818.15 \times 1000} = 18334 \text{ kmol/day}$$

Reactions involved,



$$\text{Theoretical air requirement} = \frac{18334 \times (0.9786 \times 2 + 0.0173 \times 3.5 + 0.0041 \times 5)}{0.21} = 177949 \text{ kmol/day}$$

Assuming 40% excess air [Kern 1997],

$$\text{Air consumed in furnace} = 1.4 \times 177949 = 249128 \text{ kmol/day}$$

$$\text{Air to fuel ratio} = \frac{249128}{18334} = 13.6$$

Mechanical Design:

Calculation of Tube Thickness and Tube pitch:

Tube thickness can be calculated using the equation, $t = \frac{P \times D}{2f - P}$

Where, P = Design pressure = Pressure of the vessel + 10% overdesign

$$\begin{aligned} &= (32.2 + 3.22) \text{ kg/cm}^2 \\ &= 35.42 \text{ kg/cm}^2 \\ &= 3.47 \text{ N/mm}^2 \end{aligned}$$

D = Internal diameter of tube = 0.13 m = 130 mm

f = Design stress factor = 115

$$\therefore \text{Tube thickness} = \frac{3.47 \times 130}{2 \times 115 - 3.47} = 2 \text{ mm}$$

Assuming corrosion allowance = 1 mm, total tube thickness = (2+1) mm = 3 mm

Tube pitch = 1.5 × tube outside diameter = 1.5 × (130 + 3) = 199.5 ≈ 200 mm

Calculation of Primary Stress:

$$\text{Horizontal stress, } \sigma_H = \frac{PD}{2t} = \frac{3.47 \times 130}{2 \times 3} = 75.18 \text{ N/mm}^2$$

$$\text{Longitudinal stress, } \sigma_T = \frac{PD}{4t} = \frac{3.47 \times 130}{4 \times 3} = 37.6 \text{ N/mm}^2$$

Calculation of Weight Loads

Total weight of shell (excluding internal fittings), $W_v = 240 \times t \times C_v \times D_m \times (H_v + 0.8D_m)$

Where,

C_v = a factor to account internal fittings = 1.08

H_v = length of cylindrical section = Height of tube = 12.2 m

t = 3 mm

$D_m = D + t = (130 + 3) \text{ mm} = 133 \text{ mm} = 0.133 \text{ m}$

$$\therefore W_v = 240 \times 3 \times 1.08 \times 0.133 \times (12.2 + 0.8 \times 0.133) = 1272.74 \text{ N} \approx 1273 \text{ N}$$

Dead weight of vessel = W_v + weight of catalyst in one tube

$$\begin{aligned} \text{Mass of catalyst} &= \text{catalyst charged} \times \text{bulk density of catalyst} = (7+39) \times 641.7 \text{ kg} \\ &= 29518.2 \text{ kg} \end{aligned}$$

$$\therefore \text{Weight of catalyst in one tube} = \frac{29518.2}{286} \times 9.8 = 1011.46 \text{ N}$$

$$\therefore \text{Dead weight of vessel} = 1273 + 1011.46 \text{ N} = 2284.46 \approx 2284 \text{ N}$$

Calculation of Stack height:

$$\text{Draft pressure, } P_d = 0.35 \times L_s \times P \times \left(\frac{1}{T_a} - \frac{1}{T_{ga}} \right)$$

Assuming natural type stack,

P_d = draft = 12.7 mm H₂O

P = atmospheric pressure = 1013.25 mbar

T_a = atmospheric temperature = 25°C = 298.15 K

T_{ga} = flue gas temperature = tube temperature = 775°C = 1048.15 K

$$\therefore P_d = 0.35 \times L_s \times 1013.25 \times \left(\frac{1}{298.15} - \frac{1}{1048.15} \right) = 0.851 L_s$$

$$\therefore L_s = \frac{12.7}{0.851} = 14.92 \approx 15 \text{ m}$$

11.1.2 Mechanical drawing of Primary Reformer

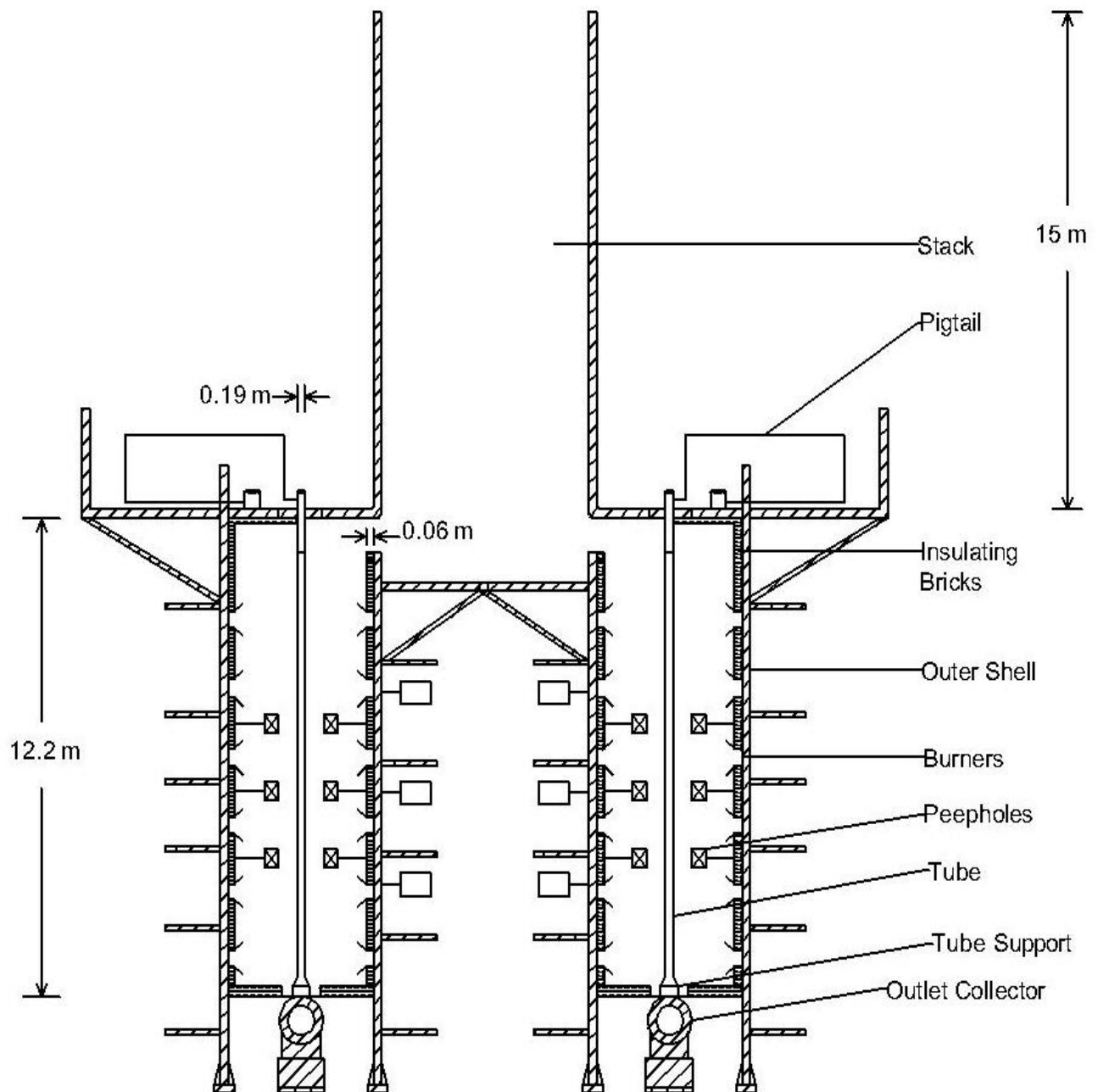


Figure 11.1.1: Mechanical drawing of Primary Reformer

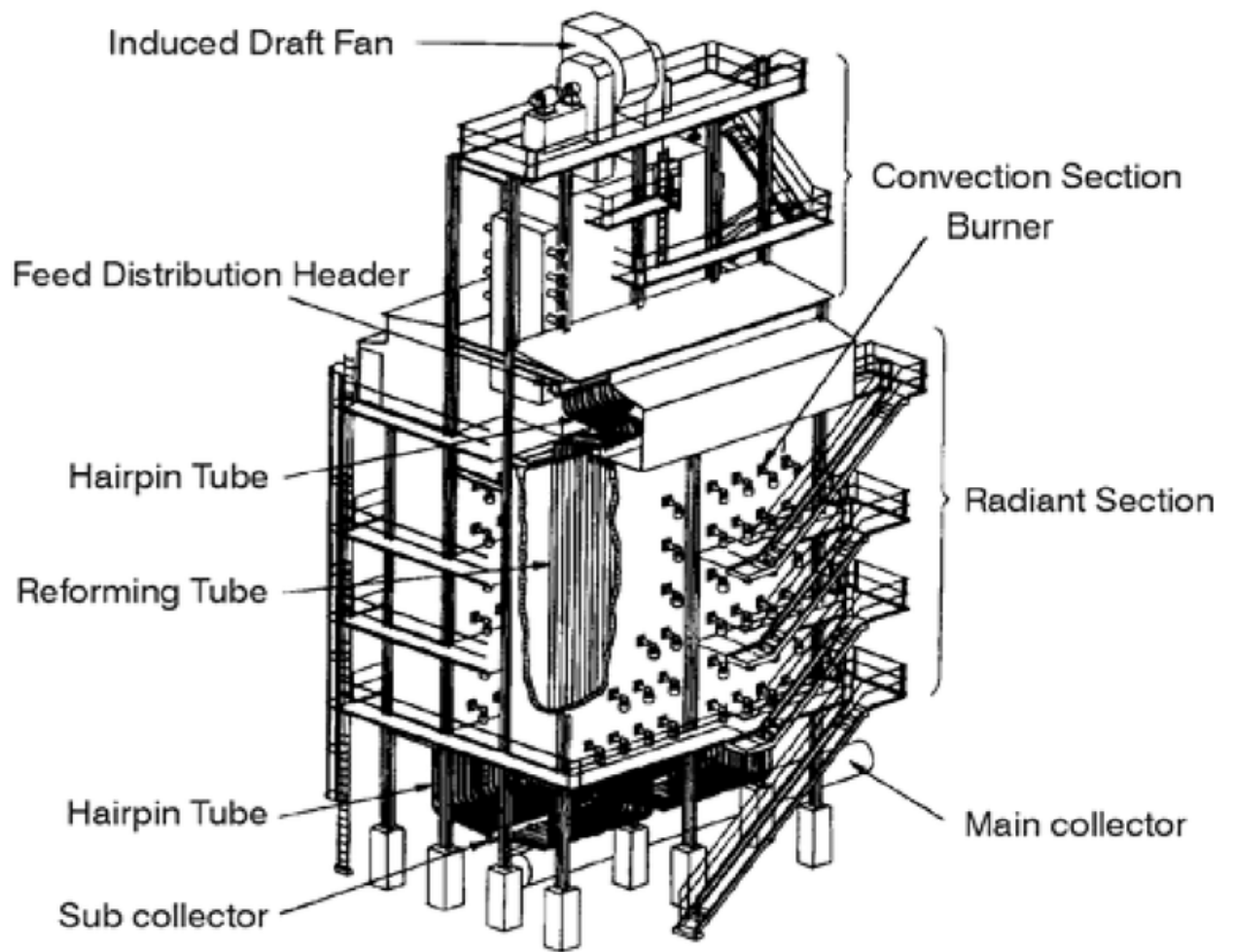


Figure 11.1.2: Schematic Diagram of topsoe side fired primary reformer

[Source:https://www.researchgate.net/figure/Topsoe-reformer-with-burners-placed-on-side-walls-1_fig1_320085255]

11.1.3 Piping & Instrumentation Diagram of Primary Reformer

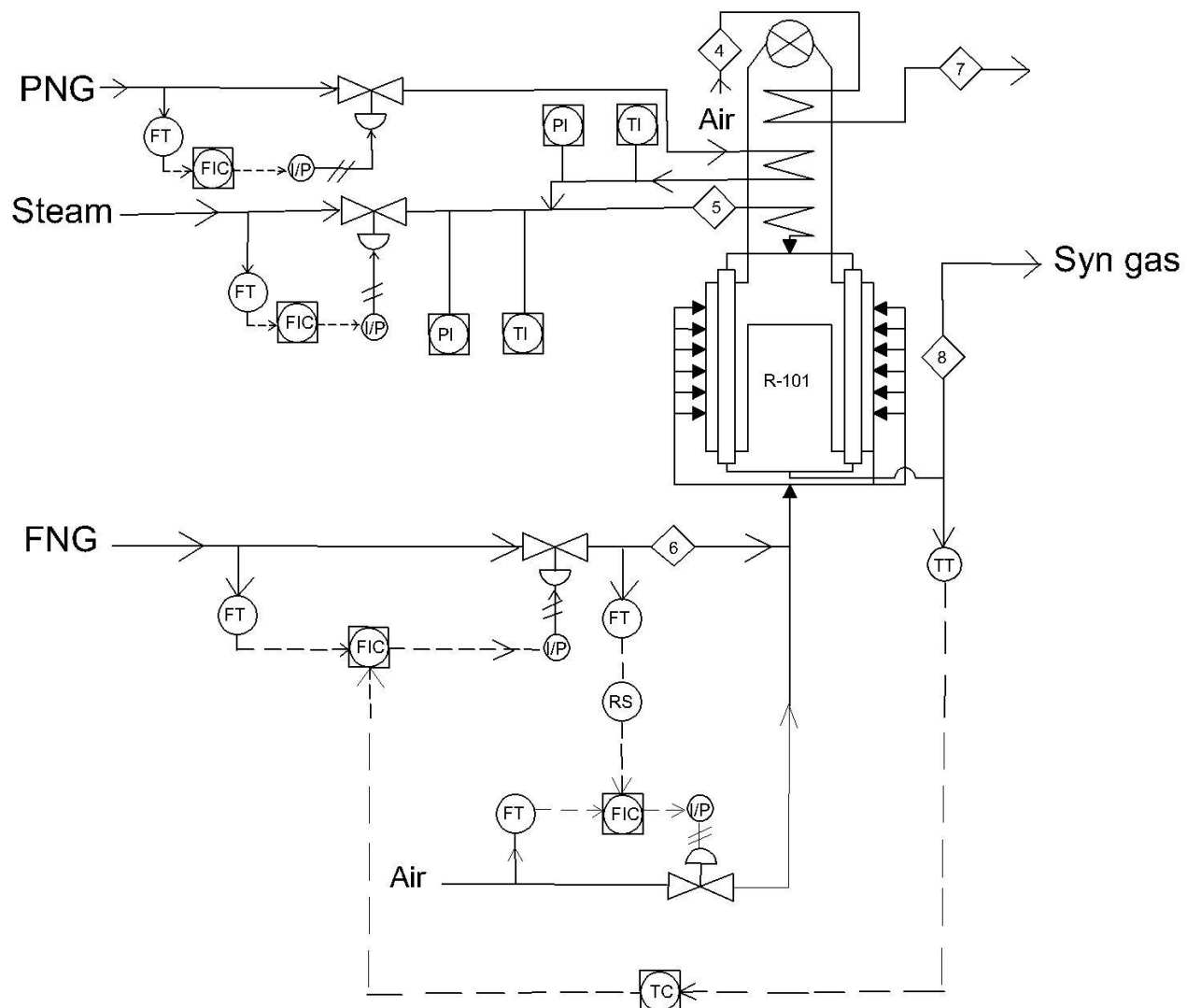


Figure 11.1.3: P & ID of Primary Reformer

11.1.4 Equipment Specification Sheet

Table 11.1.1: Specification Sheet of Primary reformer

Job No.:	Reference No.:
Customer:	Proposal No.:
Address:	Date:
Plant Location: Rangadia, Upozilla: Anowara, Chittagong	Item: Primary Reformer
Service of Unit:	Item No.: R-101
Code: ASME BPVC Section I	
Operation: Continuous	
Function: Combustion of natural gas and steam- methane reforming with necessary input of air, natural gas and steam	
Inlet Condition:	
Temperature	520°C (mixed stream)
Pressure	35.15 kg/cm ² G
Molar flowrate	160233.17 kmol/day
Outlet Condition:	
Temperature	775°C
Pressure	32.2 kg/cm ² G
Molar flowrate	202263 kmol/day
Combustion Conditions:	
Excess air	40%
Total heat duty	1.5×10 ¹⁰ kJ/day
Air to fuel ratio	13.6
Tube Design:	
Tube orientation	Vertical
Design Pressure	35.42 kg/cm ² G
Tube inside diameter	130 mm
Tube length	12.2 m
Tube wall thickness	3 mm
No. of tubes	286
Tube material	SS 310
Tube pitch	200 mm
Catalyst:	
Catalyst	Ni

Base	CaAl ₂ O ₄
Shape	Ring
Dimension (OD×ID×H)	17 mm× 16 mm× 16 mm
Bulk Density	635.66 kg/m ³
Catalyst Charged	7 m ³ Ni and 39 m ³ CaAl ₂ O ₄
Pressure drop in catalyst bed	0.63 atm
Reformer Specification:	
Vessel shell	Carbon steel
Type of burner	Side fired
No. of burners	480
Type of support	Bracket
Weight load of vessel shell	1273 N
Dead weight load of vessel	2284 N
Refractory type	Insulating firebricks
Thickness of firebricks	63.5 mm
Stack Specification:	
Type	Natural
Draft	12.7 mm H ₂ O
Ambient temperature	25°C
Flue gas temperature	775°C
Stack height	15 m

11.2 Low Temperature Shift Reactor

11.2.1 Calculation of Major Parameters

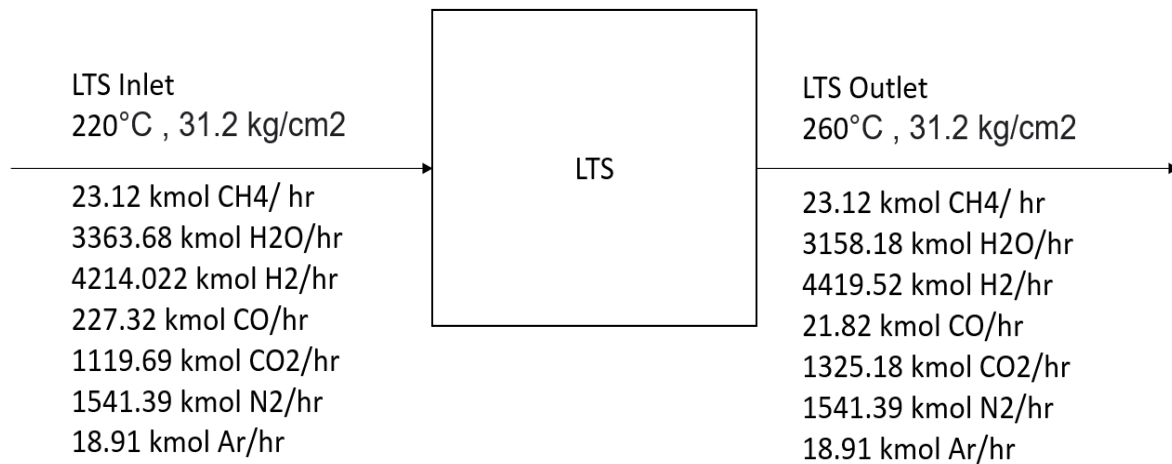


Figure 11.2.1: Inlet and outlet streams of LTS

a. Catalyst Specification:

Catalyst dimensions = 5.2 mm diameter and 3mm length (Katalco 83-3 catalyst)

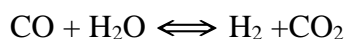
Catalyst bulk density = 1360 kg/m³ = 85 lb/ft³

Catalyst bed void fraction = 0.4

(Reference: 0156fcb2-8724-40c2-7e4f-3a55092fd7ad.Pdf, n.d.)

b. Catalyst Mass Calculation:

For the reaction,



Reaction rate can be expressed as,

$$-r = k\Psi \frac{y_{\text{CO}} \times y_{\text{H}_2\text{O}} - \frac{y_{\text{H}_2} \times y_{\text{CO}_2}}{K_{eq}}}{379 \times \rho_b} \quad (\text{Reference: Case Study 105.Pdf, n.d.})$$

Here,

$-r_{\text{CO}}$ = rate of reaction (lb mol CO reacted/ lb catalyst.hr)

T = Inlet temperature, °R

Equilibrium Constant, $K_{eq} = e^{-4.72 + \frac{8640}{T}}$ for $760 \leq T \leq 1060$

$= e^{-4.33 + \frac{8240}{T}}$ for $1060 \leq T \leq 1360$

Ψ = Activity factor = $0.86 + 0.14P$ for $P \leq 24.8$ atm

$= 4.33$ for $P > 24.8$ atm

$$k = \text{rate constant} = e^{12.88 - \frac{3340}{T}}$$

$$\rho_b = \text{bulk density of catalyst (lb/ft}^3\text{)}$$

Now,

$$T = 887.67 \text{ }^\circ\text{R}$$

$$K_{eq} = e^{-4.72 + \frac{8640}{887.67}} = 150.41$$

$$k = e^{12.88 - \frac{3340}{887.67}} = 9111.94$$

$$\Psi = 4.33 \text{ (P > 24.8 atm)}$$

$$\rho_b = 85 \text{ lb/ft}^3$$

$$y_{CO} = 0.0216, y_{H_2O} = 0.32, y_{CO_2} = 0.1066, y_{H_2} = 0.401$$

Substituting values,

$$-r_{CO} = 8.12 \times 10^{-3} \frac{\text{lb moles CO reacted}}{\text{lb catalyst.hr}}$$

Now,

$$-r_{CO} \left[\frac{\text{lb moles CO reacted}}{\text{lb catalyst.hr}} \right] \times W_c \text{ [lb catalyst required]} = m_{CO} \left[\frac{\text{lb moles CO at inlet}}{\text{hr}} \right]. \quad X_{CO}[\text{CO}$$

conversion]

$$\text{So, } W_c = 55794 \text{ lb catalyst} = 25307 \text{ kg catalyst}$$

c. Bed dimension:

Area Calculation:

$$\text{Molar density at stated condition, } \rho_s = \frac{P}{RT_{avg}} = \frac{3.06 \times 10^6}{8.314 \times (240 + 273)} = 0.7175 \text{ kmol/m}^3$$

$$\text{Volumetric flowrate over the bed, } V_b = \frac{n_{total}}{\rho_s} = 14645.48 \text{ m}^3/\text{hr}$$

$$\text{Assuming, superficial velocity across bed, } u_s = 0.3 \text{ ms}^{-1} \text{ (KAFCO)}$$

So,

$$A_b \times u_s = V_b$$

$$\text{or, } A_b = 13.56 \text{ m}^2$$

$$\text{or, } \frac{\pi \times D_i^2}{4} = 13.56$$

$$\text{So, Inside diameter of bed, } D_i = 4.16 \text{ m}$$

Adopted diameter = 4200mm

Now,

$$\text{Bed void fraction} = 0.4$$

Assuming,

Bed space velocity = 500 hr⁻¹(Reference: *Case Study 105.Pdf*, n.d.)

So,

$$\text{Bed Volume, } V_b = \frac{\text{Volumetric flow rate}}{\text{void fraction} \times \text{space velocity}} = \frac{14645.48}{0.4 \times 500} = 73.23 \text{ m}^3$$

$$\text{So, bed length} = \frac{\text{bed volume}}{\text{bed cross sectional area}} = 5.4 \text{ m}$$

Adopted length = 5400mm

Mechanical Design:

a. Pressure Drop Calculation:

$$\text{Bed length} = 5.4 \text{ m} = 17.72 \text{ ft}$$

$$\text{Bed void fraction, } \epsilon = 0.4$$

$$\text{Catalyst particle volume} = \frac{\pi D^2 h}{4} = \frac{3.14 \times 5.2^2 \times 3}{4} = 63.68 \text{ mm}^3$$

$$\begin{aligned} \text{So, catalyst particle effective diameter, } D_p &= \left(\frac{6 \times \text{Volume of catalyst particle}}{3.14} \right)^{1/3} \\ &= 4.29 \text{ mm} \\ &= 0.1689 \text{ inch} \end{aligned}$$

$$G = 500 \text{ hr}^{-1}$$

$$\text{So, Reynolds number, } N_{Re} = \frac{D_p G}{\mu} = \frac{0.1689 \times 500}{12 \times 2.42 \times 1.8 \times 10^{-2}} = 161.56$$

$$\text{Friction factor, } F_k = 1.75 + 150 \times \frac{1-\epsilon}{N_{Re}} = 2.31$$

$$\text{Feed inlet density, } \rho_f = \frac{P \times M_{avg}}{RT} = 0.746 \text{ lb/ft}^3$$

From Ergun equation,

$$\frac{\Delta P}{\Delta L} = \frac{F_k \times G^2}{D_p \times \rho_f \times g_c} \times \frac{(1-\epsilon)}{\epsilon^3} = 1.24 \text{ psf/ft} = 8.61 \times 10^{-3} \text{ psi/ft}$$

$$\text{So, pressure drop across total bed length} = 8.61 \times 10^{-3} \times 17.72 = 0.1526 \text{ psi}$$

b. Shell thickness calculation:

The material used for the vessel fabrication= ASME A516 Grade 70/ ASTM A516 Grade 70 carbon steel

Thickness of shell can be calculated using the equation,

$$t = \frac{P_i \times D_i}{2SE - 1.2P_i} \quad (\text{Towler \& Sinnott, 2008})$$

$$\text{Inner diameter of vessel, } D_i = 4200 \text{ mm}$$

$$\begin{aligned} \text{Design pressure, } P_i &= \text{total internal pressure} \times \text{safety factor} \\ &= 3060 \text{ kPa} \times 1.5 \\ &= 4590 \text{ kPa} \end{aligned}$$

Design temperature = maximum operating temperature for the catalyst
 = 285°C (Rase, 1977)

Permissible stress at design temperature for the material, S = 20 ksi
 = 138 MPa

Joint efficiency factor, E = 0.85

$$\text{So, thickness, } t = \frac{4590 \times 4200}{2 \times 138 \times 1000 \times 0.85 - 1.2 \times 4590} = 84.15 \text{ mm}$$

Corrosion allowance = 3 mm

$$\text{So, shell thickness} = 84.15 + 3 = 87.15 \text{ mm}$$

Adopted shell thickness = 90 mm

$$\text{So, outside diameter of shell, } D_o = 4200 + 2 \times 90 = 4380 \text{ mm}$$

c. Vessel head thickness calculation:

Vessel head type: Hemispherical head

Material of construction: ASME A516 Grade 70/ ASTM A516 Grade 70 carbon steel

Hemispherical head thickness can be calculated from the equation,

$$\begin{aligned} t_h &= \frac{P_i \times D_i}{4SE - 0.4P_i} \text{ (Towler \& Sinnott, 2008)} \\ &= \frac{4590 \times 4200}{4 \times 138 \times 1000 \times 0.85 - 0.4 \times 4590} \\ &= 41.25 \text{ mm} \end{aligned}$$

Adopted head thickness = 50 mm (including corrosion allowance)

d. Feed Inlet diameter:

Material of construction: ASTM A105, carbon steel

Inlet temperature, T = 220°C = 493K

Inlet pressure, P = 3060 kPa

Feed gas flowrate, n = 10508 kmol/hr

$$\begin{aligned} \text{So, volumetric flowrate, } Q &= \frac{nRT}{P} \\ &= \frac{10508 \times 1000 \times 8.314 \times 493}{3600 \times 3060 \times 1000} \\ &= 3.91 \text{ m}^3/\text{s} \end{aligned}$$

Assuming velocity, $v_n = 11.8 \text{ m/s}$ (KAFCO)

So,

$$\text{Nozzle area, } A_n = \frac{Q}{v_n} = 0.3314 \text{ m}^2$$

$$\text{and nozzle diameter, } d_n = \sqrt{\frac{4 \times 0.3314}{\pi}} = 0.6497 \text{ m} = 649.7 \text{ mm}$$

Adopted nozzle diameter, $d_n = 650$ mm

Nozzle thickness can be determined from,

$$t_n = \frac{P_i \times D_i}{2SE - 1.2P_i} \text{ (ASME Boiler and pressure vessel code, Section VIII, Division 1)}$$

$$P_i = 4590 \text{ kPa}$$

$$E = 0.85$$

$$D_i = 650 \text{ mm}$$

$$S = 184 \text{ MPa (ASME boiler and pressure vessel code, Section II, part D)}$$

So, substituting values,

$$t_n = 9.709 \text{ mm}$$

including 3 mm for corrosion allowance,

Adopted nozzle thickness = 15 mm

e. Product outlet diameter:

Since inlet and outlet flowrates are same, outlet nozzle dimensions are same as inlet nozzle.

Outlet nozzle diameter = 650 mm

Outlet nozzle thickness = 15 mm

f. Distributor:

From KAFCO Data,

Diameter of holes = 25 mm

Hole numbers/ row = 60

Number of rows = 18

Distributor radius = 1160 mm

Distributor height = 870 mm

g. Collector:

From KAFCO Data,

Mesh Type: Wire mesh AISI 304 (stainless steel) with 13 mm holes

Collector diameter = 1740 mm

Collector height = 700 mm

45 lots for drainage of 50 mm diameter and 600 mm height

h. The manhole diameter is 20 inches = 510 mm (KAFCO)

$$\text{thickness} = \frac{ID}{2} \times \left(\left(\frac{S \times E + P}{S \times E - P} \right)^{0.5} - 1 \right) + C \approx 70 \text{ mm}$$

- i. Handhole diameter = 16 inches = 405 mm (KAFCO)

Column Support Calculations:

Material of construction: ASTM A283 grade C carbon steel

Type of support: Skirt support

Inner diameter of vessel, $D_i = 4.2\text{m}$

Outer diameter of vessel, $D_o = 4.38\text{m}$

Height of catalyst bed, $Z = 5.4\text{m}$

Volume of catalyst bed, $V_b = 73.23 \text{ m}^3$

Bed void fraction, $\epsilon = 0.4$

Head thickness = 0.05m

Height of column for LTS, $h = \text{Bed height} + \text{hemispherical head height}$

$$= \text{bed height} + 2 \times \text{outer radius of hemispherical head}$$

$$= 5.4 + 2 \times \left(\frac{4.2}{2} + 0.05 \right)$$

$$= 9.7\text{m}$$

Density of vessel material (ASTM A516 grade 70 CS), $\rho_s = 7850 \text{ kg/m}^3$

Gas density at design condition, $\rho_f = 11.95 \text{ kg/m}^3$

Bulk density of catalyst, $\rho_b = 1360 \text{ kg/m}^3$

Now,

Weight of different components –

i. Shell Weight, $W_1 = \frac{\pi}{4} \times Z \times \rho_s \times (D_o^2 - D_i^2) \times g = 503638\text{N}$

ii. Hemispherical head weight (top + bottom),

$$W_2 = 2 \times \frac{2\pi}{3} \times (R_o^3 - R_i^3) \times \rho_s \times g$$

$$= 218169\text{N}$$

iii. Catalyst packing weight, $W_3 = V_b \rho_b g = 73.23 \times 1360 \times 9.8$

$$= 976009 \text{ N}$$

iv. Fluid Weight, $W_4 = V_f \rho_f g$

$$= (73.23 \times 0.4 + \frac{4}{3} \times 3.14 \times 2.1^3) \times 11.95 \times 9.8$$

$$= 7971 \text{ N}$$

From KAFCO Data,

v. Weight of nozzles, $W_5 = 24568 \text{ N}$

- vi. Weight of insulation, $W_6 = 28135 \text{ N}$
- vii. Weight of ladder, $W_7 = 1400H = 1400 \times 8.48 \text{ m} = 11872 \text{ N}$
- viii. Weight of manhole, $W_8 = 48902 \text{ N}$
- ix. Others, $W_9 = 26313 \text{ N}$

$$\text{Total } W = \sum W_i = 1845577 \text{ N}$$

Involved stresses –

Let, t_{sk} = skirt thickness

a. Stress due to dead load, $f_d = \frac{\text{Total Weight}}{D_i \times t_{sk}} = \frac{1845577}{4.2 t_{sk}} = \frac{439423}{t_{sk}} \text{ (Joshpdf.Pdf, n.d.)}$

b. Stress due to wind load:

Let,

Height of skirt is 4.42m (KAFCO Data)

$$\begin{aligned} H &= \text{height of vessel} + \text{height of skirt} - \text{height of vessel within skirt length} \\ &= 9.7\text{m} + 4.42 \text{ m} - 1.71\text{m} \text{ (KAFCO Data)} \\ &= 12.41\text{m} \end{aligned}$$

Now,

$$\text{Force due to wind load, } F_w = k p_1 h_1 D_0$$

k = coefficient depending on the shape factor = 0.7 (Cylindrical vessel)

$$p_1 = 100 \text{ kg/m}^2 = 980.66 \text{ Pa} \text{ (Assuming maximum pressure)}$$

$$h_1 = H = 12.41\text{m}$$

$$D_0 = 4.38\text{m}$$

Substituting values,

$$F_w = 37313 \text{ N}$$

Bending moment due to wind at the base of vessel,

$$M_w = F_w \times \frac{H}{2} = 231527 \text{ N.m}$$

$$\text{Stress due to wind, } f_w = \frac{4 \times M_w}{\pi D_0^2 t_{sk}} = \frac{15374}{t_{sk}} \text{ (Joshpdf.Pdf, n.d.)}$$

c. Stress due to seismic load,

Seismic load, $F = CW$

$$W = \text{weight of vessel} = 1845577 \text{ N}$$

C = seismic coefficient = 0.28 (BNBC 2020)

$$\text{So, } F = 516761\text{N}$$

$$\text{Stress due to seismic load, } f_s = \frac{2 \times F \times H}{3 \times \pi \times R_{sk}^2 \times t_{sk}} \text{ (Joshpdf.Pdf, n.d.)}$$

$$\begin{aligned}
&= \frac{2 \times 516761 \times 12.41}{3 \times 3.14 \times 2.19^2 \times t_{sk}} \\
&= \frac{283892}{t_{sk}}
\end{aligned}$$

Now, Maximum compressive stress,

$$\begin{aligned}
f_{\max} &= f_d + f_w + f_s = \frac{1}{3} \times \text{Yield point of carbon steel} \\
\Rightarrow \frac{738689}{t_{sk}} &= \frac{1}{3} \times 265 \times 10^6 \text{ N/m}^2 \\
\Rightarrow t_{sk} &= 8.36 \text{ mm}
\end{aligned}$$

Considering corrosion allowance, adopted skirt thickness, **$t_{sk} = 15 \text{ mm}$**

11.2.2 Mechanical Drawings of LTS

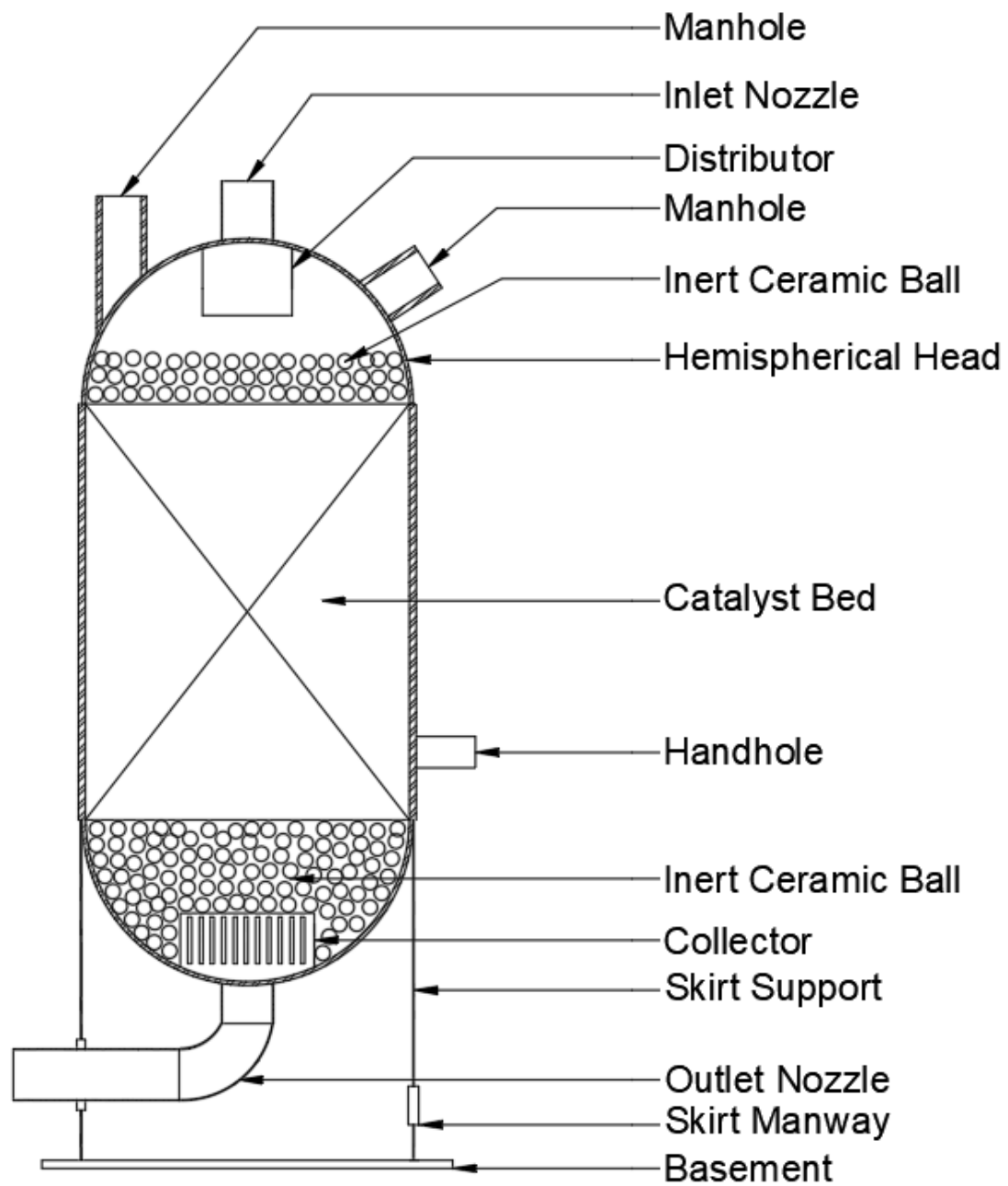


Figure 11.2.2: Different parts of low temperature shift reactor (LTS)



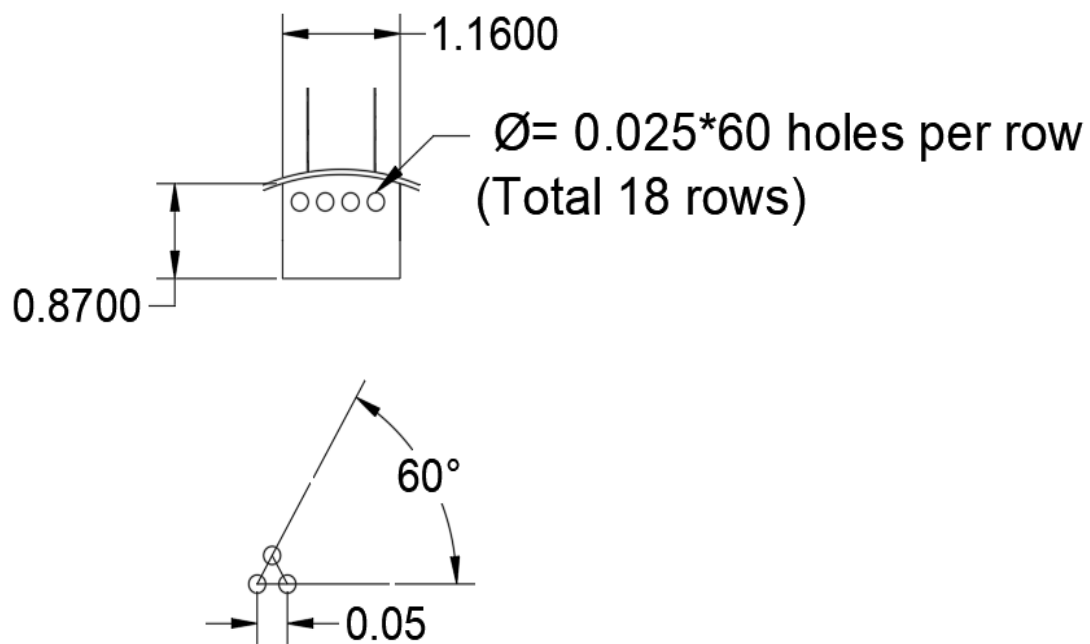


Figure 11.2.4: Detailed mechanical drawing of distributor in LTS

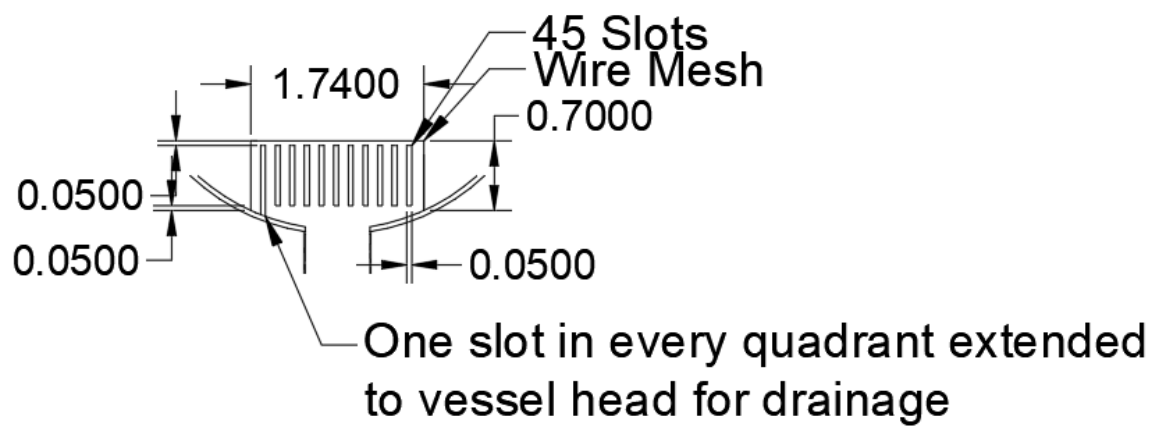


Figure 11.2.5: Detailed mechanical drawing of collector in LTS (All dimensions are in m)

11.2.3 Piping & Instrumentation Diagram of LTS

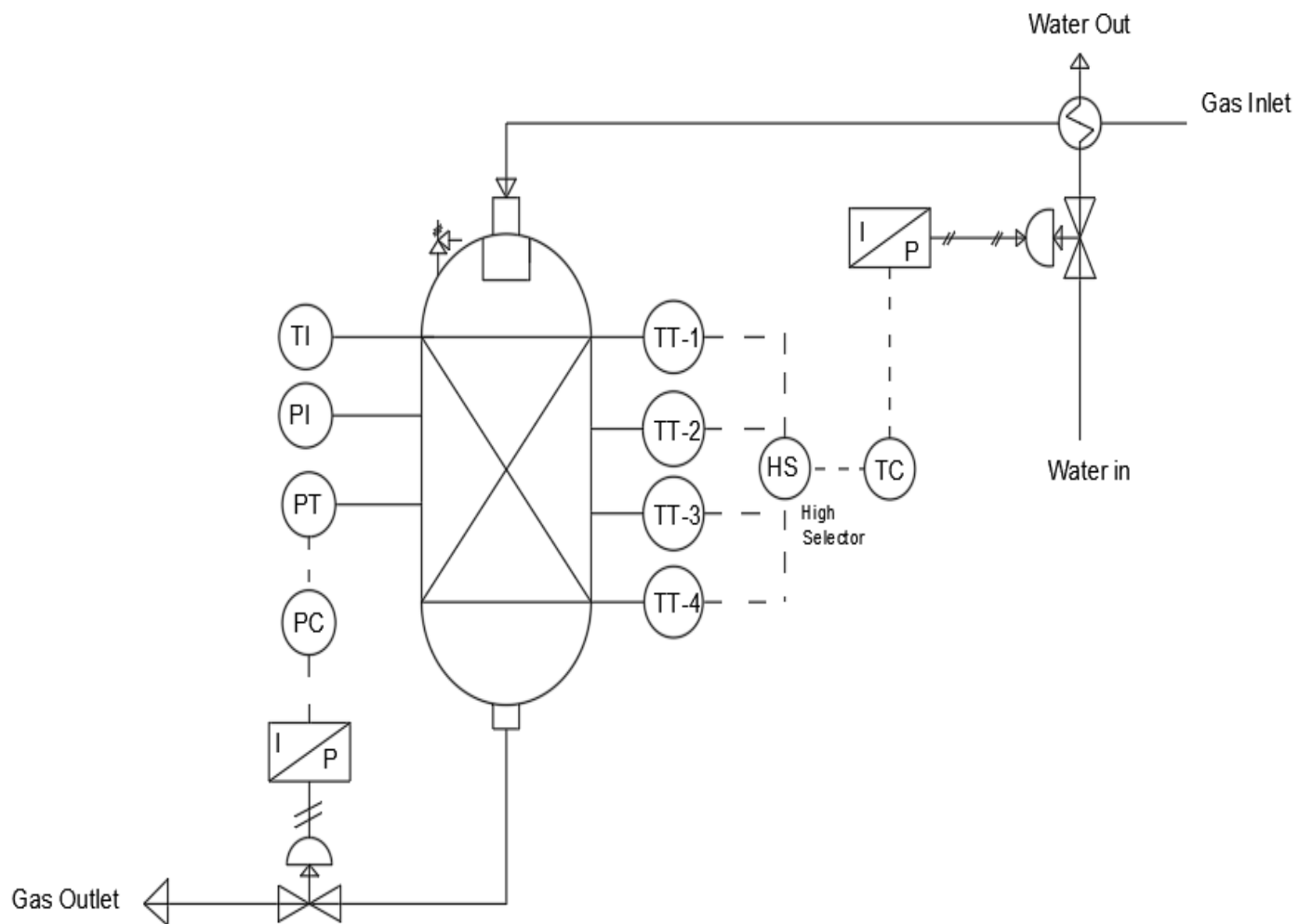


Figure 11.2.6: P&ID diagram of low temperature shift reactor

11.2.4 Equipment Specification Sheet

Table 11.2.1: LTS reactor Specification Sheet

Customer: Document no: Sheet no:		Date:19/05/2023 Item: R-104 Quantity:1
Function: Converting CO into CO ₂ and H ₂ using water gas shift reaction		
Operation type: Continuous		
Design data		
Catalyst bulk density:	Katalco 83-3	1360 kg/m ³
ΔP of catalyst bed:		0.0107 kg/cm ²
Shell length		9.7 m
Shell diameter		4.38 m
Catalyst bed length		5.4 m
Support		Skirt
Operation Condition		
Fluid		Process gas
Pressure		31.2 kg/cm ² G
Temperature		In 220 °C/ Out 260 °C
Design condition		
Code: ASME Boiler and Pressure Vessel Code: Section VIII, Div. 1		
Design Pressure and temperature: 46.8 kg/cm ² at 285 °C		
Corrosion allowance: 3 mm		
Material		
Shell		A 516 Gr. 70
Head		A 516 Gr. 70
Tray/ internal		A 283 Gr. C
Support		A 283 Gr. C
Nozzle flange		A 105
Bolt and nut: Int.		304 SS
Ext.		A 193 Gr. B7

11.3 CO₂ Absorber

11.3.1 Calculation of Major Parameters

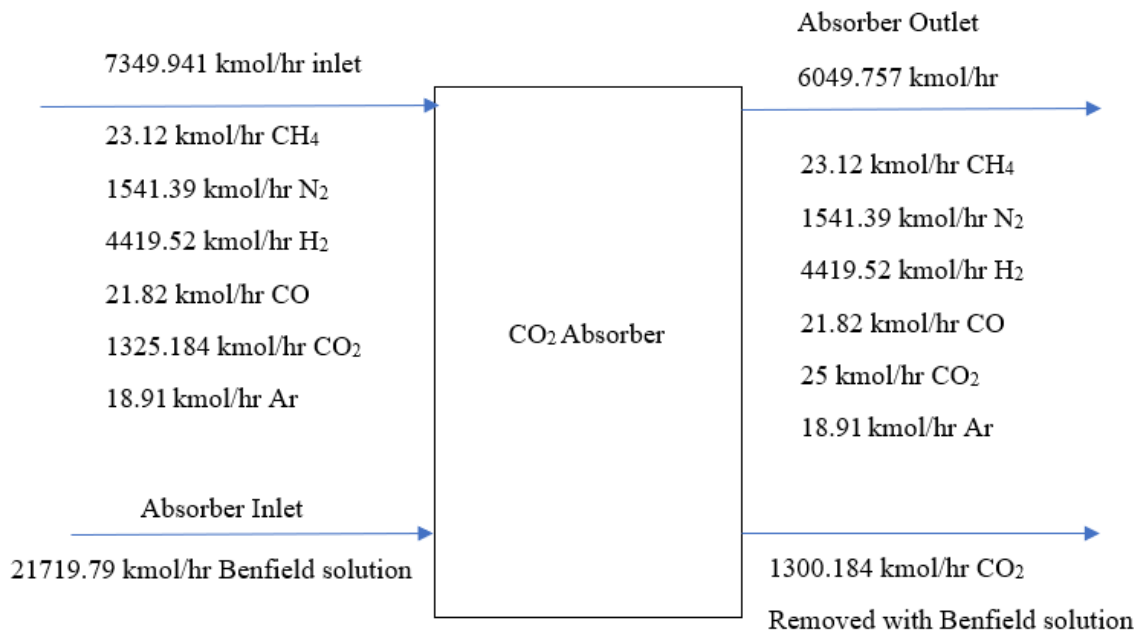


Figure 11.3.1: Inlet and outlet stream composition of absorber

Benfield solution, 30 wt% K₂CO₃, 2.9% DEA, 0.9% V₂O₅ and remaining H₂O.

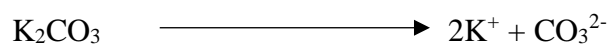
Molecular weight of Benfield solution = 27.61

Density of Benfield solution = 1297.9 kg/m³

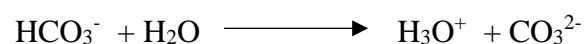
So, the molar flow rate of Benfield solution = 21719.79 kmol/hr

Chemical reactions of Absorber

Potassium carbonate dissociates first as a result of the following reaction:



The carbonate system in the solution in the absorber is created by this process. The three carbonic acid equilibrium processes make up the carbonate system:



The final reaction explains in detail how the feed stream's carbon dioxide is absorbed into the solution. This carbon dioxide reacts to mostly bicarbonate ions, which coexist with potassium bicarbonate in an equilibrium as in the following reaction:



Potassium bicarbonate, bicarbonate ion, and carbonic acid are all present in the bottom streams that exit the absorber, along with a few trace contaminants. This allows for the complete absorption of the feed stream's carbon dioxide into the bottom solution, which is subsequently sent on to the stripper.

Diameter Calculation

Taking packing size = 25 mm

Random packing system has been selected.

Using chart from (Peters et al., 2003a)(Page 784)

Packing factor = $F_p = 50$

Assuming that the Benfield (wt%) all through the tower is 30% and it is uniform all through the tower. Also, it is assumed that the temperature is fixed at 110 °C.

Liquid flow rate, $L = 21719.79$ kmol/hr

Gas flow rate, $V = 7349.94$ kmol/hr

From table 3.5 from '*Technology and manufacture of ammonia*' by Samuel Strelzeff

Molecular weight of Benfield solution, $MW_L = 27.61$

Density of Benfield solution, $\rho_L = 1297.9$ kg/m³

Viscosity of Benfield solution, $\mu_L = 11.94$ cp = 0.01193 Pa. s

From Hysys simulation,

Molecular weight of inlet gas mixture, $MW_v = 15.26$

Density of gas, $\rho_v = 10.74$ kg/m³

Viscosity of gas, $\mu_v = 1.519 \times 10^{-5}$ Pa.s

From Sherwood chart,

$$\text{Abcissa} = \frac{LMW(L)}{VMW(V)} \sqrt{\frac{\rho_v}{\rho_L - \rho_v}} = \frac{21719.79 \times 27.61}{7349.94 \times 15.26} \sqrt{\frac{10.74}{1297.9 - 10.74}} = 0.51$$

Pressure drop of the packed tower is calculated using the following co relation from '*Unit Operations of Chemical Engineers*' by McCabe, page 552

$$\Delta P_{\text{flood}} = 0.115 F_p^{0.7} = 0.115 \times 50^{0.7} = 1.79 \text{ in H}_2\text{O/ft} = 1461.36 \text{ Pa/m}$$

Actual pressure drop is assumed 80% of the flooding value.

$$\text{So, } \Delta P_{\text{actual}} = 0.8 \times 1461.36 = 1169.09 \text{ Pa/m}$$

From this pressure drop and abscissa, using Sherwood chart, the ordinate is found 0.11.

Now,

$$\frac{G^2 F \mu l^{0.1}}{\rho V (\rho_L - \rho_V)} = 0.11$$

So, $G = 6.7 \text{ Kg/m}^2.\text{s}$

$$\text{Again, } G = \frac{(Mv)V}{A}$$

Now, $A = (\pi/4) \times D^2$

So, $D = 2.48 \text{ m} = 8.2 \text{ ft} = 8.5 \text{ ft} = 102 \text{ inch}$

The diameter has been taken as 102 inch

Height Calculation

As the diameter is greater than 0.55m so, $\text{HETP} = D^{0.3} = 2.48^{0.3} = 1.31 \text{ m} = 4.3\text{ft}$

So, the HETP is taken 4.5 ft.

At the inlet and outlet of the gas stream, the composition of CO_2 is 18.03% and 0.41% respectively. The equation of the operating line is $Y_N = \frac{L}{G} X_N + (Y_1 - \frac{L}{G} X_0)$

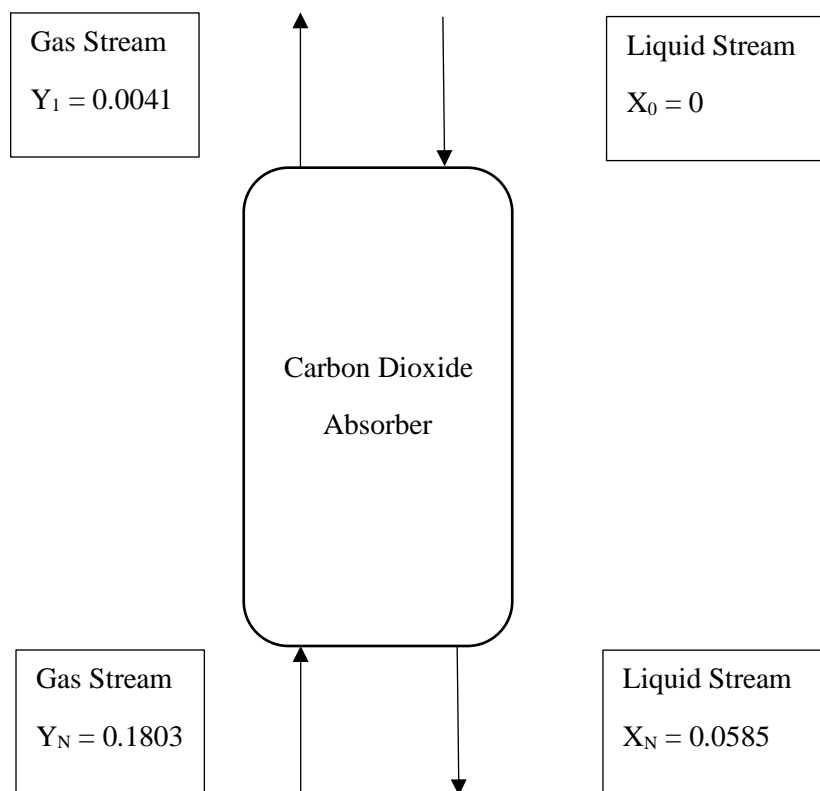


Figure 11.3.2: CO_2 Absorber tower with gas and solvent molar ratio operating at 110°C

Solving the operating line equation, $X_N = 0.0585$

The equilibrium data for Benfield solution at 110°C is given below

Table 11.3.1: Equilibrium data for Benfield solution

Equilibrium Data Table			
x	X	Y	y
0	0	0.004	0.00398
0.001	0.001	0.0057	0.00567
0.0015	0.0015	0.00655	0.00651
0.002	0.002	0.00741	0.00735
0.0025	0.00251	0.00826	0.00819
0.003	0.00301	0.00912	0.00903
0.0035	0.00351	0.00997	0.00987
0.004	0.00402	0.01083	0.01071
0.0045	0.00452	0.01168	0.01155
0.005	0.00503	0.01254	0.01239
0.0055	0.00553	0.0134	0.01322
0.006	0.00604	0.01426	0.01406
0.0065	0.00654	0.01512	0.0149
0.007	0.00705	0.01598	0.01573
0.0075	0.00756	0.01685	0.01657
0.008	0.00806	0.01771	0.0174
0.0085	0.00857	0.01857	0.01824
0.009	0.00908	0.01944	0.01907
0.0095	0.00959	0.0203	0.0199
0.0098	0.0099	0.02082	0.0204

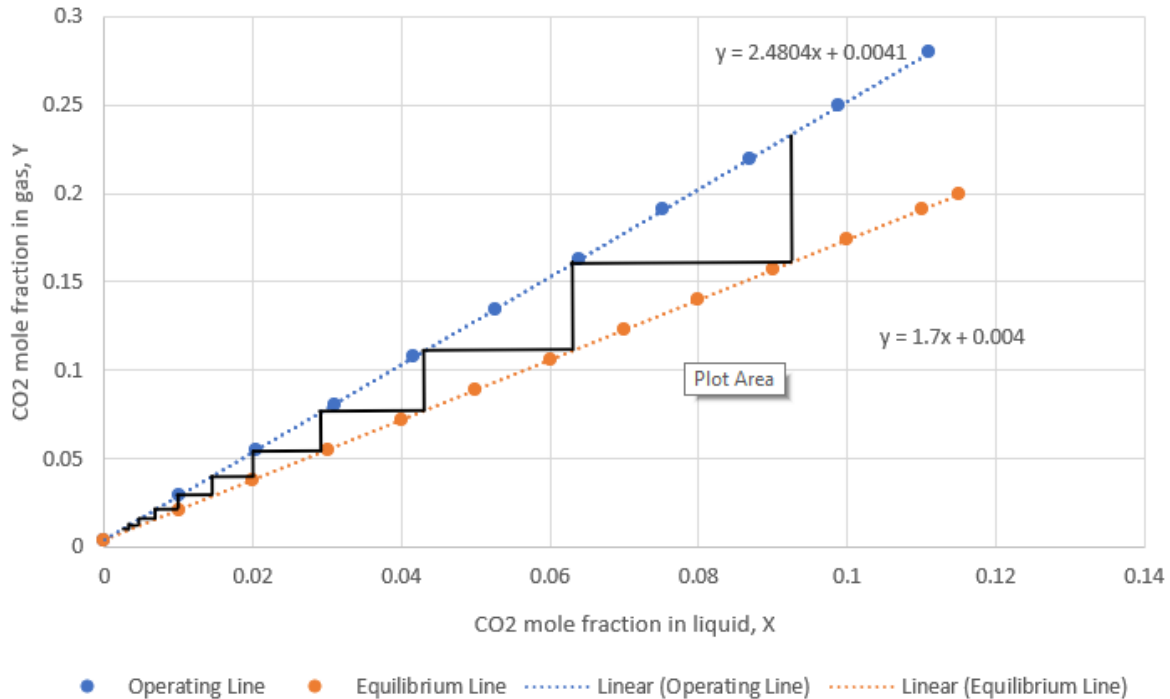


Figure 11.3.3: McCabe Thiele Diagram

From McCabe Thiele diagram, number of stages, $N = 10$

So, height of the packing = $HETP \times N = 4.5 \times 10 = 45$ ft

For random packing, over design requirement is 10% (Distillation and Other Industrial Separation by Gailf Nalven 26 pp.)

So, Actual Height will be, $H_{\text{actual}} = 1.1 \times 45 = 49.5$ ft. = 15.09 m

The Height of the packing thus taken as 49.5 ft.

15% extra space is provided above the packing for necessary attachments and 10% extra space required at the beneath of packing for liquid holdup before drainage.

So, extra height at the top + extra height at the bottom = $49.5 \times 0.15 + 49.5 \times 0.1$ ft
= 12.38 ft

So, total tower height = $49.5 + 12.38 = 61.88$ ft = **18.86 m**

Thus, Height/Diameter = $18.86 \text{ m} / 2.48 \text{ m} = 7.6$ (approximately), which is quite satisfactory.

Selection of Vessel Material and Packing Material

Carbon Steel is suitable for Benfield solution, it's a good choice for both vessel and packaging. Carbon steel can be chosen for handling Benfield solution for the following reasons:

1. **Cost-Effective:** Carbon steel is a cost-effective material compared to other corrosion-resistant alloys such as stainless steel or nickel alloys. It is a widely available and commonly used material in the industry, which makes it an economical choice.
2. **Compatibility:** Carbon steel is compatible with Benfield solution, which is used for gas treatment in the oil and gas industry. Carbon steel can withstand various concentrations of the solution and exposure to different temperatures and pressures.
3. **Mechanical Properties:** Carbon steel has excellent mechanical properties, such as high strength and toughness, which makes it a suitable material for handling the equipment and components involved in the Benfield process.
4. **Weldability:** Carbon steel is easily weldable, which makes it a popular choice for fabricating and repairing equipment and components used in the Benfield process.
5. **Durability:** Carbon steel has good durability and can withstand high temperatures, pressures, and corrosive environments. Proper maintenance and protection can extend the service life of carbon steel components and equipment.

Vessel Material specification number is **SA 1008** and this is **CS-A** type steel.

Raschig Rings of carbon steel are the packing material. (Applied Process Design for Chemical and Petrochemical Plants, Volume 2, 254 pp.).

Assuming, Outer Diameter of Raschig Ring = 1 in

Raschig Ring Wall Thickness = 1/16 in

((Applied Process Design for Chemical and Petrochemical Plants, Vol. 2 (Ernest E. Ludwig (Ed.)).Pdf, n.d.) page 249)

Liquid Distributor

Full cone Spray Nozzle has been selected as liquid distributor (Applied Process Design for Chemical and Petrochemical Plants, Vol. 2 (Ernest E. Ludwig (Ed.)).Pdf, n.d.) (page 254)

A Perforated Flash Plate is required for uniform distribution of liquid on which the sprayed liquid will be collided.

Flash Plate Diameter = Column Diameter = 102 inch

Flash Plate holes/drip points = 20 holes/ft² surface area

Total Surface area of the plate = $\pi \times D^2 / 4 = \pi \times 8.5^2 / 4 = 56.74 \text{ ft}^2$

So, holes required, $n = 20 \times 56.74 \approx 1130$ holes

Hole diameter = 3/8 in.

Liquid Redistributor

Metal Rosette Redistributor has been selected as liquid redistributor.

Using the reference(*Process Equipment Design (Joshi).Pdf*, n.d.) (page 328), for Raschig rings liquid redistributor is required at $3D$ distance interval where D is column Diameter.

Total Packing Height is 45 ft = 13.72 m.

$3D = 3 \times 8.5 = 25.5$ ft = 7.77 m

So, total liquid redistributor required = $13.72/7.77 = 1.77 \approx 2$

Using Table from(*Applied Process Design for Chemical and Petrochemical Plants, Vol. 2 (Ernest E. Ludwig (Ed.)).Pdf*, n.d.) (page 270)

Outside Diameter of the Rosette Strip = $102 + 1/16 = 102.06$ in

Where 102 inch is the internal Diameter and 1/16 in is the Rosette Thickness.

Rosette Petal Length = 3.5 in

Excluding rosette petal, the diameter of the redistributor = $102 - 2 \times 3.5 = 95$ inch

Height of metal plate = 3 in

Rosette Plate will be installed between tower flanges.

Support Plate

Support Plate is required to hold up the packings providing necessary empty spaces at the bottom.

Using Chart from(*Applied Process Design for Chemical and Petrochemical Plants, Vol. 2 (Ernest E. Ludwig (Ed.)).Pdf*, n.d.) (page 256.) for 102 in. tower diameter, (extrapolation data)

Diameter of Plate = 99.5 in

Height of Plate = 7.5 in

Diameter of holes = 5.2 in

Number of holes = 241

Metal Hold Down Plate

It rests directly top of the packing. Using Chart from(*Applied Process Design for Chemical and Petrochemical Plants, Vol. 2 (Ernest E. Ludwig (Ed.)).Pdf*, n.d.) (page 271) specification of metal hold down plate can be calculated by extrapolation.

Height of the hold down plate = 4 in

Plate Diameter = 101.75 inch

Thickness Calculation

Shell

From reference,

Tensile Strength of SA-1008 Steel = 340 MPa

Density, $\rho_s = 7872 \text{ kg/m}^3$

Maximum Allowable Stress, $S = 78600 \text{ KPa}$

Using ASME BPV Code (Sec. VIII D.1 Part UG-27) thickness of the shell can be calculated using the following equation,

$$t_s = \frac{P \times ID}{2SE - 0.6P} + C$$

Operating Pressure of the column = 2500 KPa

Using the chart from Applied Process Design for Chemical and Petrochemical Plants, Volume 1, 409 pp. design pressure is taken 121% of the operating pressure.

Thus, $P = 1.21 \times 2500 = 3025 \text{ KPa}$

ASME Code Standard Joint Efficiency, $E = 0.85$

Corrosion Allowance required for handling benfield Solution, $c = 3.8 \text{ mm} = 0.15 \text{ in}$

Thus,

$$t_s = \frac{P \times ID}{2SE - 0.6P} + c = \frac{3025 \text{ KPa} \times 102 \text{ in}}{2 \times 78600 \text{ KPa} \times 0.85 - 0.6 \times 3025 \text{ KPa}} + 0.15 = 2.49 \text{ in} = 63.27 \text{ mm}$$

So, Outer Diameter of the shell, $OD = 102 + 2 \times 2.49 = 106.98 \text{ in.} = 2.72 \text{ m}$

Absorber Head

It's been assumed that tower head is hemispherical.

Now, using UG-32(d) of ASME Section VIII Division I, 2011 edition, the thickness of the hemispherical head can be calculated using the following equation:

$$t_h = \frac{P \times ID}{2SE - 0.2P} + c$$

Putting the previously stated value, the hemispherical head thickness can be obtained.

$$t_h = \frac{3025 \text{ KPa} \times 102 \text{ in}}{2 \times 78600 \text{ KPa} \times 0.85 - 0.2 \times 3025 \text{ KPa}} + 0.15 = 2.47 \text{ in} = 62.73 \text{ mm.}$$

Manhole design: As number of liquid distributors is 2 then 3 manholes have to be used.

((*Process Equipment Design (Joshi).Pdf*, n.d.),371 page.)

$$\text{Thickness of manhole} = t_{mh} = \frac{ID}{2} \times \left(\left(\frac{S \times E + P}{S \times E - P} \right)^{0.5} - 1 \right) + C$$

$$= \frac{2.59}{2} \times \left(\left(\frac{78600 \times 0.85 + 3025}{78600 \times 0.85 - 3025} \right)^{0.5} - 1 \right) + 3.8 \times 10^{-3}$$

$$= 63.84 \text{ mm} = 0.063 \text{ m}$$

Taking manhole 42 inch in diameter and 12 inch in length.

Nozzle Design

Material selected is carbon steel.

Considering the diameter of nozzle $D_n = 0.3 \text{ m}$

Thickness of nozzle is t_n .

$$t_n = \frac{p \times D_n}{2 \times S \times E - P} + C = \frac{3025 \times 0.3 \times 1000}{2 \times 78600 \times 0.85 - 3025} + 3.8 \times 10^{-3} = 6.95 \text{ mm}$$

Insulation

Using the equation of critical insulation thickness,

$$r_o = k/h$$

Asbestos Cement Sheet is taken as Insulating Material.

Here, k = Thermal Conductivity of carbon steel at $41^\circ\text{C} = 0.181 \text{ W/m.k}$

And h = Convection Co-efficient of Air at design temperature 109°C and 20 m/s wind velocity

$$= 1.16(10.45 - v + 10 \times v^{0.5}) = 1.16(10.45 - 20 + 10 \times 20^{0.5}) = 40.79 \text{ W/m}^2 \cdot \text{k}$$

Putting these values in the above equation,

Thickness of insulation is found as, $r_o = 0.00444 \text{ m} = 4.4 \text{ mm} = 0.17 \text{ in}$ (Approximately)

Thus, The outer diameter of the column with insulator = $106.98 + 2 \times 0.17 = 107.32 \text{ in.} = 2.73 \text{ m}$

Maximum Weight of the Tower Calculation

$$W_{\max} = W_s + W_h + W_i + W_T + W_b + W_l + W_m + W_a$$

Here, W_s = Shell Weight.

$$\text{Volume of the Shell} = \frac{\pi(2.73^2 - 2.59^2)18.86}{4} = 11.03 \text{ m}^3$$

$$\text{So, } W_s = 11.03 \times 7872 = \mathbf{86828.16 \text{ kg.}}$$

W_h = Weight of Head

$$\text{Volume of Head } V_h = \left(\frac{4}{3} \times \pi \times r^3 \right) \times 0.5 \text{ m}^3 = 0.7782 \text{ m}^3$$

$$W_h = V_h \times 7872 \text{ kg} = 6126 \text{ kg}$$

Again, W_i = Weight of Insulation.

$$\text{Volume of the Insulation} = \frac{\pi \times 18.86 \times (2.7259^2 - 2.7173^2)}{4} = 0.69 \text{ m}^3$$

Density of Asbestos Cement Sheet = 270 kg/m^3

$$\text{So, } W_i = 0.69 \times 270 = \mathbf{186.3 \text{ kg}}$$

Again, W_T = Weight of Shell during Water Test.

$$\text{Volume of empty space inside the shell} = \frac{\pi \times 2.59^2 \times 18.86}{4} = 99.36 \text{ m}^3$$

Density of water is taken 1000 kg/m³.

So, the weight of water if the shell is poured with water, $W_T = 99.36 \times 1000 = \mathbf{99360 \text{ kg}}$

$$W_b = \text{Weight of Base} = 7872 \times \pi \times \left(\frac{OD}{2}\right)^2 \times t_s = 7872 \times \pi \times (2.73/2)^2 \times 0.06274 = 2915.4 \text{ kg.}$$

$W_l = \text{Weight of ladder}$

It is assumed that two 18.86 m length with 0.0618 m diameter rod and 0.65 m length with 0.0618 m diameter 10 rods are used

$$\begin{aligned} W_l = \text{Weight of ladder} &= (\pi \times (0.0618)^2 \times 18.86 + \pi \times (0.0618)^2 \times 0.65 \times 10) \times 7872 \text{ kg} \\ &= 2395.31 \text{ kg} \end{aligned}$$

$W_m = \text{Weight of manhole}$

$$\text{Volume of 1 manhole } V_m = \pi \times \frac{1.2192^2}{4} \times 0.0633 + \pi \times 1.2192 \times \left(\frac{1.28372^2 - 1.2192^2}{2}\right) = 0.3832 \text{ m}^3$$

$$W_m = 3 V_m \times 7872 = 9049.7 \text{ kg.}$$

Again, $W_a = \text{Weight of Attachments.}$

Surface Area of Flash Plate = Total Area – Hole Area

$$\begin{aligned} \text{Hole area} &= \text{hole number} \times \text{each hole area} = 1130 \times \pi/4 \times (3/8 \text{ in})^2 = 1130 \times \pi/4 \times (0.0312 \text{ ft})^2 \\ &= 0.8639 \text{ ft}^2 \end{aligned}$$

$$\text{So, surface area of flash plate} = 56.74 \text{ ft}^2 - 0.8639 \text{ ft}^2 = 55.88 \text{ ft}^2$$

$$\text{Volume of Flash Plate} = \text{Surface Area} \times \text{Height} = 55.88 \text{ ft}^2 \times 0.25 \text{ ft} = 13.97 \text{ ft}^3 = 0.3956 \text{ m}^3$$

$$\text{So, Weight} = 7872 \times 0.3956 = \mathbf{3114.16 \text{ kg}}$$

$$\begin{aligned} \text{Surface Area of Redistributor} &= \text{Redistributor Number} \times (\text{Surface Area} - \text{Hole Area}) = 2 \times \\ & (56.74 \text{ ft}^2 - 0.8639 \text{ ft}^2) = 111.75 \text{ ft}^2 \end{aligned}$$

$$\begin{aligned} \text{Volume of Redistributor} &= \text{total surface area} \times \text{Height} = 111.75 \text{ ft}^2 \times 0.25 \text{ ft} = 27.94 \text{ ft}^3 = 0.79 \\ & \text{m}^3 \end{aligned}$$

$$\text{So, Weight} = 7872 \times 0.79 = \mathbf{6218.88 \text{ kg}}$$

Support plate

$$\text{Diameter of Plate} = 99.5 \text{ in} = 8.29 \text{ ft.}$$

$$\text{Total area of support plate} = \pi/4 \times 8.29^2 = \mathbf{53.98 \text{ ft}^2}$$

$$\text{Height of Plate} = 7.5 \text{ in} = 0.625 \text{ ft}$$

$$\text{Diameter of holes} = 5.2 \text{ in} = 0.43 \text{ ft}$$

$$\text{Number of holes} = 241$$

$$\text{Hole area} = 241 \times \pi/4 \times 0.43^2 = \mathbf{35 \text{ ft}^2}$$

$$\text{Surface Area of Support Plate} = \text{Total Area} - \text{Hole Area} = 53.98 \text{ ft}^2 - 35 \text{ ft}^2 = 18.98 \text{ ft}^2$$

Volume of Support Plate = Surface Area \times Height = $18.98 \text{ ft}^2 \times 0.625 \text{ ft} = 11.86 \text{ ft}^3 = 0.336 \text{ m}^3$

. So, weight = $0.336 \times 7872 = \mathbf{2645 \text{ kg}}$

Metal Hold Down Plate approx. net weight = 498.41 lbs = **226.07 kg** (Chart from(*Applied Process Design for Chemical and Petrochemical Plants, Vol. 2 (Ernest E. Ludwig (Ed.)).Pdf*, n.d.))

Volume of Packing Space = $\pi/4 \times 2.59^2 \times 18.86 = 99.36 \text{ m}^3 = 3508.86 \text{ ft}^3$

Approximate Average Weight/ft³ = 99.68 lbm (Ludwig, Volume 2, 249 pp.)

So, Packing Weight = $99.68 \times 3508.86 = 349763.16 \text{ lbm} = \mathbf{158650 \text{ kg}}$

So, $W_{\max} = 377488.9 \text{ kg} = 377.49 \text{ Ton}$

Support Design

As the column is tall so skirt support is chosen for supporting, and the column is subjected to wind, seismic and other loads that cause bending moment to the shell.

The cylindrical shell of the skirt is designed for the combination of the stress due to dead load, wind load, seismic load. The skirt thickness is uniform and is designed for the maximum tensile or compressive stress. These stresses are:

1. Due to dead weight:

$$f_d = \frac{\text{Total weight}}{\pi \times OD \times t_{sk}} = \frac{377488.9 \times 9.8}{\pi \times 2.73 \times t_{sk}} = 431338/t_{sk}$$

2. Stress due to wind load:

$$P_{fw} = k \times P_1 \times Z \times D_0 \text{ (for height less than 20m)}$$

Where, k is the co-efficient of shape factor. = 0.7(for cylindrical shape).

Wind pressure of the vessel in lower part $P_1 = 686.465 \text{ N/m}^2$ ((*Process Equipment Design (Joshi).Pdf*, n.d.) (page 391)

Z= height of the tower = 18.86m

OD = outer diameter of the tower with insulation= 2.73 m

$$P_{fw} = 0.7 \times 686.465 \times 18.86 \times 2.73 = 24741.2 \text{ N/m}^2$$

Bending moment due to wind at the vessel base $M_w = P_{fw} \times Z \times 0.5 = 233309.5 \text{ Nm}$

$$f_{wb} = \frac{4}{\pi \times OD \times t_{sk}} M_w = 108812.8/t_{sk}$$

3. Stress due to seismic load

Load F= WC

W = total weight of the vessel

C = 0.08

R_{ok} = Radius of the skirt = half of vessel diameter = 1.365 m

Height of skirt = 2 m

$$f_{sb} = \frac{2CWZ}{3 \times \pi \times R o k^2 \times t_{sk}} = \frac{2 \times 0.08 \times 377488.9 \times 9.8 \times 18.86}{3 \times \pi \times 1.365^2 \times t_{sk}} = 635704.82 / t_{sk}$$

Maximum compressive stress

$$f_{cmax} = f_{sb} + f_d + f_{wb} = 1175855.62 / t_{sk}$$

SA-516 (Gr.70) carbon steel is chosen for skirt.

Yield strength of SA-516 = 260 MPa = 260×10^6 Pa

$f_c \leq (1/3)$ yield strength

$$f_c = 86.67 \times 10^6 \text{ Pa}$$

$$86.67 \times 10^6 = 1175855.62 / t_{sk}$$

$$t_{sk} = 73.71 \text{ mm.}$$

So, the support thickness should be 80 mm or 3.15 inch.

Absorber Design Summary

Table 11.3.2: Absorber design parameters

Parameter	Value
Column diameter	2.59 m
Packing height	13.72 m
Packing type	Raschig ring
Packing material	Carbon steel
Column height	18.86 m
Shell material	SA-1008 carbon steel
Shell thickness	63.27 mm
Head type	Hemispherical
Head thickness	62.73 mm
No. of manhole	3
Manhole diameter	1.07 m
Manhole thickness	63.84 mm
Nozzle thickness	6.95 mm
Skirt height	2 m
Skirt thickness	73.71 mm
Standards followed	ASME

11.3.2 Mechanical Drawing of CO₂ Absorber

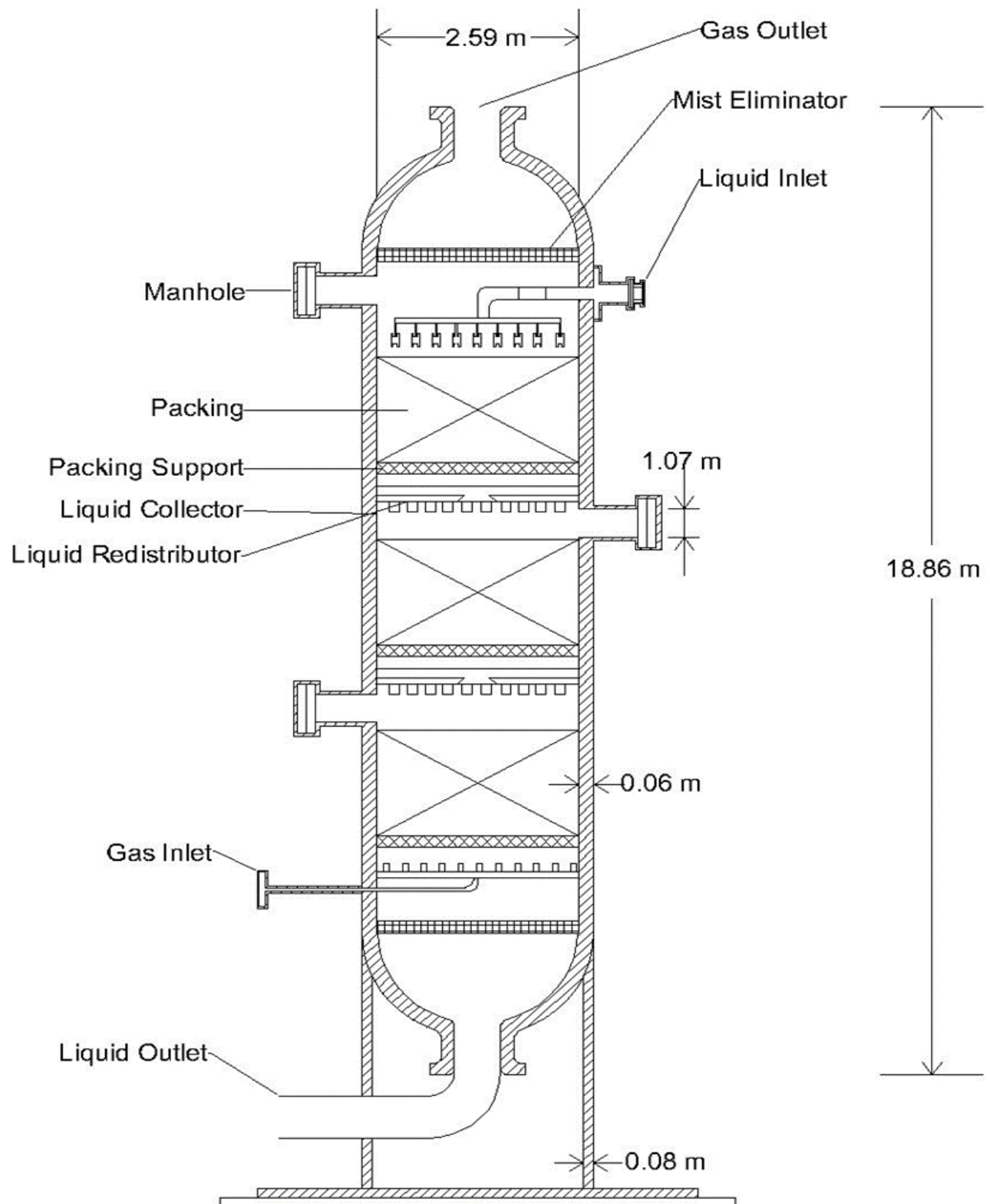


Figure 11.3.4: Mechanical Design of the absorber tower

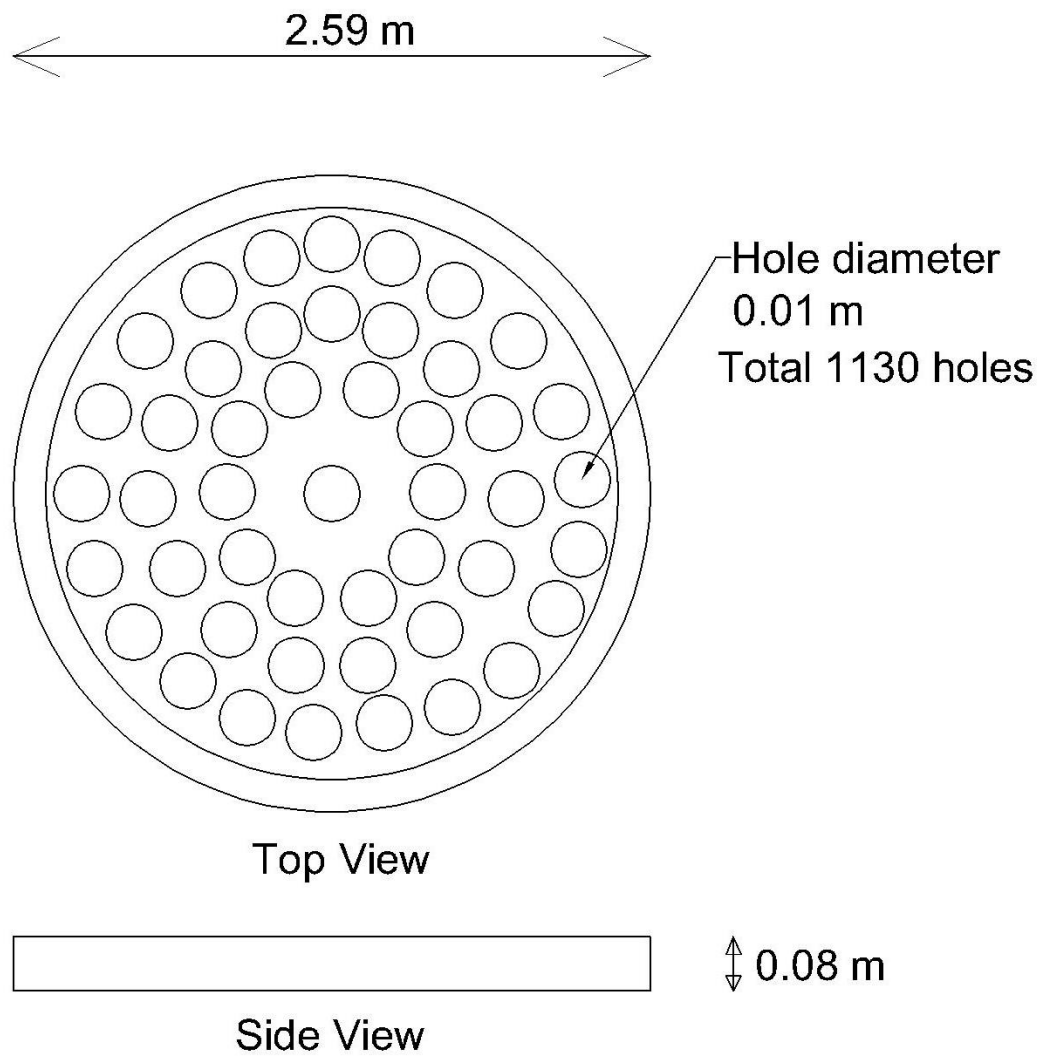


Figure 11.3.5: Flash Plate

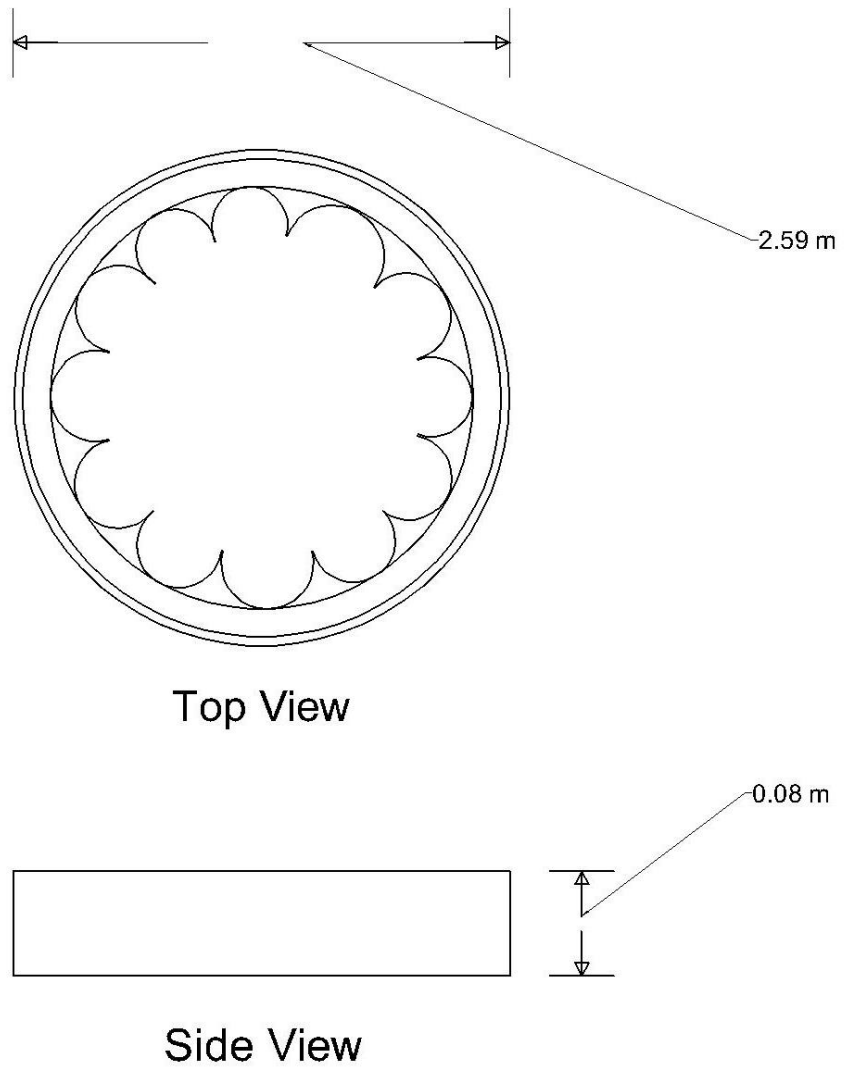


Figure 11.3.6: Liquid Distributor

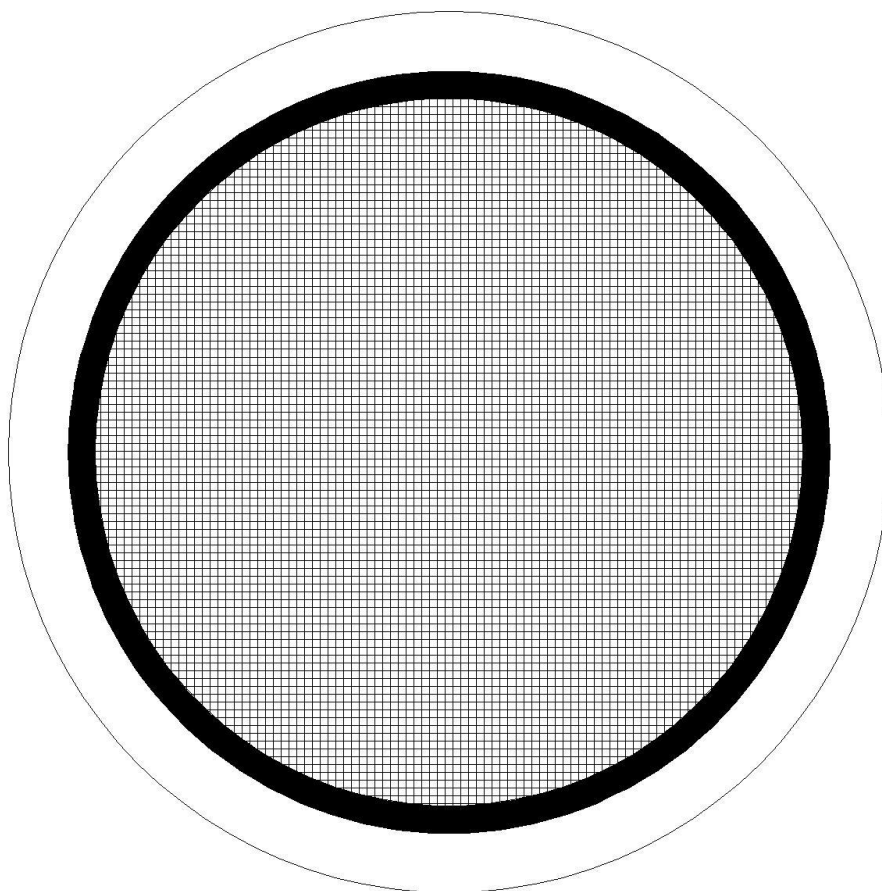


Figure 11.3.7: Mist Eliminator

11.3.3 Piping and Instrumentation Diagram of CO₂ Absorber

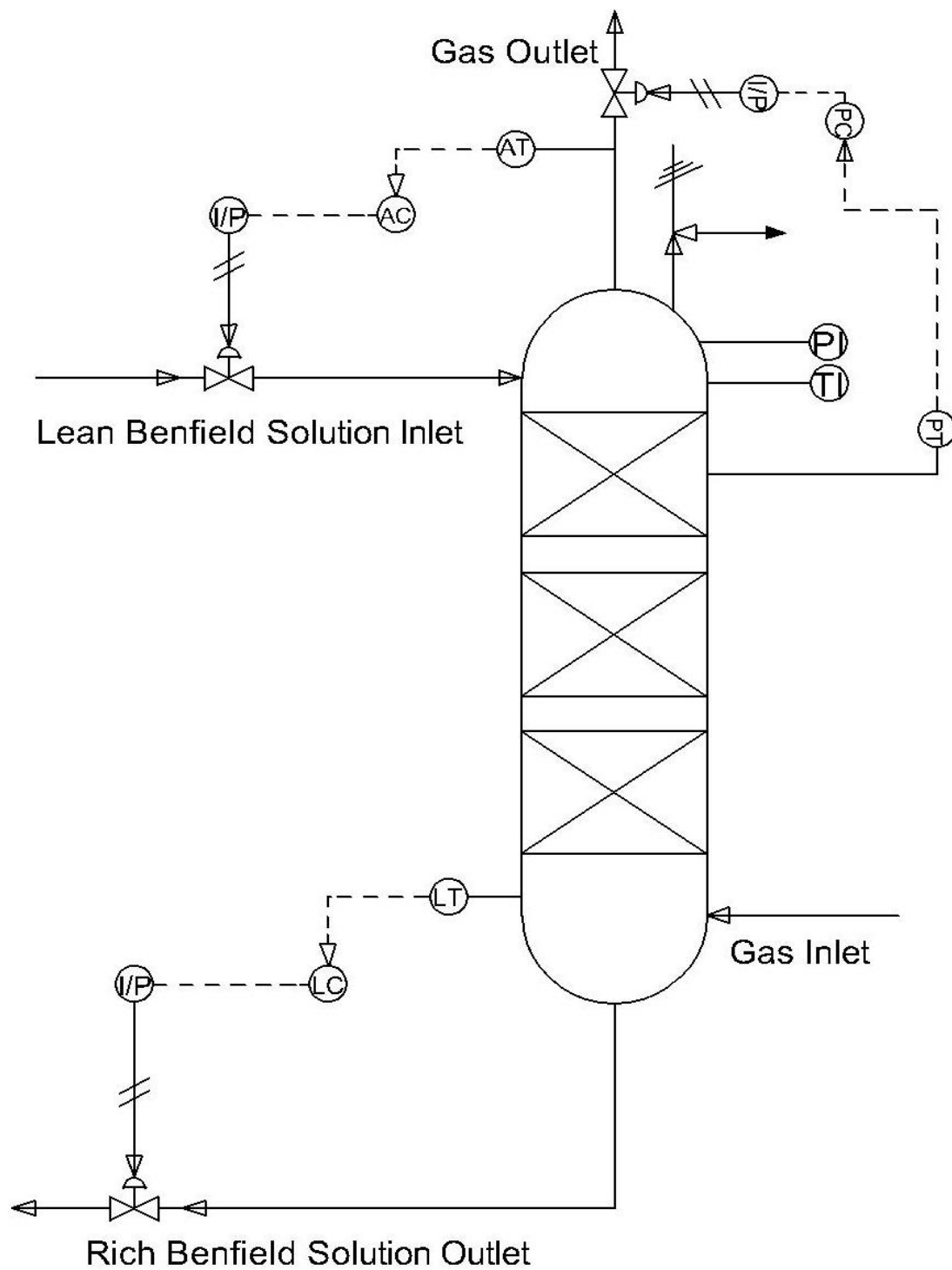


Figure 11.3.8: Piping and Instrumentation Diagram of Absorber

11.4 Ammonia Converter

11.4.1 Calculation of Major Parameters

For the ammonia production process, the type of converter used is 3 bed quench type Halder Topsoe converter where inlet feed is split into three equal parts. Each part is used as a feed to each bed. The second bed takes in the outlet product from the first bed and quenches it with $1/3^{\text{rd}}$ of the initial converter feed. Similarly, the third bed takes in outlet product from the second bed and mixes it with the final $1/3^{\text{rd}}$ of the feed.

The schematic diagram of the converted bed is given below for better representation.

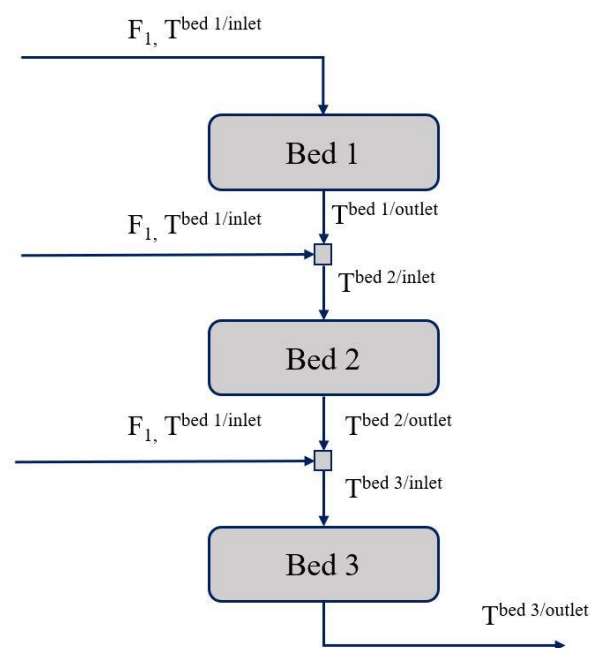


Figure 11.4.1: The converter bed, temperature and flow rate of inlet and outlet stream

The inlet and outlet temperature of each bed is preset to be able to solve the consequent design equations

Table 11.4.1: Temperature distribution of different bed of ammonia converter

$T^{\text{bed1/inlet}}$	250°C	$T^{\text{bed2/inlet}}$	300°C	$T^{\text{bed3/inlet}}$	350°C
$T^{\text{bed1/outlet}}$	350°C	$T^{\text{bed2/outlet}}$	375°C	$T^{\text{bed3/outlet}}$	450°C

Algorithm for solving each bed

- Rate equation of the ammonia conversion reaction is, (*'Chemical Reactor design for process plants'* by Rase,H.F)

$$r_A = 1.77 \times 10^{15} e^{-\frac{40765}{R \times T}} \times [K^2 \times P^{\frac{3}{2}} \left(y_N \times \frac{y_H^{\frac{3}{2}}}{y_A} - \frac{1}{P} \times \frac{y_A^{\frac{3}{2}}}{y_H^2} \right)]$$

Where r_A is the rate equation in $\frac{\text{kmol}}{\text{m}^3 \text{ cat} \times \text{hr}}$

T is the temperature in Kelvin

K is the equilibrium constant of the ammonia conversion reaction

P is the pressure in atm

y_N is the mole fraction of Nitrogen

y_H is the mole fraction of Hydrogen

y_A is the mole fraction of ammonia

- The expression for equilibrium constant is, (*'Chemical Reactor design for process plants'* by Rase,H.F)

$$\log K = -20691122 \log T - 50519265 \times 10^{-5} T + 1.848863 \times 10^{-7} T^2 + \frac{2001.6}{T} + 2.6899$$

Where T is in Kelvin

- The actual bed temperature is taken as an average of inlet and outlet temperature and after that the following design equation is used to solve for each bed.

$$\Delta W r_A = \Delta n_A$$

Where ΔW is the volume of the catalyst bed in m^3

r_A is the rate of reaction in $\frac{\text{kmol}}{\text{m}^3 \text{ cat} \times \text{hr}}$

Δn_A is the change in number of moles of ammonia

$$\Delta n_A = \frac{3}{2} \Delta n_H = \frac{1}{2} \Delta n_N$$

- Now the energy balance for each bed is done by using the following equation

$$\Delta n_A \times -\Delta H_A = \sum (n_{\text{component}}^{\text{bed number}})_{\text{exit}} \times \int_{T_{\text{inlet}}^{\text{bed number}}}^{T_{\text{exit}}^{\text{bed number}}} C_p \times dT$$

- Where $-\Delta H_A$ is the heat of reaction. The equation to find the heat of reaction is, (*'Chemical Reactor design for process plants'* by Rase,H.F)

$$\Delta H_A(T, P) = -9184 - 7.2949 \times T + 0.34996 \times 10^{-2} \times T^2 + 0.03356 \times 10^{-3} - 0.11625 \times 10^{-9} \times T^4 - (6329.3 - 3.1619P) + (14.3595 + 4.4552 \times 10^{-3} \times P) \times T - T^2(8.3395 \times 10^{-3} + 1.928 \times 10^{-6}P) - 51.251 + 0.14215 \times P \frac{\text{cal}}{\text{g} \times \text{mole Ammonia}}$$

Where T is in Kelvin and P is in atm

The Expression for specific heat capacity is taken up to 2nd power ('*Introduction to chemical engineering thermodynamics*' by J.M Smith, H.C Van Ness, M.M Abbott, 2005)

$$C_{pH} = 29.07 - 0.837 \times 10^{-3} \times T + 2.012 \times 10^{-6} \times T^2$$

$$C_{pN} = 26.98 - 5.912 \times 10^{-3} \times T - 0.3376 \times 10^{-6} \times T^2$$

$$C_{pA} = 25.88 - 32.58 \times 10^{-3} \times T - 3.046 \times 10^{-6} \times T^2$$

$$C_{pI} = 50$$

- At quench point between the two beds, the mixed stream has the component molar flow of

$$n_j^{n+1}(i) = n_j^n(e) + F_1 y_j^I$$

The temperature change is found from the energy balance equation

$$\sum n_j^n(e) \times \int_{T^{n+1}}^{T^n} C_{pj} \times dT = F_1 \times \sum (y_j \times \int_{T^{Ii}}^{T^{n+1i}} C_{pj} \times dT)$$

n_j^{n+1} = number of moles of particular component entering in next bed

T^{n+1i} = temperature of the feed entering into the next bed

Design of Bed I

The inlet condition of the converter

Stream No	28
Stream Name	Converter Inlet
Molar Flow	1090154.4 (kmol/day)
Temperature	250 degC
Pressure	210 atm

Therefore, the condition of inlet feed flow rate F1

Molar Flow	363384.8 (kmol/day)
Temperature	300 degC
Pressure	210 atm
Mole fraction of ammonia (y_A)	0.247
Mole fraction of Nitrogen (y_N)	0.171
Mole fraction of Hydrogen (y_H)	0.504
Mole fraction of Inerts (y_I)	0.076

The inlet and outlet temperature of each bed are preset to be able to solve the consequent design equations.

The volume of catalyst bed 1 is $\Delta W_1 \text{ m}^3$

Average Temperature of Bed I is $T^{\text{bed1}} = 573 \text{ K}$

So the equilibrium constant $K^I = 1.2567$

The reaction rate

$$r_A = 1.77 \times 10^{15} e^{-\frac{40765}{8.314 \times 573}} \times \left[1.6567^2 \times 210^{\frac{3}{2}} \left(0.171 \times \frac{0.504^{\frac{3}{2}}}{0.247} - \frac{1}{210} \times \frac{0.247}{0.504^{\frac{3}{2}}} \right) \right]$$

$$= 1.98 \times 10^3 \frac{\text{kmol}}{\text{m}^3 \text{ cat} \times \text{hr}}$$

So,

$\Delta W_1 \times 1.98 \times 10^3 = \Delta n_A^I$	$\Delta W_1 \times 0.99 \times 10^3 = \Delta n_N^I$	$\Delta W_1 \times 2.97 \times 10^3 = \Delta n_H^I$
---	---	---

So the change of mole number

$$n_N^I(e) = 363384.8 \times 0.171 - \Delta W_1 \times 0.99 \times 10^3$$

$$n_H^I(e) = 363384.8 \times 0.504 - \Delta W_1 \times 2.97 \times 10^3$$

$$n_A^I(e) = 363384.8 \times 0.247 + \Delta W_1 \times 1.98 \times 10^3$$

$$n_I^I(e) = 363384.8 \times 0.076$$

From equation of heat of reaction

$$\Delta H_A = -12400 \frac{\text{cal}}{\text{mol NH}_3} = -52185 \frac{\text{kJ}}{\text{kmol NH}_3}$$

So from the design equation

$$\begin{aligned} & \Delta W_1 \times 1.98 \times 10^5 \times -52185 \\ &= (363384.8 \times 0.171 - \Delta W_1 \times 0.99 \times 10^3) \\ & \times \int_{523}^{623} (26.98 - 5.912 \times 10^{-3} \times T - 0.3376 \times 10^{-6} \times T^2) dT \\ &+ (363384.8 \times 0.504 \\ &- \Delta W_1 \times 2.97 \times 10^3) \\ & \times \int_{523}^{623} (26.98 - 5.912 \times 10^{-3} \times T - 0.3376 \times 10^{-6} \times T^2) dT \\ &+ (363384.8 \times 0.247 \\ &+ \Delta W_1 \times 1.98 \times 10^3) \\ & \times \int_{523}^{623} (26.98 - 5.912 \times 10^{-3} \times T - 0.3376 \times 10^{-6} \times T^2) dT \\ &+ 363384.8 \times 0.076 \times \int_{523}^{623} (50) dT \end{aligned}$$

Solving the equation

$$\Delta W_1 = 23.61 \text{ m}^3$$

So

$$n_N^I(e) = 363384.8 \times 0.171 - 23.61 \times 0.99 \times 10^5 = 38764.90 \text{ kmol/day}$$

$$n_H^I(e) = 363384.8 \times 0.504 - 23.61 \times 2.97 \times 10^3 = 113024.23 \text{ kmol/day}$$

$$n_A^I(e) = 363384.8 \times 0.247 + 23.61 \times 1.98 \times 10^3 = 136503.54 \text{ kmol/day}$$

$$n_I^I(e) = 363384.8 \times 0.076 = 27617.24 \text{ kmol/day}$$

Design of bed II

At quench point

$$n_N^{II}(i) = 38764.90 + 363384.8 \times 0.171 = 100903.70 \text{ kmol/day}$$

$$n_H^{II}(i) = 113024.23 + 363384.8 \times 0.504 = 296167.16 \text{ kmol/day}$$

$$n_A^{II}(i) = 136503.54 + 363384.8 \times 0.247 = 226259.58 \text{ kmol/day}$$

$$n_I^{II}(i) = 27617.24 + 27617.24 = 55234.48 \text{ kmol/day}$$

From equation ----- the value of $T^{\text{bed 2/inlet}} = 300.13^\circ\text{C}$

Molar Flow	678864.92
	(kmol/day)
Temperature	337.5 degC
Pressure	225 atm
Mole fraction of ammonia (y_A)	0.333
Mole fraction of Nitrogen (y_N)	0.148
Mole fraction of Hydrogen (y_H)	0.436
Mole fraction of Inerts (y_I)	0.081

The volume of the catalyst bed required for bed II is ΔW_2

Repeating the calculation process for bed I

$$\Delta W_2 = 46.3 \text{ m}^3$$

Design of Bed III

Repeating the same calculation process

$$\Delta W_3 = 55.75 \text{ m}^3$$

Total Volume of the bed

$$\Delta W_1 + \Delta W_2 + \Delta W_3 = 23.61 + 46.3 + 55.75 = 125.66 \text{ m}^3$$

Using $\frac{L}{D} = 4:1$

$$V = \frac{\pi}{4} \times D^2 \times L = \frac{\pi}{4} \times D^3 \times \frac{L}{D} = \frac{\pi}{4} \times D^3 \times 4 = 125.66$$

$$D = 3.42 \text{ m}$$

Adding 50% to internal piping diameter, total shell diameter = $3.42 \times 1.5 = 5.12 \text{ m}$

Using $D=3.42 \text{ m}$

$$\text{Height of bed I } \frac{4 \times 23.61}{\pi \times 3.42^2} = 2.57 \text{ m}$$

$$\text{Height of bed II } \frac{4 \times 46.3}{\pi \times 3.42^2} = 5.04 \text{ m}$$

$$\text{Height of bed III } \frac{4 \times 55.75}{\pi \times 3.42^2} = 6.06 \text{ m}$$

Total height of the bed = $2.57 + 5.04 + 6.06 = 13.67 \text{ m}$

Actual height of the bed = $13.67 \times 1.2 = 16.4 \text{ m}$

Pressure drop calculation

Ergun equation,

$$\frac{\Delta P}{L} = \frac{150 \times \mu \times (1 - \varepsilon)^2}{D_e^2 \times \varepsilon} \times v_s + \frac{1.75 \times \rho \times v_s^2 \times (1 - \varepsilon)}{D_e \times \varepsilon^3}$$

Where

ΔP = Pressure drop across packed tube

L = Height of tube

ρ = density of the fluid

D_e = Equivalent spherical diameter of the particle

μ = Dynamic viscosity of the fluid

v_s = superficial velocity

ε = void Fraction

Equivalent spherical diameter $D_e = 0.003$ m

$\mu = 2.254 \times 10^{-5}$ Pa \times s

Catalyst properties [from KAFCO data sheet]

Shape-Granule (assumed to have sphericity of 1)

Size – 3mm diameter

Converter outlet mass density $\rho = 0.494 \frac{\text{kg}}{\text{m}^3}$

Average Molecular weight $M_w = 11.08 \frac{\text{kg}}{\text{kmol}}$

Volumetric flow rate $\frac{\text{molar flow rate} \times M_w}{\rho} = 35173.47 \times \frac{11.08}{0.494} = 78891 \frac{\text{m}^3}{\text{hr}}$

superficial velocity $v_s = \frac{\text{volumetric flow rate}}{\text{flow area}} = \frac{78891}{\frac{\pi}{4} \times 3.42^2} = 2.38 \frac{\text{m}}{\text{s}}$

Assuming void fraction = 0.4

By Ergun equation

$$\frac{\Delta P}{L} = 14456.23 \frac{\text{Pa}}{\text{m}} = 14456.23 \times 16.4 = 2.34 \text{ atm}$$

Mechanical Design

Thickness of shell

The minimum thickness required to resist internal pressure $t = \frac{P \times D}{2f - P}$

Where P = design pressure = 121 atm = 12 N/mm²

D = Internal diameter = 3.42m = 3420mm

f = Design stress factor = 171

$$\text{Now } t = \frac{12 \times 3420}{2 \times 171 - 12} = 178.43 \text{ mm}$$

Corrosion rate for ammonia converter is around 0.02-0.25 mm/yr (reference

www.supermetal.com)

Taking maximum corrosion rate of 0.25 mm/yr for a plant life of 20 years, corrosion

allowance $0.25 \times 2 \times 20 = 10\text{mm}$

So the total thickness = 188.43mm

Vessel head thickness calculation:

Vessel head type: Hemispherical head

Material of construction: ASME A516 Grade 70

head thickness from the equation,

$$\begin{aligned} t_h &= \frac{P_i \times D_i}{4SE - 0.4P_i} \\ &= \frac{250 \times 101.325 \times 5.12 \times 1000}{4 \times 138 \times 1000 \times 0.85 - 0.4 \times 250 \times 101.325} \\ &= 236.48 \text{ mm} \end{aligned}$$

Adopted head thickness = 240 mm = 0.24 m (including corrosion allowance)

Feed Inlet diameter:

Material of construction: ASTM A105, carbon steel

Inlet temperature, $T = 250^\circ\text{C} = 523\text{K}$

Inlet pressure, $P = 21278 \text{ kPa}$

Feed gas flowrate, $n = 32173.47 \text{ kmol/hr}$

$$\begin{aligned} \text{So, volumetric flowrate, } Q &= \frac{nRT}{P} \\ &= \frac{32173.47 \times 1000 \times 8.314 \times 5823}{3600 \times 21278 \times 1000} \\ &= 20.33 \text{ m}^3/\text{s} \end{aligned}$$

Assuming velocity, $v_n = 13.9 \text{ m/s}$

So,

$$\text{Nozzle area, } A_n = \frac{20.33}{13.9} = 1.46 \text{ m}^2$$

and nozzle diameter, $d_n = \sqrt{\frac{4 \times 1.46}{\pi}} = 1.36 \text{ m}$

Adopted nozzle diameter, $d_n = 1.4 \text{ m}$

Nozzle thickness can be determined from,

$$t_n = \frac{P_i \times D_i}{2SE - 1.2P_i}$$

$$P_i = 21278 \text{ kPa}$$

$$E = 0.85$$

$$D_i = 1360 \text{ mm}$$

$$S = 184 \text{ MPa (ASME boiler and pressure vessel code, Section II, part D)}$$

So, substituting values,

$$t_n = 13.78 \text{ mm}$$

including 2mm for corrosion allowance,

Outlet diameter:

Material of construction: ASTM A105, carbon steel

Outlet temperature, $T = 450^\circ\text{C} = 723\text{K}$

Outlet pressure, $P = 21278 \text{ kPa}$

Outlet gas flowrate, $n = 32519 \text{ kmol/hr}$

$$\begin{aligned} \text{So, volumetric flowrate, } Q &= \frac{nRT}{P} \\ &= \frac{32519 \times 1000 \times 8.314 \times 723}{3600 \times 21278 \times 1000} \\ &= 25.24 \text{ m}^3/\text{s} \end{aligned}$$

Assuming velocity, $v_n = 13.9 \text{ m/s}$

So,

$$\text{Nozzle area, } A_n = \frac{25.24}{13.9} = 1.81 \text{ m}^2$$

$$\text{and nozzle diameter, } d_n = \sqrt{\frac{4 \times 1.81}{\pi}} = 1.51 \text{ m}$$

Adopted nozzle diameter, $d_n = 1.55 \text{ m}$

Nozzle thickness can be determined from,

$$t_n = \frac{P_i \times D_i}{2SE - 1.2P_i}$$

$$P_i = 21278 \text{ kPa}$$

$$E = 0.85$$

$$D_i = 1550 \text{ mm}$$

$$S = 184 \text{ MPa (ASME boiler and pressure vessel code, Section II, part D)}$$

So, substituting values,

$$t_n = 15.23 \text{ mm}$$

including 2mm for corrosion allowance,

Support of the converter

- For a vessel with diameter 5.12m and height 16.4m for a 0.5 Mo steel ASTM A 204 grade A, the total weight is found from interpolation of the data table CS-6.4 of reference 'Chemical Reactor design for process plants' by Rase, H.F

Table CS-6.4 Converter Summary (1972 Costs)

Case	Nominal Pressure atm	ID Inner Shell, ft	Total Bed Height, ft	Total Bed ΔP , atm	Inner Shell ^a		Outer Shell ^b			Total Weight lb	Vessel Cost \$'
					Height ft	Weight lb	ID ft	Height ft	Weight lb		
1	150	10.4	49	3	55	135,147	10.7	56	944,307	1.08×10^6	756,000
2	225	7.05	58	4.6	64	100,958	7.45	65	794,448	0.895×10^6	626,500
3	300	5.6	76.3	6.1	82	105,799	6.1	83	924,911	1.031×10^6	721,700

If total weight is directly proportional to height \times ID²

$$\text{Total weight } W = 1.08 \times 10^6 \times \frac{53.8 \times 16.79}{56 \times 10.7^2} = 2.55 \times 10^6 \text{ lb} = 1.15 \times 10^6 \text{ kg}$$

Bracket system has been adopted as support of the tower and six brackets are necessary as tower diameter is 5.12 m ('Process equipment design' by Joshi, 380 pp)

Using the equation design from 'Process equipment design by Joshi' (380 pp), maximum total compressive load,

$$P = \frac{4P_w(H - L)}{nDb} + \frac{W_{\max}}{n}$$

$$\text{Wind load of the tower } P_w = 0.5\rho C_p V_a^2$$

C_p = wind pressure coefficient = 1.05 for cylindrical pressure vessel

$$\rho = \text{air density} = 1.225 \text{ kg.m}^3$$

$$V_a = \text{air velocity} = 30 \frac{\text{m}}{\text{s}} (\text{at worst condition})$$

$$P_w \text{ per unit surface area} = 578.81 = \frac{0.0059 \text{ kg}}{\text{cm}^2}$$

$$\text{Outside surface area of the tower} = \pi \times 8.395 \times 53.8 = 1418 \text{ ft}^2 = 1418 \times 0.3048^2 \times 10^4 = 1.31 \times 10^6 \text{ cm}^2$$

$$\text{So } P_w = 0.0059 \times 1.31 \times 10^6 = 7729 \text{ kg}$$

Using the chart from [4]

The height of bracket from base $L = 5\text{ft}$

The height of vessel bottom from base $= 3\text{ft}$

Height of tower from ground $= 53.8+3=17.31\text{m}$

Diameter of the bolt circle $D_b = 0.01\text{m}$

Now putting all these values

Maximum total compressive load at the most remote corner of the column from support

$$P = \frac{4 \times 7729 \times (17.31 - 1.524)}{6 \times 0.01} + \frac{1.15 \times 10^6}{6} = 8.32 \times 10^6 \text{ kg}$$

Design consideration

Main parts of the converter

- Cylindrical pressure shell
- Converter basket for catalyst
- Internal piping to allow quenching
- Pressure shell covers

Design codes

- ASME section VIII division 2(USA)
- AD-Merkblatter

Material Selection

0.5 Mo steel, ASTM A 204 grade A (*'Chemical Reactor design for process plants'* by Rase, H.F)

- Alloying with Cr and Mo reduces hydrogen attack
- Operation below 400 deg C reduces nitriding
- Hydrogen induced cracking is prevented by steel selection in which proper heat treatment has been done to achieve maximum hardness of 250 HV

11.4.2 Mechanical Drawing of Ammonia Converter

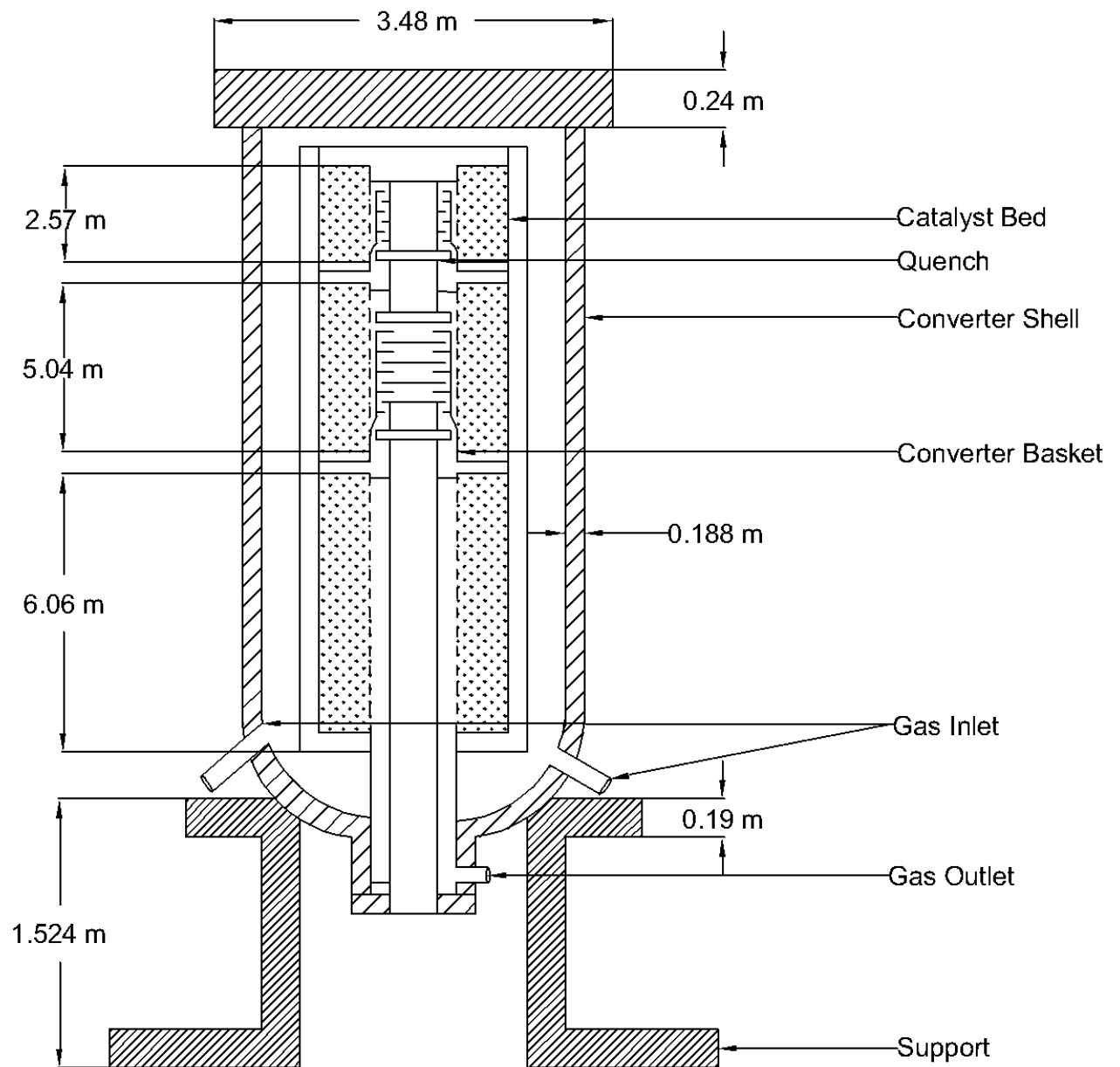


Figure 11.4.2: Mechanical drawing of the ammonia converter

11.4.3 Piping & Instrumentation Diagram of Ammonia Converter

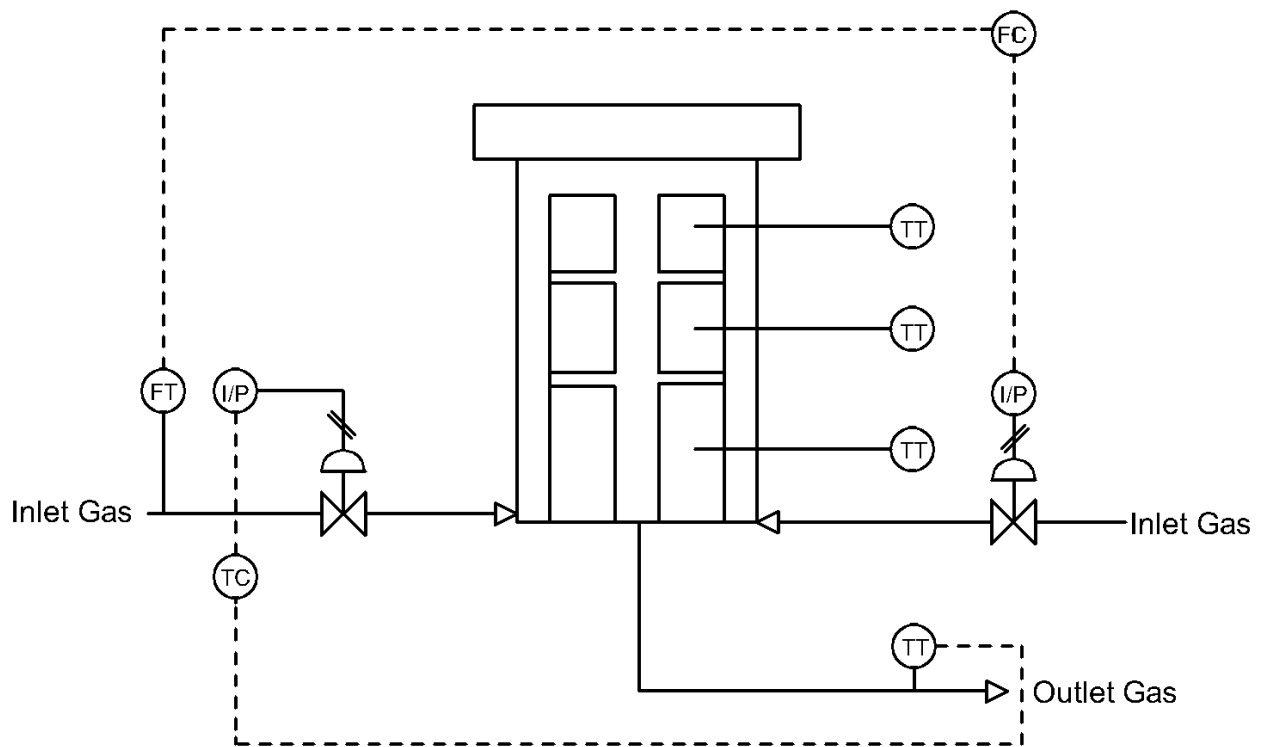


Figure 11.4.3: Piping and Instrumentation diagram of ammonia converter

11.4.4 Equipment Specification Sheet

Table 11.4.2: Specification Sheet of Ammonia Converter

Parameters		Design data
Length of catalyst bed (m)	1 st bed	2.57
	2 nd bed	5.04
	3 rd bed	6.06
Diameter (m)		5.12
Volume of catalyst bed (m ³)		125.66
Pressure vessel head thickness (mm)		56.00
Pressure vessel wall thickness (mm)		188.43
Pressure drop (atm)		2.34
Inlet flow (kmol/h)		35173.47
Outlet flow (kmol/h)		32519.08
Inlet mole composition (mole fraction)	H ₂	0.504
	N ₂	0.171
	NH ₃	0.247
	Ar	0.076
Outlet mole composition (mole fraction)	H ₂	0.418
	N ₂	0.139
	NH ₃	0.360
	Ar	0.082
Inlet and outlet temperatures of catalyst bed (K)	1 st bed inlet temperature	523
	3 rd bed outlet temperature	723

Chapter 12
ECONOMIC ANALYSIS

12.1 Determination of cost of major individual equipment

Primary Reformer (PRF)

$$\text{Heat duty } 9.07 \times 10^{10} \frac{\text{kJ}}{\text{day}} = 3.58 \times 10^9 \frac{\text{BTU}}{\text{hr}}$$

Tube internal pressure = 518 psia

Cost of primary reformer in 2014 = \$ 43349400 (*Matches' Reactor Cost - Autoclave, Fermenter, Kettle, Mixer Settler.*, n.d.)

$$\begin{aligned} \text{Total cost of primary reformer in 2023} &= \frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014} \\ &= \frac{800.6}{576.1} \times 43349400 = \$ 6.02 \times 10^7 \end{aligned}$$

Secondary Reformer

Assuming secondary reformer cost is 1/3 of primary reformer

Total cost of secondary reformer in 2023 = $6.02 \times 10^7 \times \frac{1}{3} = \$ 2.01 \times 10^7$ (Inc & Francisco, 2006)

High Temperature shift converter

Volume of high temperature shift converter = $24.75 \text{ m}^3 = 6538.25 \text{ gal}$

Cost of High temperature shift converter in 2014 = \$137700 (*Matches' Reactor Cost - Autoclave, Fermenter, Kettle, Mixer Settler.*, n.d.)

$$\begin{aligned} \text{Total cost of high temperature shift converter in 2023} &= \frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014} \\ &= \frac{800.6}{576.1} \times 137700 = \$1.91 \times 10^5 \end{aligned}$$

Low Temperature Shift converter

Total volume of the low temperature shift converter = 73.25 m^3

Total cost of low temperature shift converter = $\$ 5.64 \times 10^5$ (*Matches' Reactor Cost - Autoclave, Fermenter, Kettle, Mixer Settler.*, n.d.)

Carbon Dioxide Absorber

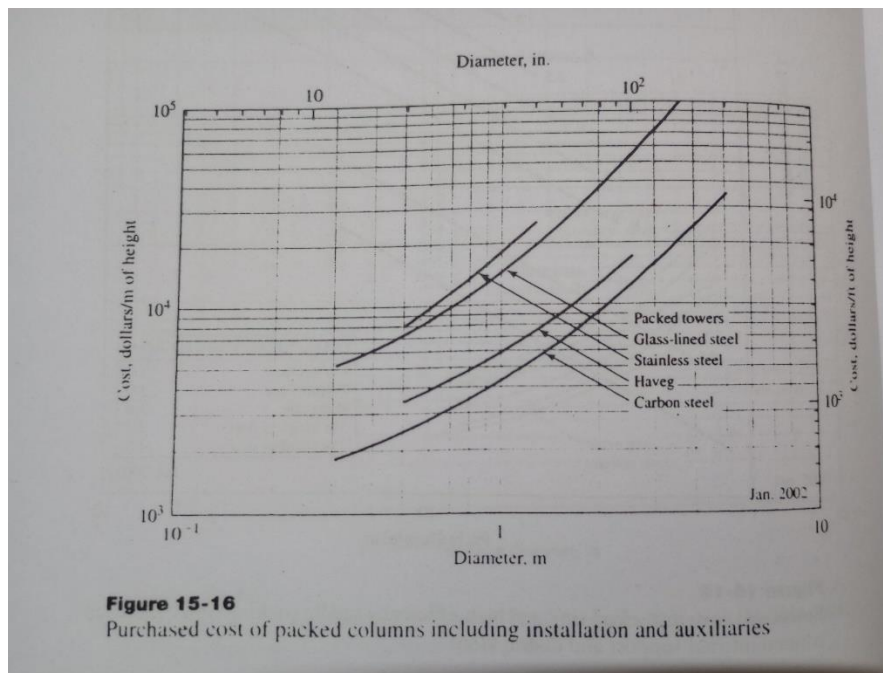


Figure 12.1.1 : Plot of column cost estimation

The designed column contains Raschig ring and the material of construction is carbon steel.

Column height = 18.86 m

Column diameter = 2.72 m

From graph, cost (in 1991) = $(14800 \times 18.86) = \$ 279128$ (Peters et al., 2003b)

Total cost = cost index 2023 /cost index 1991 = $\$(800.6/361.3) \times 279128 = \$ 6.19 \times 10^5$
(CEPCI_2002_2.Pdf, n.d.)

Regenerator

Regeneration tower height = 21.69 m

Diameter = 3.13 m (Both height and diameter is assumed to be 1.15 times the absorber dimensions)

From graph, cost (in 1991) = $(17300 \times 21.69) = \$ 375237$

Total cost = (cost index 2023/cost index 1991) $\times \$ 375227 = (800.6/361.3) \times 375237 = \$ 831482$
(CEPCI_2002_2.Pdf, n.d.)

Methanator

Total volume of the Methanator reactor = $11.85 \text{ m}^3 = 3130.43 \text{ gal}$

Total cost of methanator reactor in 2014 = \$ 110300

Total cost of Methanator reactor in 2023 = $\frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014}$

$$= \frac{800.6}{576.1} \times 110300 = \$ 1.53 \times 10^5$$

(Matches' Reactor Cost - Autoclave, Fermenter, Kettle, Mixer Settler., n.d.)

Ammonia Converter

Total inner shell diameter of the ammonia converter = 5.13 m

Total height of the ammonia converter = 16.4 m

Volume of the ammonia converter = $337.65 \text{ m}^3 = 89197.69 \text{ gal}$

Total cost of ammonia converter in 2014 = 348900 \$ (Matches' Reactor Cost - Autoclave, Fermenter, Kettle, Mixer Settler., n.d.)

$$\begin{aligned} \text{Total cost of ammonia converter in 2023} &= \frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014} \\ &= \frac{800.6}{576.1} \times 348900 = \$ 4.85 \times 10^5 \end{aligned}$$

Purge Gas Recovery Unit (PGR unit)

This unit includes-ammonia absorber, distillation column, cold-box, adsorber and auxiliary units.

Total installation cost of PGR unit 2013= \$ 666000 (Hossain et al., 2013)

$$\begin{aligned} \text{Total installation cost of PGR unit 2023} &= \frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014} \\ &= \frac{800.6}{576.1} \times 666000 = \$ 9.26 \times 10^5 \end{aligned}$$

(2 Updating the CE Plant Cost Index - The Chemical Engineering Plant Cost Index (CEPCI) is an - Studocu, n.d.)

Pump

Cost of total 7 centrifugal pump in 2014 = $6300 \times 7 = \$ 4.41 \times 10^4$

(Matches' Centrifugal Pump Cost - Horizontal, Vertical, ANSI, API-610., n.d.)

$$\begin{aligned} \text{Total cost of centrifugal pump in 2023} &= \frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014} \\ &= \frac{800.6}{576.1} \times 4.41 \times 10^4 = \$ 6.13 \times 10^4 \end{aligned}$$

Compressor

Cost of total 2 compressor of power 7731 kW used in the plant in 2014 = \$153600

(Matches' Compressor Cost - Centrifugal, Reciprocating, Rotary Screw., n.d.)

$$\text{Total cost of 2 pumps in 2023} = \frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014}$$

$$= \frac{800.6}{576.1} \times 153600 = \$ 2.13 \times 10^5$$

Heat Exchanger

Surface area of total 7 heat exchanger = $422.24 \text{ m}^2 = 4544.95 \text{ ft}^2$

Total cost of Heat exchanger in 2014 = \$ 95600

(Matches' Heat Exchanger Cost - Air-Cooled, Kettle, Shell and Tube, Tank, Water., n.d.)

Total cost of heat exchanger in 2023 = $\frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014}$

$$= \frac{800.6}{576.1} \times 95600 = \$ 1.13 \times 10^5$$

Ammonia Storage Tank

Volume of the ammonia storage tank = $3781.82 \text{ m}^3 = 999098.7 \text{ gal}$

Cost of Ammonia storage tank in 2014 = \$ 213400

(Matches' Tank Cost - API, Horizontal, Vertical, Cone Roof, Bottom, Flat Bottom, Round End., n.d.)

Total cost of Ammonia storage tank in 2023 = $\frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014}$

$$= \frac{800.6}{576.1} \times 213400 = \$ 2.97 \times 10^5$$

Separator

Pre-Absorption Separator

Diameter of the pre-Absorption separator = 1.153 m = 45.4 in

Total cost of the pre absorption separator in 2014 = \$ 13300

(Matches' Separator Cost - Air-Sweep, Cyclone, Gas-Liquid Vane, Screw Classifier., n.d.)

Total cost of separator in 2023 = $\frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014}$

$$= \frac{800.6}{576.1} \times 13300 = \$ 1.85 \times 10^4$$

Pre-compression Separator

Diameter of the pre compression separator = 1.255 m = 49.40 in

Total cost of the precompression separator in 2014 = \$ 14200

(Matches' Separator Cost - Air-Sweep, Cyclone, Gas-Liquid Vane, Screw Classifier., n.d.)

Total cost of separator in 2023 = $\frac{(\text{cost index in 2023})}{(\text{cost index in 2014})} \times \text{Total cost in 2014}$

$$= \frac{800.6}{576.1} \times 14200 = \$ 1.97 \times 10^4$$

Total cost of separator = $\$ 3.82 \times 10^4$

12.2 Determination of Total Capital Investment

Table 12.2.1: Total cost of Purchased Equipment

Name of the Equipment	Total Cost in $\$ \times 10^3$
Primary Reformer	60200
Secondary Reformer	20100
High Temperature Shift Converter	191
Low Temperature Shift Converter	564
Methanator	153
Carbon-di-oxide Absorber	619
Regenerator	831
Ammonia Converter	485
Compressor	213
Heat Exchanger	113
Pump	613
Separator	38
Purge Gas Recovery Unit	926
Ammonia storage tank	297
Catalyst and package unit cost	100
Total purchased equipment cost	85443

Table 12.2.2: Estimation of capital Investment

	Factor	Cost in million dollars
Direct Costs		
Purchased equipment delivered	1.00	85.443
Purchased equipment installation	0.47	40.158
Instrumentation and control	0.36	30.759
Piping	0.68	58.101
Electrical systems	0.11	9.398
Buildings	0.18	15.380
Yard improvements	0.10	8.544
Service facilities	0.70	59.810
Total direct plant cost	3.60	307.590
Indirect Costs		
Engineering and supervisions	0.33	28.196
Construction expenses	0.41	35.032
Legal expenses	0.04	3.417
Contractor's fee	0.22	18.797
Contractor's fee	0.44	37.595
Total indirect costs	1.44	123.040
Fixed capital investment	5.04	430.630
Working capital Investment	0.89	76.044
Total capital investment	5.93	506.680

12.3 Determination of Operating cost

The total product cost = C

Estimation of total product cost

1. Manufacturing cost = Direct production cost + fixed charges + plant overhead cost

A. Direct production costs

Table 12.3.1: Direct production costs

Components	Cost \$
Raw materials (10-80 % of total product cost)	0.28C
Operating labor (10 – 20 % of total product cost)	0.13C
Direct supervisory and clerical labor (10-20 % of total product cost)	0.10C
Utilities (10-20 % of total product cost)	0.12C
Maintenance and repairs (2-10 % of total product cost)	0.04C
Operating supplies (0.5 - 1 % of total product cost)	0.005C
Laboratory charges (10-20 % of operating labor cost)	0.013C
Patents and royalties (0-6 % of total product cost)	0.005C

B. Fixed charges (10-20% of total product costs)

- Depreciation (depends on life period, salvage value, and method of Calculation, about 10% of fixed capital investment for machinery and equipment and 2-3% of building value for buildings) = $0.07FCI$
- Local taxes (1-4% of fixed capital investment) = $0.01 FCI$
- Insurance (0.4-1% of fixed capital investment) = $0.004 FCI$

C. Plant overhead cost (5-15 % of total product cost) = $0.075C$

- (60-70% of cost for operating labor, supervision, and maintenance, or 5-

15% of total production cost); includes costs for the followings: general plant upkeep and overhead, payroll overhead, packaging, medical services, safety and protection, restaurants, recreation, salvage, laboratories, and storage facilities)

Thus, Manufacturing cost = $0.768C + 0.084FCI$

D. General expenses = administrative costs + distribution and selling costs + R & D costs

i. Administrative costs (15- 20 % of operating labor cost) = $0.02C$

ii. Distribution and selling costs (2 - 20 % of total product cost) = $0.04C$

iii. R & D costs (about 5 % of total product cost) = $0.02C$

iv. Financing (0 – 10 % of total capital investment) = $0.005 TCI$

Thus, general expenses = $0.08 C + 0.005 TCI$

Hence, Total production cost = manufacturing cost + general expenses

Or, $C = 0.768 C + 0.084 FCI + 0.08 C + 0.005 TCI$

Or, $C = 0.768 \times C + 0.084 \times 430632720 + 0.08 \times C + 0.005 \times 506676990$

Or, $C = 0.848 C + 38706534$

Or, $C = \$ 254648250 = \$ 254.65 \times 10^6 / yr$

\therefore Total production cost = $\$ 254.65 \times 10^6 / yr$

(Assuming this production cost includes raw materials cost, utility cost, maintenance cost etc.)

12.4 Determination of Cashflow Diagram

Calculation of Depreciation cost by sinking fund method

Project Life = 20 years

Salvage value at the end of the project life = 10% of the Fixed Capital Investment

$$= 0.1 \times \$ 430632720$$

$$= \$ 43.06 \times 10^6$$

$$\therefore \text{Depreciation per year} = \frac{(506.68 \times 10^6 - 43.06 \times 10^6)}{20} = \$ 23.18 \times 10^6$$

Determination of Revenue

Calculation of total income from sales = 1200 ton/day = 396000 ton/year

Production of CO₂ = 1373 ton/day = 453090 ton/year

Market price of ammonia = \$ 798 /ton

$$\text{Revenue by selling ammonia} = 798 \times 396000 = \$ 316.01 \times 10^6 \text{ /year}$$

(Global Chemical and Petrochemicals, Specialty Chemicals, Elastomer and Rubber, Fertilizer and Feedstock - Latest Chemical Prices, News and Market Analysis | ChemAnalyst, n.d.)

Market price of CO₂ = \$ 73/ton

$$\text{Revenue by selling CO}_2 = 73 \times 453090 = \$ 33.08 \times 10^6 \text{ /year}$$

(Carbon Dioxide Price per Ton June 2022 - News and Statistics - IndexBox, n.d.)

So, The total sales = \$ 349.09 × 10⁶ /year

Gross Annual earning = Total Annual Sales Revenue –

Total Annual Production cost –

Depreciation

$$= 349.09 \times 10^6 - 254.65 \times 10^6 - 23.18 \times 10^6$$

$$= \$ 71.17 \times 10^6$$

Assuming income tax = 25% of gross profit

$$\text{Net profit} = 71.17 \times 10^6 - (0.25 \times 71.17 \times 10^6) = \$ 53.37 \times 10^6$$

So, the plant will be operating on net profit of 53.37 million dollars

$$\begin{aligned} \therefore \text{Net annual cash flow} &= \text{Net profit} + \text{Depreciation} = \$ 53.37 \times 10^6 + \$ 23.18 \times 10^6 \\ &= \$ 76.55 \times 10^6 \end{aligned}$$

∴ Net annual cash flow for the cash flow diagram of the plant will be 76.55 million dollars.

A summary of production cost and selling price is given in Table 12.4.1

Table 12.4.1: Production cost and selling price of ammonia/ton

Production cost of ammonia per ton	\$ 643.06
Selling price of ammonia per ton	\$ 798

Calculation of Rate of Return

Net profit = \$ 53.37×10^6

Percent rate of return = $\frac{53.37 \times 10^6}{506.68 \times 10^6} \times 100 = 10.53\%$

Cash flow diagram

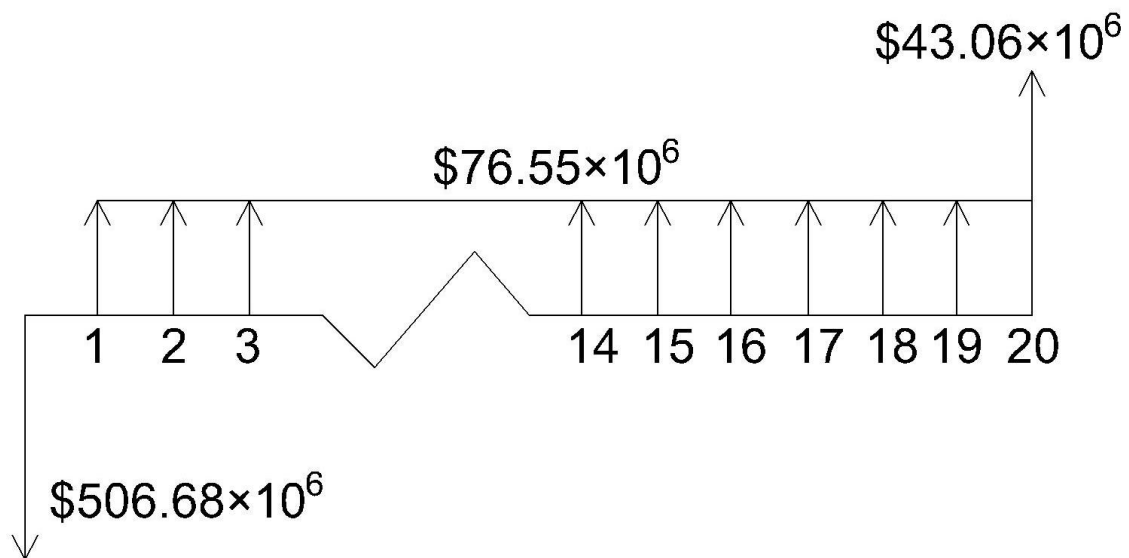


Figure 12.4.1: Cash flow diagram

12.5 Feasibility Analysis

12.5.1 IRR method

Let, MARR = 9 %

Useful life of an ammonia plant is usually 20 years. So, $n = 20$

PW = - Total capital investment + Annual income (P/A, i' %, 20) + Salvage value (P/F, i' %, 20)

So, $0 = -506.68 \times 10^6 + 76.55 \times 10^6 (P/A, i'\%, 20) + 43.06 \times 10^6 (1+i')^{-20}$

Or, $0 = -506.68 \times 10^6 + 76.55 \times 10^6 \frac{[(1+i')^{20}-1]}{i'(1+i')^{20}} + 43.06 \times 10^6 (1+i')^{-20}$

After trial and error, $i' = 0.14 = 14 \%$

\therefore The plant is economically justified because $IRR > MARR$

12.5.2 ERR method

Let, MARR = 9%

Present worth of cash outflows = Total capital investment = \$ 506.68×10^6

Future worth of cash inflows = Salvage value + FW of annual return

$= 43.06 \times 10^6 + 76.55 \times 10^6 (F/A, 9\%, 20)$

$= 43.06 \times 10^6 + 76.55 \times 10^6 \frac{[(1+0.09)^{20}-1]}{0.09}$

$= \$ 3959 \times 10^6$

Now, $F = P (1 + i'')^n$

$3959 \times 10^6 = 506.68 \times 10^6 (1 + i'')^{20}$

$\therefore i'' = 10.8\%$

Since $ERR > MARR$, the project is economically justified.

12.5.3 Break even time

Break even time = $\frac{506.68 \times 10^6}{76.55 \times 10^6} = 6.62$ years

12.5.4 Spider Plot

Spider plot is drawn to show the sensitivity of the project with changing of capital investment, annual revenue, salvage value and useful life. These parameters are being changed from -50% to +50% and corresponding relative present worth is calculated. The obtained spider plot is shown in Figure 12.5.1.

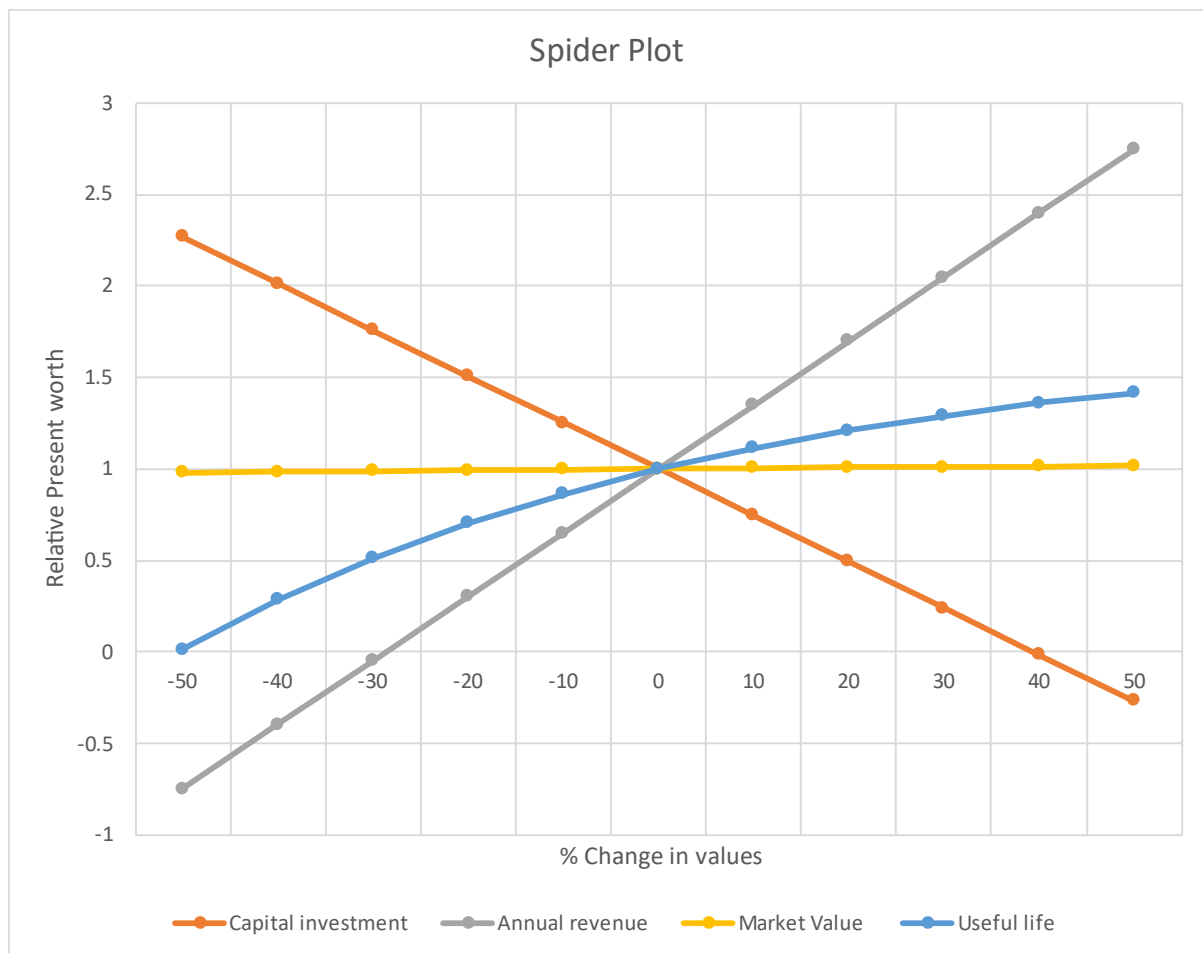


Figure 12.5.1: Spider plot of the project

From the spider plot, it can be seen that the project is sensitive to annual revenue change, capital investment change and useful life change. Sensitivity to annual revenue change is higher than the others as slope of the graph is higher. The project is insensitive to market change.

12.5.5 Payback Period

Payback period is calculated using cumulative cashflow. The cumulative cashflow is given in Table 12.5.1.

Table 12.5.1: Determination of cumulative PW

Year	Cashflow (in million dollars)	Cumulative PW (in million dollars)
0	-506.68	-506.70
1	76.55	-436.50
2	76.55	-372.00
3	76.55	-312.90
4	76.55	-258.70
5	76.55	-208.90
6	76.55	-163.30
7	76.55	-121.40
8	76.55	-82.99
9	76.55	-47.74
10	76.55	-15.41
11	76.55	14.26
12	76.55	41.47
13	76.55	66.44
14	76.55	89.35
15	76.55	110.37
16	76.55	129.65
17	76.55	147.33
18	76.55	163.56
19	76.55	178.45
20	119.61	199.79

From the above calculations, payback period comes 11 years as cumulative present worth becomes positive at EOY = 11.

Chapter 13
EQUIPMENT LAYOUT AND PLOT PLAN

13.1 Equipment Layout

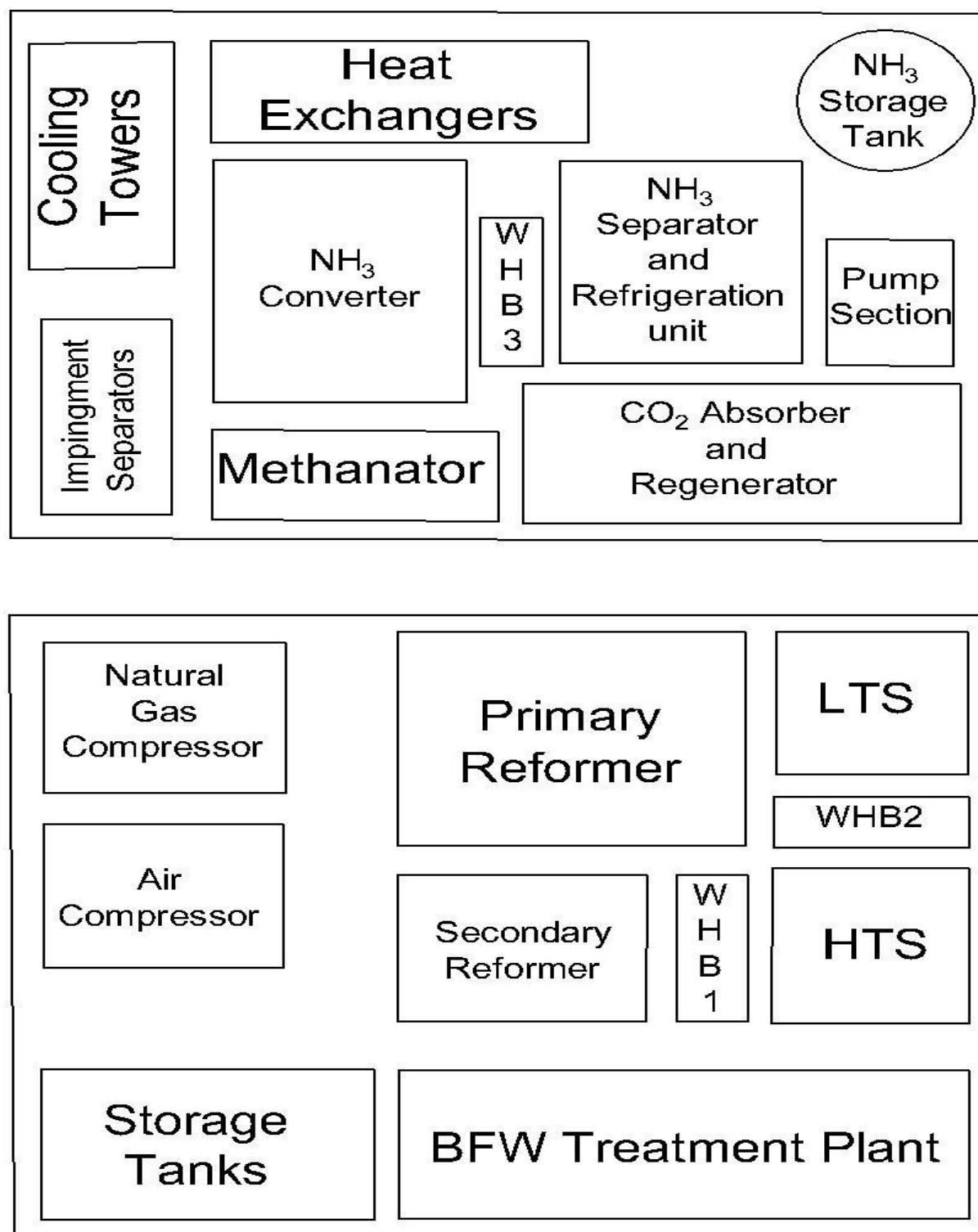


Figure 13.1.1: Equipment layout of the plant

13.2 Plot Plan

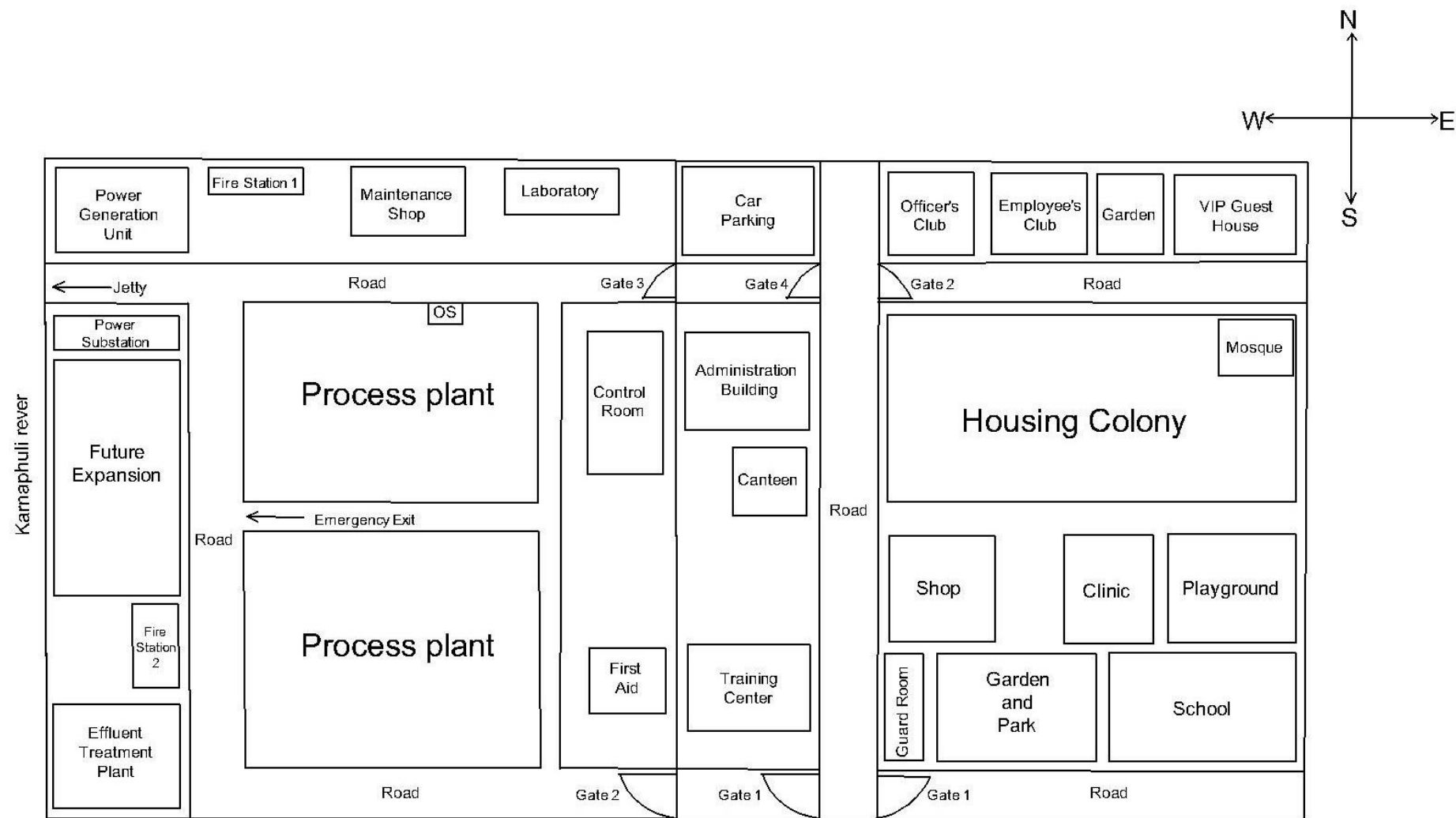


Figure 13.2.1: Plot Plan

Conclusion

As ammonia plays an important role in producing N-based fertilizers, production of ammonia should be increased. A 1200 TPD ammonia production plant has been designed based on this insight. The plant is chosen at Rangadia, Chittagong, near Karnaphuli river because of availability of raw material, ease of transportation and availability of utility facilities within this area. By doing this design project, practical knowledge about process designing has been gained. Process flow diagram has been drawn in detail. 4 major equipment such as primary reformer, low temperature shift reactor, CO₂ absorber and ammonia converter have been designed. Their mechanical drawings, P & IDs are provided. Economic analysis has been done to check the feasibility of this plant. Equipment layout and plot plan are also included in this report to show the battery limit of this plant. Plot plan is drawn based on the assumed area of 75 acres.

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Appendices

Appendix A

Calculations for Sizing of Equipment

Secondary Reformer

Total inlet molar flow rate = 202263.2 kmol/day = 2341 mol/s

Inlet temperature = 775°C

Inlet pressure = 32.2 kg/cm² G

$$\therefore \text{Volumetric flow rate} = \frac{2341 \times 8.314 \times (775 + 273)}{3157741.3 + 1.01325 \times 10^5} = 6.26 \text{ m}^3/\text{s}$$

Assuming superficial velocity = 0.3048 m/s

$$\text{Cross sectional area of SRF, } A = \frac{6.26}{0.3048} = 20.54 \text{ m}^2$$

$$\therefore \text{Bed diameter of SRF, } D = 2 \times \sqrt{\frac{20.54}{3.14}} = 5.11 \text{ m}$$

Catalyst specification:

[**Reference:** A. A. AL-Dhfeery and A. A. Jassem, “Evaluation Performance of Different Types Catalysts of an Industrial Secondary Reformer Reactor in the Ammonia Plants,” *Mod. Res. Catal.*, vol. 01, no. 03, pp. 43–51, 2012, doi: 10.4236/mrc.2012.13006.]

Type- RKS-2-7H

Shape- Cylinder with 7 holes

Dimension (OD×ID×H)- 16 mm×3 mm×11 mm

Base component- NiO

Carrier- MgAl₂O₄

Bulk density- 1000 kg/m³

Porosity- 46.52%

Using the equation, $R_{CH_4} \times W = M_{CH_4} \times X_{CH_4}$

Where, rate of reaction of CH₄, $R_{CH_4} = 9.4 \times 10^{-3}$ kmol/kg cat. hr

inlet mass flow rate of CH₄, $M_{CH_4} = \frac{6775.82 \times 16}{24} = 4517.21$ kg/hr

conversion of CH₄, $X_{CH_4} = 0.918$

$$\therefore \text{Weight of catalyst used, } W = \frac{4517.21 \times 0.918}{9.4 \times 10^{-3}} = 441148 \text{ kg}$$

$$\therefore \text{Length of catalyst bed, } L = \frac{W}{1000 \times A} = \frac{441148}{1000 \times 20.54} = 21.47 \text{ m}$$

Methanator

Inlet molar flow rate of methanator = 6049.757 kmol/h

Methanator inlet temperature = 363°C

Methanator inlet pressure = 26 kg/cm² G

$$\therefore \text{Volumetric flow rate} = \frac{6049.757 \times 10^3 \times 8.314 \times (363 + 273)}{(26 + 1) \times 10^5} = 11.85 \times 10^3 \text{ m}^3/\text{h}$$

From literature, space velocity = 1000 h⁻¹

[Reference: Technology and Manufacture of Ammonia by Samuel Strelzoff]

$$\therefore \text{Volume of methanator} = \frac{11.85 \times 10^3}{1000} = 11.85 \text{ m}^3$$

Impingement separator

Souder Brown equation for separator

$$V_a = k \sqrt{\frac{\rho_l - \rho_v}{\rho_v}}$$

V_a = Maximum allowable superficial velocity across inter face of mesh $\frac{\text{ft}}{\text{s}}$

k = Constant for separator based on application

ρ_l = liquid density $\frac{\text{lb}}{\text{ft}^3}$

ρ_v = vapor density $\frac{\text{lb}}{\text{ft}^3}$

Sizing of pre-absorption separator S101

Density of liquid stream $\rho_l = 59.84 \frac{\text{lb}}{\text{ft}^3}$ at 101°C and 27.52 bar (separator condition)

Density of Vapor stream $\rho_v = 1.14 \frac{\text{lb}}{\text{ft}^3}$ (assumed at separator condition)

$$V_a = \sqrt{\frac{(59.84 - 1.14)}{1.14}} = 7.175 \frac{\text{ft}}{\text{s}}$$

Design vapor velocity $V_d = 0.75 V_a = 5.38 \frac{\text{ft}}{\text{s}} = 1.63 \frac{\text{m}}{\text{s}}$

Total molar flow rate of vapor stream = $7349.944 \frac{\text{kmol}}{\text{h}}$

Average molecular weight of vapor stream

$$= \frac{1325.184}{7349.944} \times 44 + \frac{21.82}{7349.944} \times 28 + \frac{4419.52}{7349.944} \times 2 + \frac{23.12}{7349.944} \times 16 + \frac{1541.39}{7349.944} \times 28$$

$$+ \frac{18.91}{7349.944} \times 40 = 15.244 \frac{\text{kg}}{\text{kmol}}$$

Volumetric flow rate of vapor stream

$$= 7349.944 \frac{\text{kmol}}{\text{h}} \times \frac{24\text{h}}{1\text{day}} \times 15.244 \frac{\text{kg}}{\text{kmol}} \times \frac{1}{18.26} \frac{\text{m}^3}{\text{kg}} = 1.47 \times 10^5 \text{ m}^3/\text{day}$$

$$\text{Area} = \pi \times r^2 = \frac{1.47 \times 10^5}{1.63 \times 24 \times 3600} \text{ m}^2 = 1.045 \text{ m}^2$$

Radius of the separator $r = 0.576 \text{ m}$

Diameter $D = 2 \times 0.576 = 1.153 \text{ m}$

$$\frac{L}{D} = 4$$

$L = 4.61 \text{ m}$

$$\text{Volume} = \frac{\pi \times D^2 \times L}{4} = \frac{(\pi \times 1.153^2 \times 4.61)}{4} = 4.81 \text{ m}^3$$

Sizing of pre-compression separator S102

Density of liqued stream $\rho_l = 57.36 \frac{\text{lb}}{\text{ft}^3}$ at 300°C and 14.7 bar (separator condition)

Density of Vapor stream $\rho_v = 1.12 \frac{\text{lb}}{\text{ft}^3}$ (assumed at separator condition)

$$V_a = \sqrt{\frac{(57.36 - 1.12)}{1.12}} = 7.08 \frac{\text{ft}}{\text{s}}$$

Design vapor velocity $V_d = 0.75 V_a = 5.31 \frac{\text{ft}}{\text{s}} = 1.618 \frac{\text{m}}{\text{s}}$

Total molar flow rate of vapor stream = $5887.594 \frac{\text{kmol}}{\text{h}}$

Average molecular weight of vapor stream

$$= \frac{0.5}{5887.594} \times 44 + \frac{0.44}{5887.594} \times 28 + \frac{4257.36}{5887.594} \times 2 + \frac{69}{5887.594} \times 16 + \frac{4257.36}{5887.594} \times 28$$

$$+ \frac{18.91}{5887.594} \times 40 = 22.01 \frac{\text{kg}}{\text{kmol}}$$

Volumetric flow rate of vapor stream

$$= 5887.594 \frac{\text{kmol}}{\text{h}} \times \frac{24\text{h}}{1\text{day}} \times 22.01 \frac{\text{kg}}{\text{kmol}} \times \frac{1}{17.94} \frac{\text{m}^3}{\text{kg}} = 1.73 \times 10^5 \text{ m}^3/\text{day}$$

$$\text{Area} = \pi \times r^2 = \frac{1.73 \times 10^5}{1.618 \times 24 \times 3600} \text{ m}^2 = 0.4 \text{ m}^2$$

Radius of the separator $r = 0.627$ m

Diameter $D = 2 \times 0.576 = 1.255$ m

$$\frac{L}{D} = 4$$

$L = 5.02$ m

$$\text{Volume} = \frac{\pi \times D^2 \times L}{4} = \frac{(\pi \times 1.255^2 \times 5.02)}{4} = 6.2 \text{ m}^3$$

Heat Exchanger

The heat exchanger is used to cool the gas coming from Secondary reformer before it enters HTS. The data are taken from HYSYS,

Total heat transferred in the process, $Q = 6.719 \times 10^7$ kJ/h

LMTD = 220°C

Overall heat transfer coefficient, $U = 2991$ kJ/ h m^2C

Total surface, $A = 6.719 \times 1075064 \times 220$
 $= 60.32 \text{ m}^2$

Pump

The pump is used to carry the lean Benfield solution from regenerator to the absorber section.

$\rho = 1297.9$ kg/ m^3

$h = 13$ m

$\Delta P = (\rho gh)/10^5 = 1.655$ bar

Power required = $Q \times \Delta P$

$$= 317893.51000 \times 3600 \times 1.655 \times 105$$
$$= 14.62 \text{ kW}$$

Pump shaft power required = $\frac{\text{theoretical power required}}{\text{efficiency}} = \frac{14.62}{0.75} = 19.5 \text{ kW}$

Compressor

This compressor is used to increase the pressure of the air before it goes to secondary reformer.

A compressor sizing calculator is used for sizing.

The input values of the calculator are:

Suction pressure = 1.01 bar

Discharge pressure = 31.58 bar

Suction temperature = 28°C

Compressibility factor, $Z = 1$

Molecular weight = 29

$$C_p/C_v (k) = 1.4$$

Required flowrate = 61112.05 kg/h

The output from the calculator is:

Selected stage number : 3	Compressor Ratio Per Stage : 3.1		
	Stage 1	Stage 2	Stage 3
Suction pressure (bar a)	1.01325	3.1	9.7
Discharge pressure (bar a)	3.1	9.7	31.6
Adiabatic discharge temperature (°C)	143	143	148
Actual volumetric flow (m ³ /h)	52062	16794	5418
Isentropic stage power (kW)	1952	1952	2043

Figure A.1: Compressor sizing result from calculator

Total absorbed power = 7731 KW

(Reference: <https://www.cstfirenze.com/knowledge-center/tools-andcalculators/reciprocating-compressors-sizing>)

Storage Tank

Ammonia production rate = 70594 kmol/day

If ammonia is stored at atmospheric pressure below boiling point,
Let,

Pressure, P = 1.1 atm

Temperature, T = -33°C

Density of liquid ammonia, $\rho = 680 \text{ kg/m}^3$

So, Volumetric flowrate of ammonia, $V = \frac{m}{\rho}$

$$= \frac{70594 \times 17}{680}$$

$$= 1764.85 \text{ m}^3/\text{day}$$

Let storage tank duration is 36 hours and 70% of tank space occupied.

So, Required Volume, $V = \frac{1764.85 \times 1.5}{0.7}$

$$= 3781.82 \text{ m}^3$$

If tank height is H and diameter is D,

Then

$$\frac{\pi D^2 H}{4} = 3781.82 \text{ m}^3$$

Assuming $H/D = 2$

$$H = 16.9\text{m}$$

$$D = 8.45\text{m}$$

High Temperature Shift Reactor (HTS)

Volume per hour of feed = $252169.8 \text{ kmol/day} = 10507075 \text{ mol/hr} = 17326 \text{ m}^3/\text{hr}$

Assuming, Liquid Hourly Space Velocity (LHSV), $Z = 700 \text{ hr}^{-1}$

So,

$$\begin{aligned} \text{Catalyst Volume} &= \frac{\text{Volume per hour of feed}}{\text{LHSV}} \\ &= \frac{17326}{700} \\ &= 24.75 \text{ m}^3 \end{aligned}$$

So,

$$\frac{\pi D^2 H}{4} = 24.75 \text{ m}^3$$

and assuming $H/D = 2.5$

$$D = 2.33\text{m}$$

$$\text{and } H = 5.82\text{m}$$

Workload Distribution

Project title: Design of a 1200 TPD Ammonia Production Plant			Supervisor: Dr. Nahid Sanzida	
Student ID	Name	Individual Work	Individual Equipment Design	Signature
1702032	Md. Mostafizur Rahman	PBD, PFD Environmental Management Plan, Sizing of impingement separator and ammonia separator, Equipment pricing	Ammonia Converter	
1702040	Zulkar Nyene Bhuyan	Energy balance, Sizing of HTS and all storage tanks	Low Temperature Shift Reactor	
1702042	Sadman Islam	Material Balance, Utility balance, Sizing of heat exchangers, pumps and compressors, Economic analysis	CO ₂ Absorber	
1702047	Tasfia Ferdoush	Introduction, Project definition, Selection of process and Process Description, Design Basis, Sizing of secondary reformer and methanator, Equipment layout and Plot plan	Primary Reformer	