

**FEED GAS PRE-HEATER EFFICIENCY
CALCULATION AND ENHANCEMENT OF
AMMONIA PLANT**



By

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DEDICATION

TO

OUR TEACHERS, PARENTS AND FRIENDS WHO ENABLED US
TO BE ENGINEERS

Acknowledgements

We praise Almighty Allah, the sole benefactor of this universe, who guided us through most difficult times and granted us the knowledge and enlightened us through Holy Quran.

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In the end we thank all our friends and teachers for the help and guidance throughout.

Motivation

Pakistan is an agricultural country and exports its products worldwide. Hence Pakistan's economy is largely dependent on the agricultural products. This makes production of fertilizer very critical to Pakistan's economy. Fertilizer's demand over the years has also drastically increased with increasing export of Agro-based products.

Ammonia being the foundation of nitrogen fertilizer is of immense importance. It can be directly added to the soil or used to produce other fertilizers like urea.

Natural gas is the main raw material in the production of ammonia process. It is used as feed and fuel in the process. Nitrogen from air and hydrogen from natural gas are combined under high temperature and pressure to produce Ammonia. Ammonia production is an energy intensive process. By making the process more efficient we can reduce the energy consumption of the process and reduce its environmental impact.

Abstract

PakArab fertilizer limited was established on November 15, 1972 by the joint venture of Islamic Republic of Pakistan and state of Abu Dhabi. The aim was to strengthen the ties between the two states, however PakArab was privatized under the privatization policy in July 2005 and it was then acquired by Fatima Group and Arif Habib Group.

PakArab is a natural gas-based steam reforming power plant which was designed by KELLOG INTERNATIONAL CORPORATION. It has a nominal capacity of 910 t/day of liquid Ammonia at -34 C.

Pakarab is in competition with other fertilizer companies like Engro, Fauji fertilizer etc. Due to economic constraints Pakarab plant is shut down currently. Therefore, this project is very critical to the industry. Under the scope of the project we are trying to minimize energy utilization and save cost.

Pakarab faces 4 shutdowns annually due to the Catacarb system installed to remove CO₂ from process gas. This thesis report compares the solvents used to remove CO₂ with respect to efficiency and cost.

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Nomenclature

A_{cp} = Cold plain area

A_T = Total area

A_R = Refractory area

α = effectiveness factor for cold plain area

ε = Gas emissivity

ctc = center to center distance

d_o = outer diameter of tube

f = overall exchange factor

T_g = Flue gas temperature

T_s = Surface temperature of tube

HETP = Height equivalent of theoretical plates

NTU = Number of transfer units

F_{LV} = Flow factor

V_w = Vapor flow rate per unit column cross-sectional area

L_{min} = Minimum solvent flow rate

F_p = Packing factor

K_n = equilibrium constant

ρ_v = density of vapor

ρ_l = density of solvent

Z = Height of packing

Chapter #01

Introduction and Review

(1.1) Ammonia

Ammonia is a colorless gas with a characteristic pungent smell. It is lighter than air, its density being 0.589 times that of air. It is easily liquefied due to the strong hydrogen bonding between molecules; the liquid boils at $-33.3\text{ }^{\circ}\text{C}$ and freezes at $-77.7\text{ }^{\circ}\text{C}$ ($-107.86\text{ }^{\circ}\text{F}$) to white crystals.

Haber's process:

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

The catalyst

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

Temperature and Pressure:

A high temperature of about 450°C .

A high pressure of about 200 atmospheres is required.

Uses:

Fertilizer: Globally, approximately 88% (as of 2014) of ammonia is used as fertilizers either as its salts, solutions or anhydrously. When applied to soil, it helps provide increased yields of crops such as maize and wheat 30% of agricultural nitrogen applied in the USA is in the form of anhydrous ammonia and worldwide 110 million tons are applied each year.

Refrigerant: R717 is used as a refrigerant.

Ammonia is also used to make the following compounds:

Hydrazine, in the Olin Rasching process and the peroxide process

Hydrogen cyanide, in the BMA process and the Andrussov process

Hydroxylamine and ammonium carbonate, in the Raschig process

Phenol, in the Raschig–Hooker process

Urea, in the Bosch–Meiser urea process and in Wöhler synthesis

Fuel composition

POSITION	FEED GAS	
COMPOSITION	MOL%	Lbmol/hr
METHANE	73.00	4270.50
ETHANE	0.30	17.55
CO	0.00	0.00
CO ₂	10.00	585.00
OXYGEN	0.00	0.00
NITROGEN	16.70	976.95
AMMONIA	0.00	0.00
ARGON	0.00	0.00
HYDROGEN	0.00	0.00
TOTAL DRY	-	5850.00
WATER VAPORS	-	0.00
TOTAL WET	-	5850.00
MOLECULAR WEIGHT	20.88	

Table 1: Feed Composition

RAW MATERIALS: Natural Gas and Air

PRODUCTS: Ammonia and Carbon dioxide

(1.2) AMMONIA SYNTHESIS PROCESS DESCRIPTION

(1.2.1) Process description

Ammonia is synthesized by Haber's process. Hydrogen and nitrogen in the ratio of 3:1 react in the presence of iron catalyst bed to produce ammonia. Main raw materials are hydrogen obtained from natural gas by steam-Methane reforming and Nitrogen from air. Oxygen in air is fully consumed by oxidation of combustible gases in the Natural gas.

Following are the process involved in the synthesis of Ammonia

- 1) Feed gas compression
- 2) Desulfurization
- 3) Pre-reformer
- 4) Primary reforming
- 5) Secondary reforming
- 6) Carbon monoxide shift conversion
- 7) Carbon dioxide removal
- 8) Methanation
- 9) Residual CO₂ removed
- 10) Synthesis gas compression
- 11) Synthesis of Ammonia

(1.2.2) Feed Gas Compression

Natural Gas at about 590 psig and 75 F is compressed to 605 psig by a single stage compressor.

(1.2.3) Feed

Natural gas, air and steam are feed to an Ammonia plant. Feed is preheated to 520 F and passed through a desulfurizer. It is then preheated to 840 F and enters the reformer with steam. This mixture is then passed over the catalyst and then reformed to hydrogen and carbon dioxide. Outlet temperature of a reformer is maintained at 1435 F.

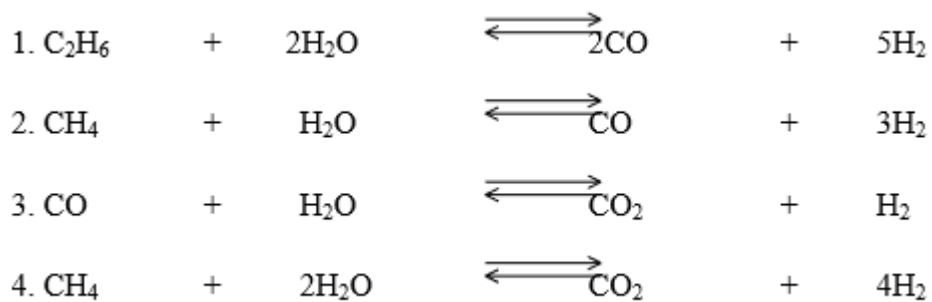
(1.2.4) Feed Gas Desulfurization

Sulfur can seriously affect the performance of the catalyst used in reforming. Sulfur is removed by passing hot feed gas through a bed of NiMo, Zinc oxide and Cu. NiMo converts the organic sulfur contents in the feed gas to inorganic sulfur and then is removed by zinc oxide.

(1.2.5) Pre-Reforming

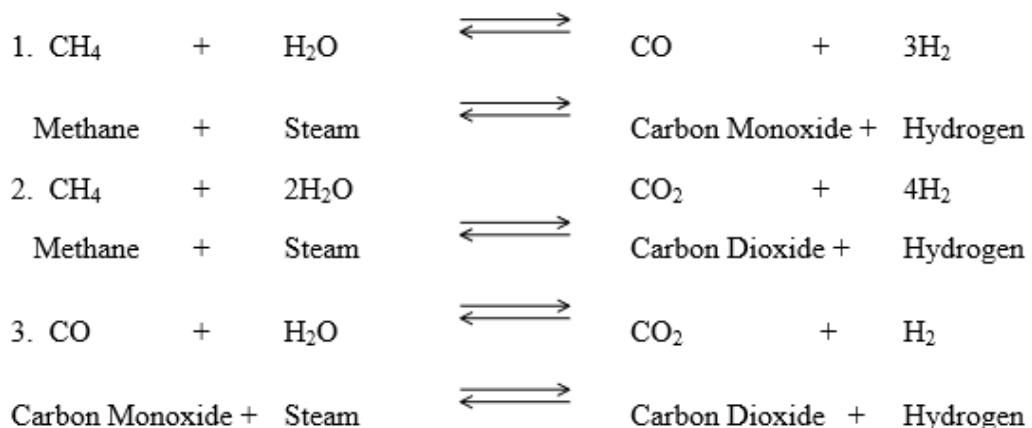
Steam reforming occurs in fixed bed adiabatic reactor that is filled with catalyst.

Following are the set of reactions that occur in a primary reformer



(1.2.6) Primary Reformer

Reforming by definition is the conversion of hydrocarbons to hydrogen. Natural gas is reacted with steam and passed over nickel catalyst. Reaction occurs in tubes of the reformer at 510 psig. Hydrogen carbon monoxide and carbon dioxide come out as products at 1435 F.



(1.2.7) Compressed Air

Ambient air is compressed to 500 psig and then heated in convection coils in the primary reformer. It then enters the secondary reformer, where oxygen is completely reacted with hydrogen.

(1.2.8) Secondary reformer

Air and steam at about 510 psig and 1150 F enter the secondary reformer along with primary reformer effluent. Hydrogen to nitrogen ratio is maintained at 3:1. Oxygen in air is fully consumed by reaction with combustible gases like methane, carbon monoxide and hydrogen. Since reforming reactions are endothermic, combustion reaction provides enough energy to raise the temperatures to the desired level. The products leave the secondary reformer at about 1700 F and methane slip of about 0.55% on dry gas basis.

(1.2.9) Carbon monoxide shift conversion

Carbon monoxide is converted to carbon dioxide by reacting it with steam. it is an exothermic reaction. Although this reaction occurs in reforming section too, however due to high temperatures in the reformer it is retarded and reformer effluent has a high concentration of carbon monoxide.

(1.2.10) High temperature shift conversion

Low temperatures favor shift conversion reaction. Therefore, effluent from secondary reformer is cooled and entered in the HTSC. Iron oxide and chromium catalyst are used in HTSC. Most of carbon monoxide is converted to carbon dioxide and only about 2.1% of CO leaves the HTSC.

(1.2.11) Low temperature shift conversion

This converter charged with ZnO and CuO catalyst practically completes carbon monoxide conversion to dioxide. The CO in the effluent gas stream is less than 0.17 vol% on a dry basis.

(1.2.12) Carbon dioxide removal system

There is approximately 18% carbon dioxide on a dry gas basis. It is compulsory to remove all the CO₂ from the gas. Potassium carbonate solution is used to absorb CO₂ and about 0.1 vol% is left in the effluent of removal section.

Carbon dioxide removal system has an absorber and a stripper with carbonate solution circulating in a closed loop. Process gas enters the absorber from downwards and comes in contact with the carbonate solution coming downwards. Absorber Pressure is kept at 405 psig. The hot absorber solution enters the stripper where it is stream stripped to remove absorbed CO₂. Stripper Operates at about 16 psig.

(1.2.13) Methanation

Methanation is the final step in the purification section. Any carbon monoxide and dioxide left in the process gas is converted to methane. Gas enters the methanator at 550 F and leaves at 590 F. The catalyst used in the methanator is nickel oxide and is supported on alumina.

Gas leaving the methanator contains CO₂ in small amount and water in significant amounts. Both of these are poisonous to the catalyst in the converter and hence must be removed. So, sodium zeolite is used to remove the impurities.

(1.2.14) Ammonia synthesis gas compression

A multistage compressor compresses the purified synthesis gas to 2900 to 2950 psig.

(1.2.15) Ammonia production

Hydrogen and Nitrogen combine to form Ammonia at 380 F in the presence of iron oxide catalysts. Inert gases like argon and methane does not participate in any reaction. However, they retard the synthesis reaction by reducing the concentration of reactants. Inerts are maintained at 5 to 7 % with the help of a purge.

(1.2.16) Condensation process

Synthesis gas containing 17% ammonia is cooled from 745 F and generates 585 psig pressure steam. Further cooled to 35 F, it heats the boiler feed water then converted feed effluent exchanger. It is then chilled in ammonia chillers and about 75% of ammonia is condensed as a result.

(1.2.17) Product

Anhydrous ammonia is supplied to two urea plants, whereas surplus is stored in tanks. The product is 99.75 % ammonia and rest are water. Ammonia once a week is inspected for oil/grease, percent water and percent ammonia content.

(1.3) PROCESS FLOW DIAGRAM

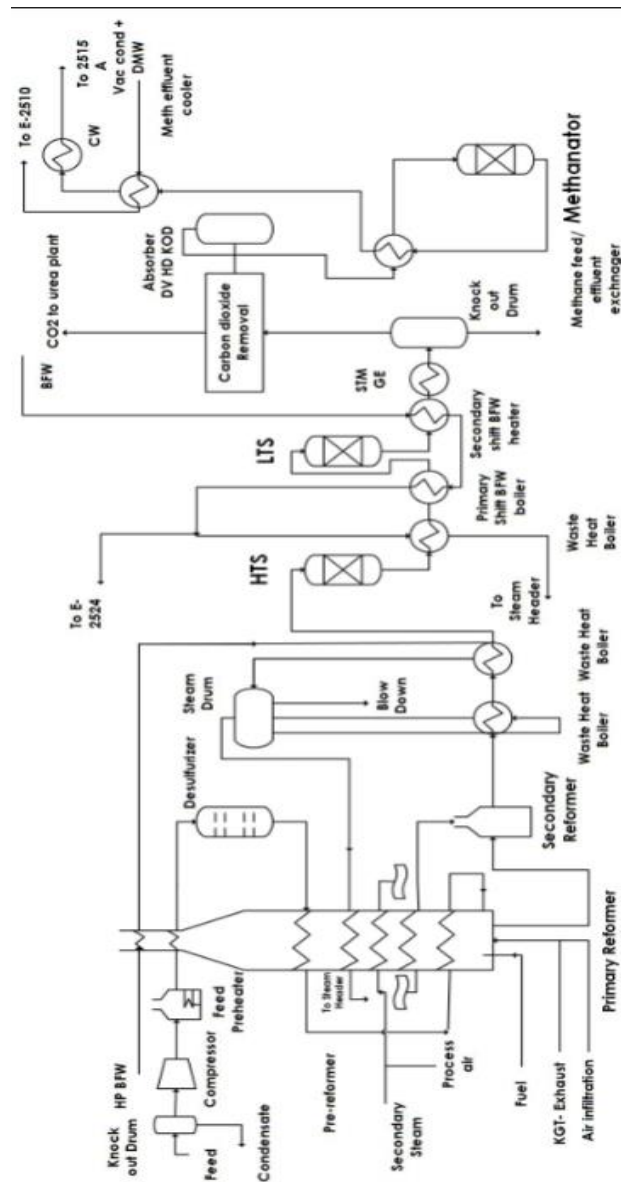


Figure 1 Process Flow Diagram

Chapter # 02

MATERIAL BALANCE

(2.1) Reforming section

POSITION	TOTAL FEED	PRIMARY REFORMER IN	SECONDARY REFORMER IN	SECONDARY REFORMER OUT
Composition	Mol %	Mol %	Mol %	Mol %
Methane	70.87	55.40	9.03	0.080
Ethane	0.29	0	0	0
CO	0	0.16	10.11	12.62
CO ₂	9.7	12.12	11.64	9.19
O ₂	0	0	0	0
N ₂	16.92	14.16	6.42	22.42
Ammonia	0	0	0	0
Argon	0.01	0.01	0	0.02
H ₂	2.10	18.08	62.82	54.93
Molar flow rate dry Lbmol/hr	6027.83	7218.89	15900.71	21617.75
Molecular weight	20.53	18.22	14.48	16.16
Molar flow rate wet Lbmol/hr	6027.83	21291.58	26430.42	33796.07

Table 2 Material Balance, Reforming section

POSITION	AIR IN PRIMARY REFORMER	FLUE GAS	FUEL IN
Composition	Mol %	Mol %	Mol %
Methane	0	0	73.0
Ethane	0	0	0.2
CO	0	0	0
CO ₂	1.80	11.14	9.6
O ₂	16.80	2.5	0
N ₂	80.45	86.09	17.2
Ammonia	0	0	0
Argon	0.95	0.18	0
H ₂	0	0.03	0
Molar flow rate dry Lbmol/hr	19643	24496.23	2841.23
Molecular weight	29.20	27.55	20.84
Molar flow rate wet Lbmol/hr	21575.34	30564.5	2841.23

Table 3 Material Balance, Reforming section

(2.2) Purification section

POSITION	HTSC IN	HTSC OUT	LTSC OUT	KNOCK OUT DRUM IN
Composition	Mol %	Mol %	Mol %	Mol %
Methane	0.80	0.73	0.71	0.71
Ethane	0	0	0	0
CO	12.63	2.53	0.27	0.27
CO ₂	9.19	17.33	19.16	19.15
O ₂	0	0	0	0
N ₂	22.42	20.41	19.96	19.96

Ammonia	0	0	0	0
Argon	0.02	0.02	0.02	0.02
H ₂	54.70	58.97	59.88	59.89
Molar flow rate dry Lbmol/hr	21571.45	2375.48	24288.36	24285.17
Molecular weight	16.16	16.16	16.16	16.97
Molar flow rate wet Lbmol/hr	33749.77	33800.36	33800.70	30748.90

Table 4 Material Balance, Purification section

POSITION	METHANATOR IN	METHANATOR OUT
Composition	Mol %	Mol %
Methane	0.92	1.45
Ethane	0	0
CO	0.35	0
CO ₂	0.20	0
O ₂	0	0
N ₂	24.63	24.48
Ammonia	0	0
Argon	0.23	0.23
H ₂	73.637	70.79
Molar flow rate dry Lbmol/hr	19672.19	19307.78
Molecular weight	9.04	9.14
Molar flow rate wet Lbmol/hr	20174.70	19307.78

Table 5 Material Balance, Purification section

(2.3) Carbon Dioxide Removal Section

POSITION	ABSORBER IN	CLEAN GAS	STRIPPER OUT	STEAM ENTERING STRIPPER
Composition	Mol %	Mol %	Mol %	Mol %
Methane	0.71	0.92	0	0
Ethane	0	0	0	0
CO	0.27	0.35	0	0
CO ₂	19.14	0.20	99.98	0
O ₂	0	0	0	0
N ₂	19.96	24.63	0.02	0
Ammonia	0	0	0	0
Argon	0.02	0.23	0	0
H ₂	59.90	73.67	0	0
Molar flow rate dry Lbmol/hr	24279.59	19672.19	4607	0
Molecular weight	15.52	9.04	25.07	18
Molar flow rate wet Lbmol/hr	25219.60	20174.70	16982	3605184

Table 6 Material Balance, CO₂ removal section

Chapter # 03

Energy Balance

(3.1) Pre-Reformer:

Energy carried by feed: -

Temperature= 515°C

$$Q_{in} = m \int c_p dt = 1.8426 \cdot 10^8 \text{ KJ/h}$$

Heat of Reactions



Total Energy absorbed by reactions = $2.2 \cdot 10^7$ kJ/hr.

Energy Leaving the pre-reformer:

Temperature = 456°C

Total Energy out= Q_{in} - Total Energy absorbed by reactions= $1.6206 \cdot 10^8$ kJ/hr

(3.2) Primary Reformer:

Energy carried by Feed gas:

Temperature = 526.66°C

$$Q_f = m \int c_p dt = 1.9448 \cdot 10^8 \text{ KJ/h.}$$

Energy carried by fuel gas:

Temperature = 23.8°C

$$Q_{\text{fuel}} = 2.22 \cdot 10^5 \text{ KJ/h.}$$

Energy carried by Combustion Air:

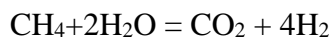
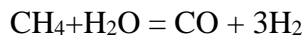
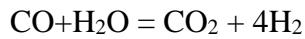
Temperature = 394°C

$$Q_{\text{air}} = 1.485 \cdot 10^8 \text{ KJ/h.}$$

Energy released by combustion of methane and ethane = $8.35 \cdot 10^8$ KJ/h.

$$\text{Stack losses} = Q_{\text{stack}} = m \int c_p dt = 2.21 \cdot 10^8 \text{ KJ/h}$$

Energy absorbed by reforming reactions: -



$$\Delta H_{\text{ref}} = 2.21 \cdot 10^8 \text{ KJ/h.}$$

Energy carried by reformer feed out:

$$Q_{\text{fout}} = 3.4304 \cdot 10^8 \text{ KJ/h.}$$

Energy utilized by side streams = $6.1371 \cdot 10^8$ KJ/h.

$$\text{Total Energy in} = Q_f + Q_{\text{fuel}} + Q_{\text{air}} + \Delta H_{\text{rxn}} = 11.7 \cdot 10^8 \text{ KJ/h.}$$

$$\text{Total energy out} = Q_{\text{fout}} + \Delta H_{\text{ref}} + Q_{\text{stack}} = 11.6 \cdot 10^8 \text{ KJ/h.}$$

(3.3) Secondary reformer:

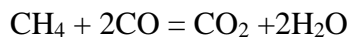
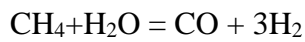
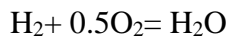
Energy of secondary reformer feed:

$$Q_{\text{feed}} = 3.43 \times 10^8 \text{ KJ/h at temperature} = 785 \text{ }^\circ\text{C}$$

Energy carried by Steam/air mixture:

$$Q = 6.02 \times 10^7 \text{ KJ/h at temperature} = 599 \text{ }^\circ\text{C}$$

Energy evolved by reactions:



$$\Delta H_{\text{rxn}} = 8.48 \times 10^7 \text{ KJ/h.}$$

Energy of outlet stream:

$$Q_{\text{out}} = 4.88 \times 10^8 \text{ KJ/h at temperature} = 919^\circ\text{C}$$

$$\text{Total Energy in} = 4.88 \times 10^8 \text{ KJ/h.}$$

$$\text{Total Energy out} = 4.86 \times 10^8 \text{ KJ/h.}$$

(3.4) HTS converter:

$$\text{Energy in} = \text{Energy feed} + \Delta H_{\text{rxn}}$$

$$= (1.83 \times 10^8) + (3.97 \times 10^7) = 2.45 \times 10^8 \text{ KJ/h.}$$

$$\text{Energy out} = Q_{\text{out}} = 2.45 \times 10^8 \text{ KJ/h.}$$

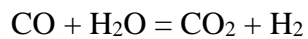
(3.5) LTS converter:

LTS feed in at $T = 212.78^\circ\text{C}$

$$\text{Energy in} = 1.3889 \times 10^8 \text{ KJ/h}$$

Energy evolved:

$$\Delta H_{\text{rxn}} = 9.96 \cdot 10^6 \text{ KJ/h.}$$



$$\text{Total Energy in} = 1.5426 \cdot 10^8 \text{ KJ/h.}$$

$$\text{Total Energy out} = 1.54 \cdot 10^8 \text{ KJ/h.}$$

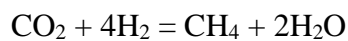
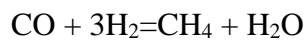
(3.6) Methanator:

Methanator feed in at temperature 273.9°C

$$Q_{\text{in}} = 6.645 \cdot 10^7 \text{ KJ/h.}$$

$$\text{Energy evolved} = \Delta H_{\text{rxn}} = 5.145 \cdot 10^6 \text{ KJ/h}$$

Reactions:



$$\text{Total energy in} = 7.14 \cdot 10^7 \text{ KJ/h.}$$

$$\text{Total energy out} = 7.14 \cdot 10^7 \text{ KJ/h.}$$

(3.5) Overall energy balance:

Table 7 Overall Energy Balance

Energy Carrier's	Energy in	Energy Carrier's	Energy out
	KJ/h		KJ/h
Feed gas	6.9×10^5	Stack losses	2.21×10^8
Compressor	5.22×10^6	Steam	1.37×10^8
Stripper steam	2.93×10^7	Stripper steam	2.93×10^7
Fuel in	2.22×10^5	Waste energy	8.9×10^7
Combustion air	1.585×10^8	CO ₂ recovered	1.86×10^7
Process steam	4.8×10^7	Methanator energy	7.139×10^7
Energy Absorbed(reaction)	8.347×10^8	Reaction energy	8.35×10^8
Secondary steam	2.1×10^8		
Process air	1.1×10^8		
Total	1.4×10^9	Total	1.39×10^9

Chapter # 04

LITERATURE REVIEW

(4.1) FEED PRE-HEATER (103B):

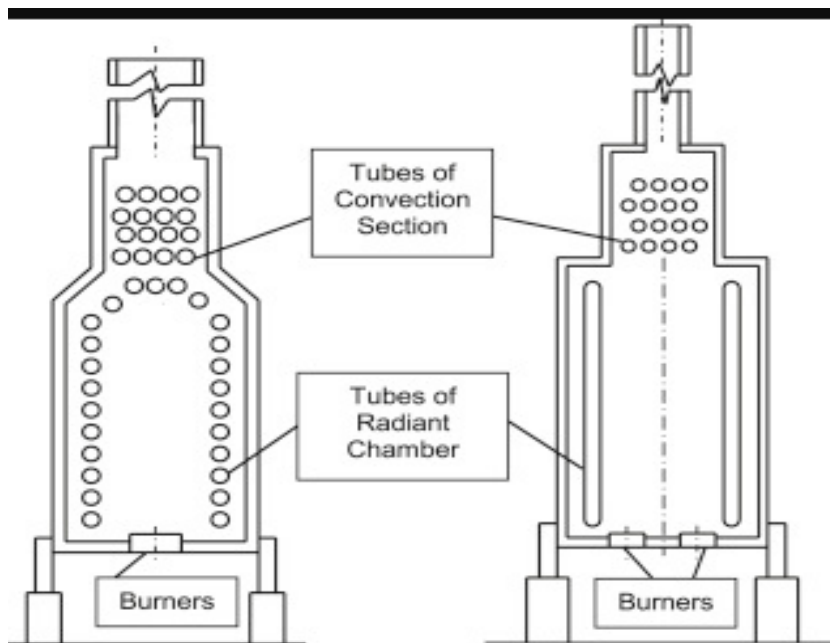


Figure 2 Fired Heater

Furnace mainly has 4 sections

- 1) Shield
- 2) Radiant
- 3) Convection
- 4) Stack

A fired heater maybe a box or vertical in shape. Fired heaters raise the temperature of the feed by combustion of natural Gas or oil. Depending upon the use they are classified as furnaces or processed heaters. Processed heaters use the combustion heat to perform a reaction in the tubes.

(4.2) FEED GAS PREHEATER 103 B DATA SHEET

Table 8 Feed gas pre-heater Data Sheet

HEAT DEMAND	518 Gcal/hr
3% LOSSES ASSUMED	0.16Gcal/hr
HEAT DEMAND THEORETICAL	534Gcal/hr
HEAT DEMAND INCL FLUE GASES	636 Gcal/hr
NATURAL GAS CONSUMPTION	840Nm ³ /hr
HEATER EFFICIENCY	80.2%

CONDITIONS	INLET	OUTLET
TEMPERATURE C	38	399
PRESSURE Kg/cm ²	41.6	40.22
LIQUID FLOW kg/hr	-	-
VAPOUR FLOW kg/hr	20012	20012
LIQ DEG API	-	-
VAP MOLECULAR WEIGHT	16.3	16.3

(4.4) CO₂ REMOVAL SECTION

Process gas leaves the low temperature shift convertor containing 18% carbon dioxide on a dry basis. In order to avoid fouling and denaturing of catalysts in following processes carbon dioxide has to be removed. At present carbon dioxide is being removed using hot potassium carbonate solution. The gas leaving the absorber contains 0.1% Vol CO₂ on a dry basis.

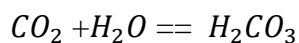
CO₂ removal system consists of an absorber and stripper. Potassium carbonate solution circulates in a closed loop between absorber and stripper. Lean solution enters the absorber from top and counter currently comes in contact with the process gas containing carbon dioxide. It strips off the CO₂ from process gas. Clean gas leaves the absorber from top and goes to Methanation section. Rich solution exits the absorber from the bottom and enters the stripper from top. It now comes in contact with steam counter currently where CO₂ is stripped off from the rich solution. Rich solution becomes lean and leaves the stripper from bottom. It again enters the absorber from top and the circle continues.

CO₂ removal efficiency depends upon following factors

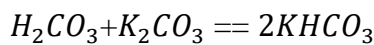
- 1) Concentration of CO₂ in process gas entering absorber
- 2) Volumetric rate of solvent
- 3) Stripping steam quantity
- 4) Heat duty of reboilers

An increase in any of the above factors increases removal efficiency.

(4.4.1) REACTIONS IN ABSORBERS



Carbon dioxide + water == carbonic acid



Carbonic acid + potassium carbonate == potassium bicarbonate

(4.5) ABSORBER

(4.5.1) OPERATING CONDITIONS

Absorber slip	0.1%
Process gas inlet temperature	204 F
Process gas outlet temperature	184 F
Lean solution temperature	185 F
Semi lean temperature	245 F
Rich solution outlet temperature	255 F

Table 9 Absorber operating conditions

(4.6) STRIPPER

Stripper top temperature	234 F
Stripper top pressure	16 psig
CO ₂ outlet temperature	230 F

Table 10 Stripper operating conditions

(4.7) COMPARISON OF BENFIELD AND MDEA AS A SOLVENT

(4.7.1) Benfield Process

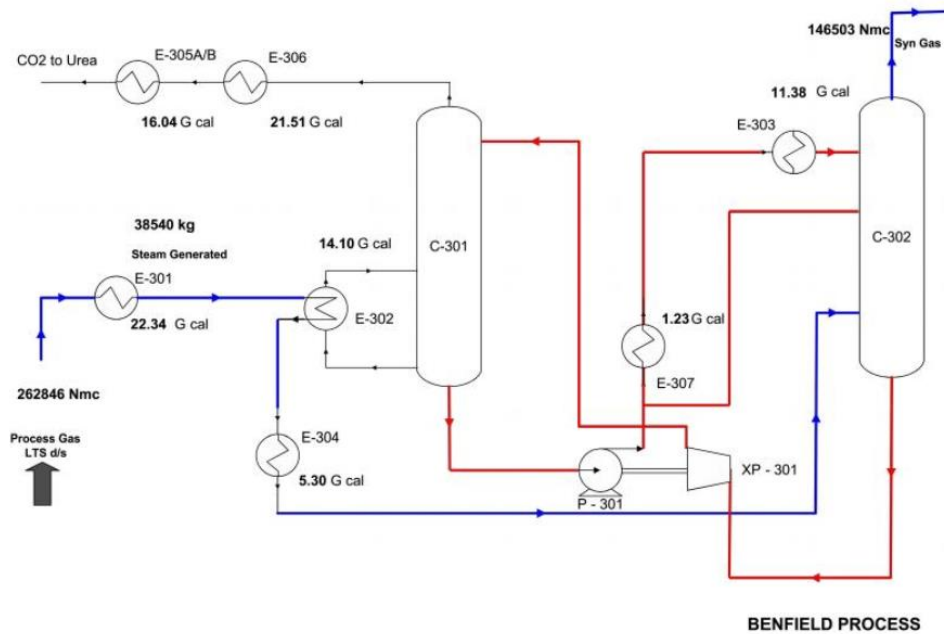


Figure 3 Benfield's process flow diagram

(4.7.2) MDEA Process

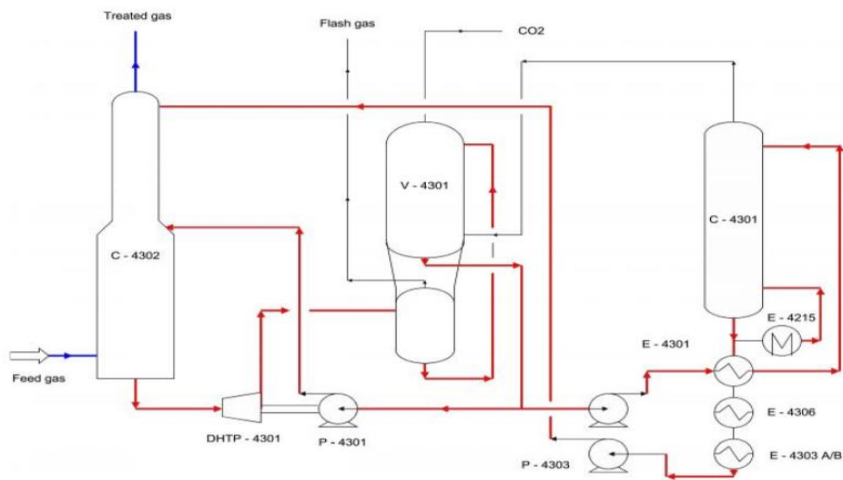


Figure 4 MDEA process flow diagram

Comparison of solvent MDEA and Benfields

	MDEA	Benfield
Circulation rate m ³ /hr	2150	1270
Concentration	40%	28%
Additives	piperazine	DEA, V2O5
Antifoam	required	Required
Steam required t/hr	Nil	44
CO ₂ I/L %	18-19	18-19
CO ₂ O/L ppm	Not detected	500-700
Heat required for regeneration Kcal/m ³	167	669
Solution cost per one charge	29.8	6.98
Activator degradation	Nil	DEA degrades

Table 11 Comparison of solvent MDEA and Benfields

Chapter # 05

PLANT DESIGN

(5.1) Furnace Design

We will be designing vertical type indirect fired heater furnace

We will calculate number of tubes, Area of furnace, tube length by assuming flux.

Assumptions:

Average flux density = 5000 BTU/hr

Known data:

Heater Duty = 3908962 BTU/hr

Overall efficiency = 82 %

Fuel LHV = 23852 BTU/LB

Excess air = 15%

Diameter of tube = 0.718 Ft

Tube length = 35.07 Ft

Tube wall/ surface temperature = 637 F

(5.1.1) Using Loeb and Evan's Method

Method from (Kern, 2002):

Average flux = 5000 BTU/hr

$Q/\dot{a} * A_{cp} * F$

Where

$Q = \text{Net Energy BTU/hr}$

$\dot{\alpha} = \text{relative effectiveness factor of the tube bank}$

$A_{cp} = \text{Cold plane area of the tube bank Ft}^2$

$F = \text{Exchange Factor}$

Assuming $Q / \dot{\alpha} * A_{cp} * F$ is equal to twice of avg Flux

So

$$Q / \dot{\alpha} * A_{cp} * F = 2 * \text{avg flux}$$

$$= 10000 \text{ BTU/hr}$$

Assuming exchange factor as 0.5129

$$Q / \dot{\alpha} * A_{cp} * F = 19496.978$$

So

From the graph

$$Y\text{-axis} = 10,000$$

$$T_s = 600 \text{ F}$$

So, from graph $T_g = 1050 \text{ F}$

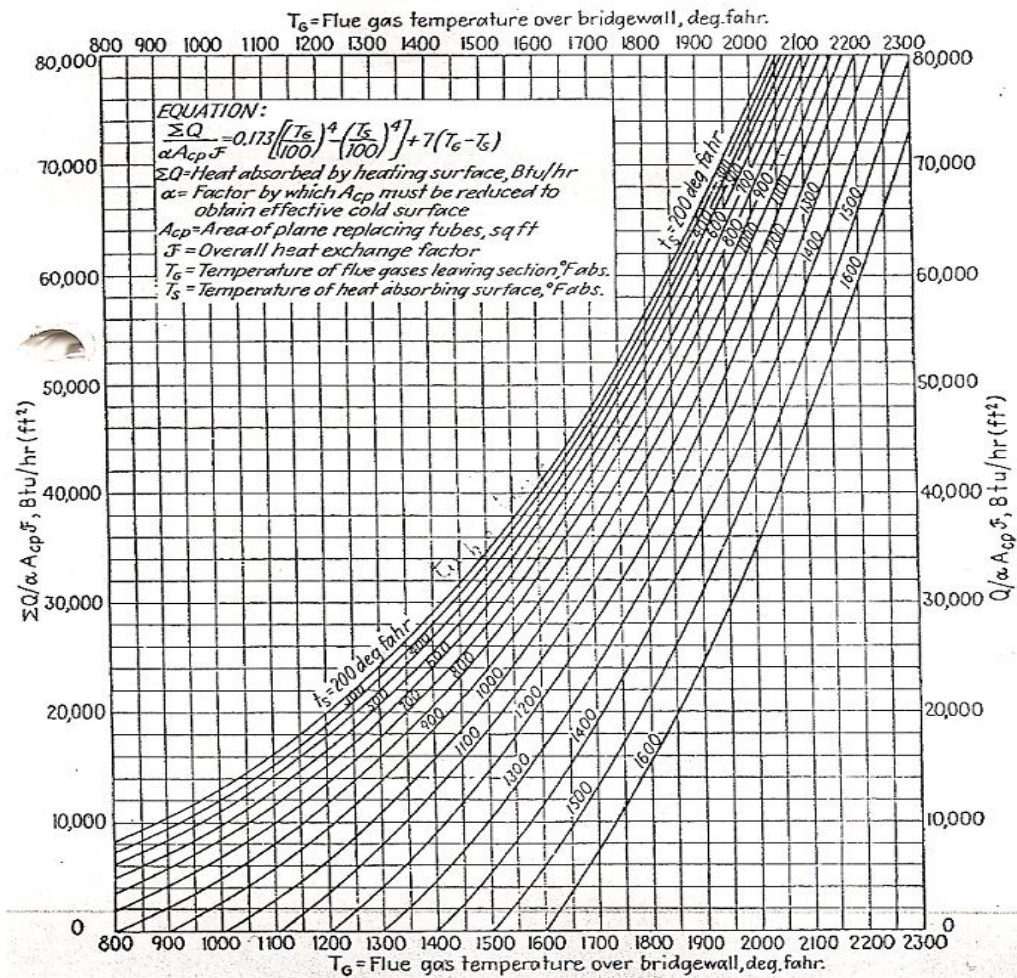


Figure 5.1.1: Loeb and Evan's graph (Kern, 2002)

(5.1.2) Energy balance on furnace

$$Q_f = 21027927.77$$

Value is taken from energy and material balance section

$$\text{Fuel quantity} = Q / \text{LHV}$$

$$= 21027927.77 / 23852$$

$$= 861 \text{ Lb/hr}$$

$$\text{Mass of air} = \text{fuel mass} * 17.77$$

$$= 861 * 17.77$$

$$= 15375.10 \text{ Lb/hr}$$

$$Q_{\text{air}} = m \cdot \Delta H_{\text{at}}$$

$$= 15375 \cdot 82$$

$$= 1260758$$

$$Q_w = 2\% \text{ of } Q_f \text{ (assumed) losses}$$

$$= 0.02 \cdot 21024927.77$$

$$= 420558.555$$

$$\text{Total } Q_{\text{in}}$$

$$= Q_f + Q_{\text{air}} - Q_w$$

$$= 21027927.77 + 1260758 - 420558.555$$

$$= 21868128.02$$

$$Q_{\text{flue gasses}} = 9578146$$

$$Q_{\text{net}} - Q_{\text{fluegasses}}$$

$$= 12289981.99 \text{ BTU/hr}$$

$$N_{\text{tubes}} = Q / (2 \cdot r \cdot L \cdot q \cdot 3.14)$$

$$N_{\text{tubes}} = (12289981.99) / (2 \cdot 3.142 \cdot 35.07 \cdot 5000)$$

$$\text{Area} = 2 \cdot \pi \cdot r \cdot L$$

$$N_{\text{tubes}} = 35.11$$

Centre to center distance between tubes

$$c_{tc} = 16 \text{ in}$$

$$A_{cp} = c_{tc} \cdot L$$

$$= 16/12 \cdot 35.07$$

$$= 46.76 \text{ ft}^2$$

ctc / OD

$$= 16/12 * 0.718$$

$$= 0.95733$$

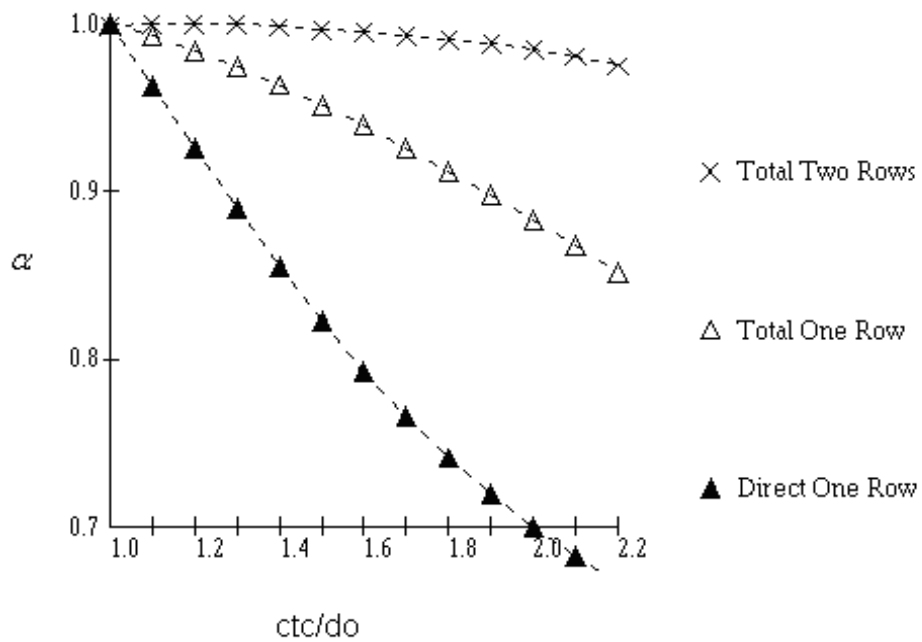


Figure 5 Effectiveness factor for A_{cp} , Graph (Kern, 2002)

So \dot{a} comes out to be from graph

$$\dot{a} = 0.8$$

$$\dot{a} * A_{cp} / \text{tube}$$

$$= 0.8 * 46.76$$

$$= 37.408$$

$$\dot{a} * A_{cp}$$

$$= 1313.55$$

Assumed $A_t = 2103 \text{ ft}^2$

$$Ar = At - \dot{a} * Acp$$

$$= 2103 - 1313$$

$$= 789.44$$

$$Ar / \dot{a} * Acp = 0.601$$

Mean beam length = 13.32 ft

Gas Emissivity

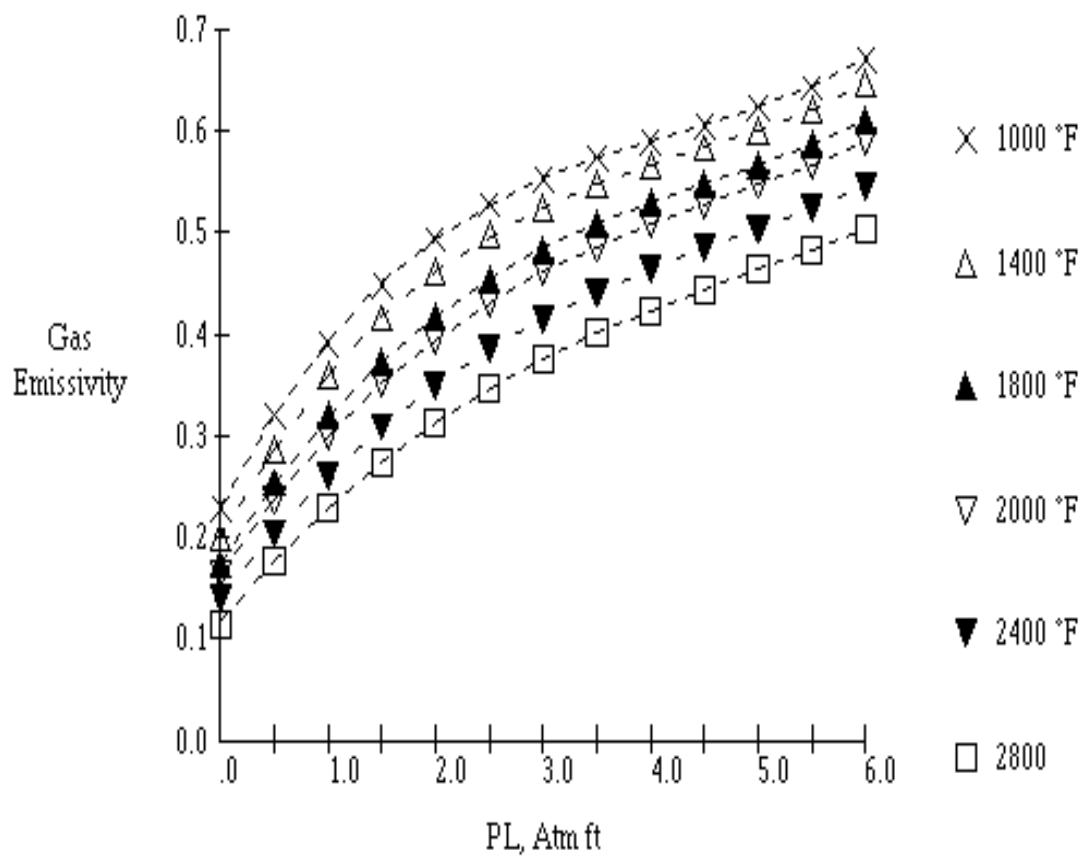


Figure 6 Gas emissivity graph (Kern, 2002)

$P = 0.256 \text{ atm}$

$L = 13.32 \text{ m}$

$PL = 3.40992 \text{ matm}$

Gas emissivity = 0.55

$Ar/\alpha A_{cp} = 0.6$

$E_g = 0.5087$

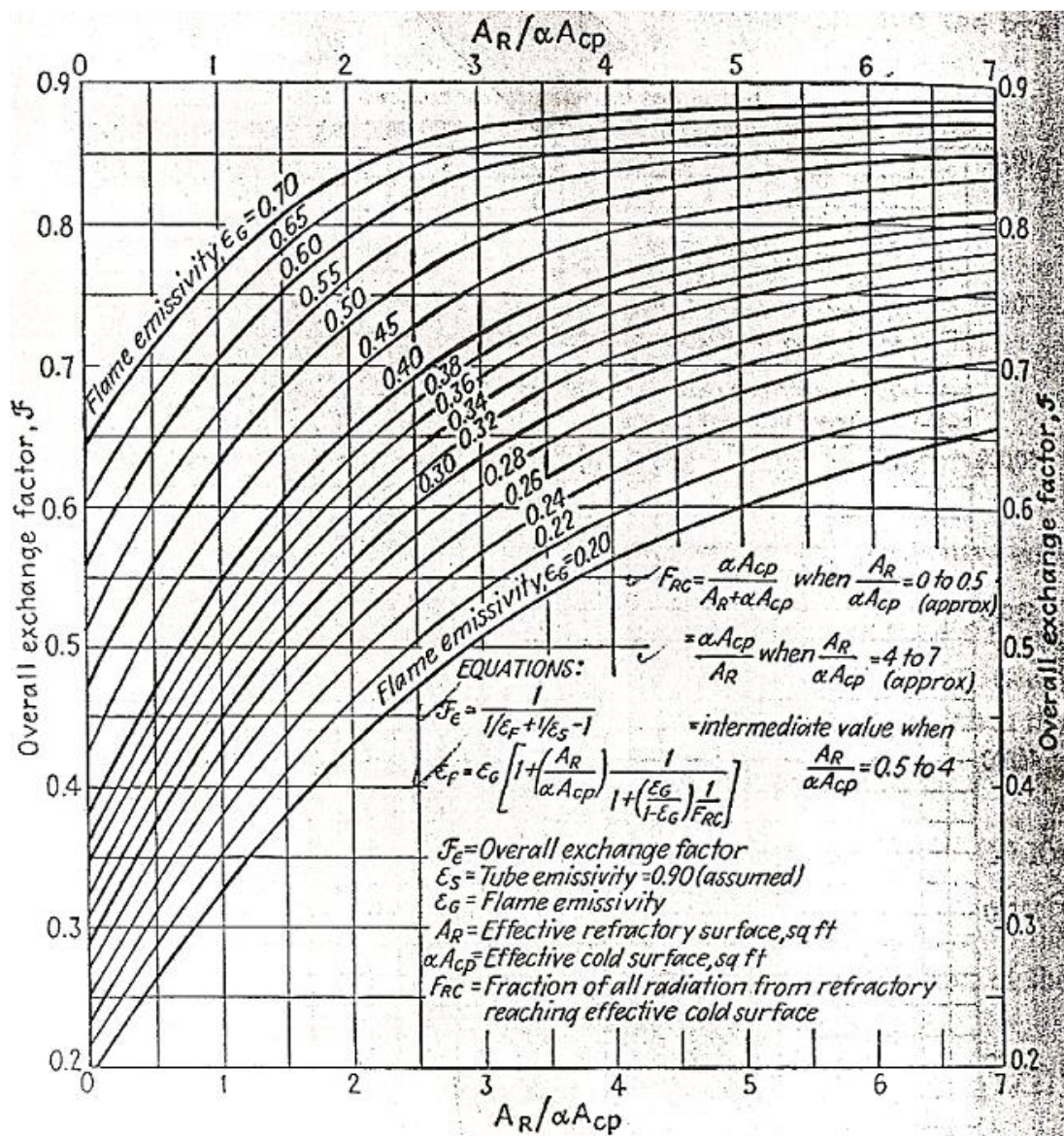


Figure 7: Overall exchange factor graph (Kern, 2002)

So, from graph find the value of F (overall exchange factor)

$$F = 0.5$$

Now calculate

$$Q / (\dot{a} * A_{cp} * F)$$

$$= 18712.575$$

Calculate % error in assumed and calculated value

$$19496.978 - 18712.575 / 19496.978$$

$$= 4\% \text{ error}$$

(5.1.3) ALTERNATIVE CALCULATION FOR EMISSIVITY

Gas emissivity was calculated using Kern's Method

$$p_{CO_2} = 0.1084$$

$$p_{H_2O} = 0.1248$$

$$L = 15$$

$$E_g = 0.489 \sim 0.5 \text{ from graph}$$

(5.2) Design of absorber

Data known:

Vapor gas flow rate = 11439 kgmol/hr

Pressure = 2.75 MPa

Temperature = 385 K

CO₂ vapor mol fraction IN = 0.19

CO₂ vapor mol fraction out = 0.001

Ln (H) = a + b/T + c Ln(T) + dT (Zhang)

a = 19.899

b = -1072.2

c = 0

d = 0

Ln H = 17.11

H = 27.07 MPa

Roult's law says:

p' = y_{n+1}*P

p' = 524 kPa

Henry's law says

p' = x_n*H

x_n = 0.019355 fraction of CO₂ absorbed in liquid

Finding equilibrium constant

K_n = Y_{n+1} *(x_{n+1})/x_n*(y_{n+1} + 1) (Richardson, 2005)

K_n = 9.8163425

$$L_{\min} = V * kn * \text{fraction of solute absorbed}$$

$$L_{\min} = 111698 \text{ kgmol/hr}$$

$$L = L_{\min} * 1.2$$

$$L = 134038 \text{ Kgmol/hr}$$

$$N_{OG} = \ln [(1 - (m * G_m / L_m) * y_1 / y_2) / [1 - (m * G_m / L_m)]] \text{ (Richardson, 2005)}$$

$m = kn$ equilibrium constant

$$m \text{ (Gm/Lm)} = 0.83774$$

$$(1 - m \text{ (Gm/Lm)}) = 0.16226$$

$$N_{og} = 21.2949 \text{ m}$$

$$N_{\text{tubes}} = \mathbf{23.2356}$$

Packing type and its physical properties were selected from design data for various packings. Given in Coulson 6th edition. (Richardson, 2005)

$$\text{HETP taken from literature} = 0.75$$

$$\text{Hence } Z = NTU * \text{HETP}$$

$$= 23.2356 * 0.75$$

$$Z = 17.4267 \text{ m}$$

Diameter calculations

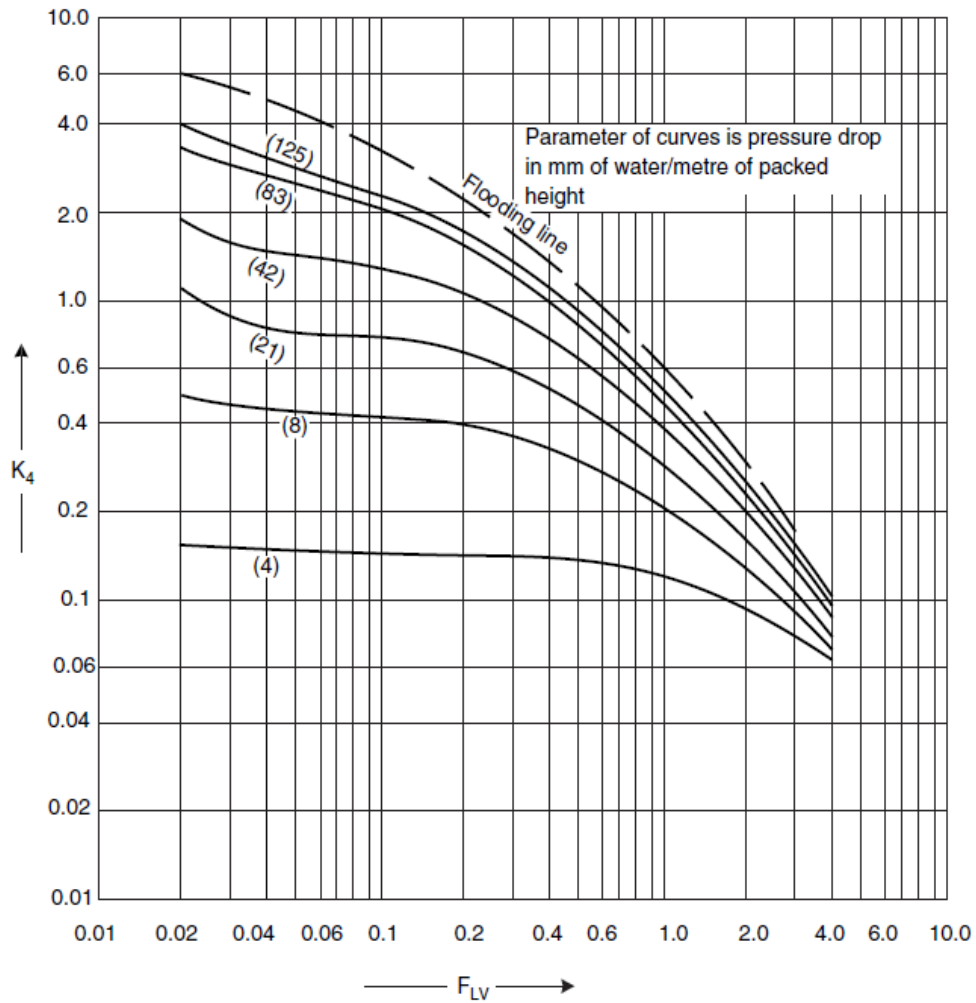


Figure 5.2.2: Flow factor & pressure drop correlation (Perry, 1997)

$$F_{LV} = L_w^* / V_w^* \sqrt{(\rho_v / \rho_L)} \quad (\text{Richardson, 2005})$$

$$\rho_v = 137702 \text{ kg/m}^3$$

$$\rho_L = 13372$$

$$F_{lv} = 3.0429$$

From graph K4 at flooding

$$K_4 = 0.2$$

$$p_v^*(p_l - p_v)$$

$$= 1662540760$$

$$F_p = 66 / \text{m}$$

$$U_1 = 0.000627$$

$$13.1 * F_p * (u_l / p_l)^{0.1} = 126.66$$

At flooding

$$V_w = 1620.23186 \text{ kg/m}^2\text{sec}$$

$$\text{At } 60 \% \text{ flooding } V_w^* = 972.139$$

$$\text{Area} = \text{mass flow rate} / \text{mass flux}$$

$$= 137702 / 972.139$$

$$= 11.766 \text{ m}$$

$$D = \text{under root } (4 * A / \text{pie})$$

$$\mathbf{D = 3.8704 \text{ m}}$$

(5.3) Design of stripper

Data known:

Liquid flow rate = 8700 kgmol/hr

Pressure = 1.1 kPa

Temperature = 320 K

$\ln H = a + b/T + c \ln(T) + dT$ (Zhang)

$a = 19.899$

$b = -1072.2$

$c = 0$

$d = 0$

$\ln H = 16.5$

$H = 15376870 \text{ Pa}$

Henry's law

$p' = x_n * H$

$p' = 297626 \text{ Pa}$

Roult's law says

$p' = y_{n+1} * P$

$y_{n+1} = 0.2698$

Finding equilibrium constant

$$K_n = Y_{n+1} * (x_{n+1}) / x_n * (y_{n+1} + 1) \text{ (Richardson, 2005)}$$

$$V_{\min} = 7768.66 \text{ kgmol/hr}$$

$$N_{OG} = \ln [(1 - (m * G_m / L_m) * y_1 / y_2) / [1 - (m * G_m / L_m)]] \text{ (Richardson, 2005)}$$

M = kn equilibrium constant

$$m \text{ (Gm/Lm)} = 0.808$$

$$N_{og} = 12.6 \text{ m}$$

$$N_t = 14.04$$

From coulsan and richardson volume 6

50 mm packing size was selected .

HETP taken from book = 1

Hence $Z = NTU * \text{HETP}$

$$= 15.59 * 1$$

$$\text{Packing height} = 15.59\text{m}$$

Diameter calculations:

From (Perry, 1997) and (Richardson, 2005)

$$F_{LV} = L_{w*} / V_{w*} \sqrt{(\rho_v / \rho_L)}$$

$$pV = 7462 \text{ kg/m}^3$$

$$pL = 137702$$

$$F_{LV} = 3.3429$$

From graph K4 at flooding = 0.2

$$p_v*(p_l-p_v) = 971850880$$

$$F_p = 66 / \text{m}$$

$$U_1 = 0.000627$$

$$13.1 * F_p * (u_1/p_l)^{0.1} = 126.66$$

At flooding:

$$V_w = 1238.23186 \text{ kg/m}^2\text{sec}$$

At 60 % flooding

$$V_{w*} = 743.139$$

Area = mass flow rate / mass flux

$$= 1238 / 743.139$$

$$= 10.45 \text{ m}$$

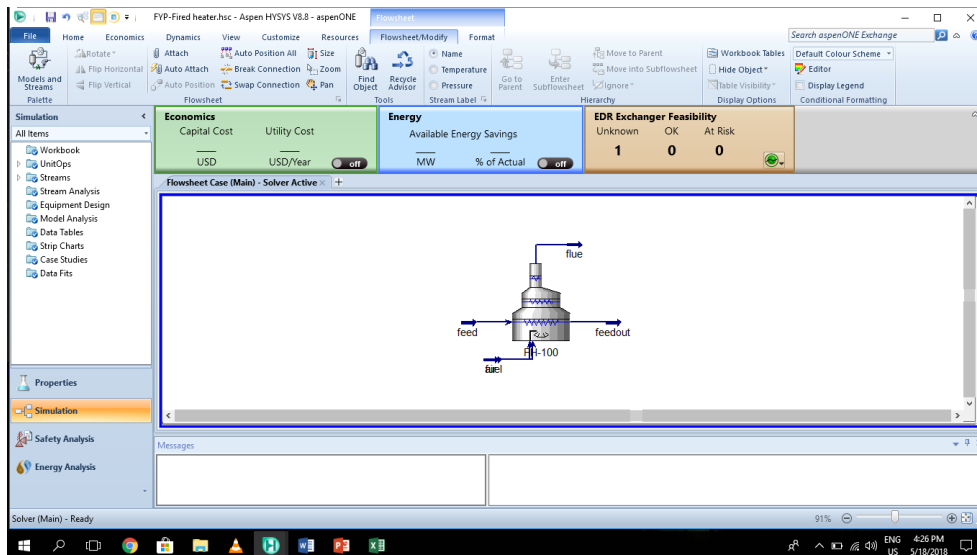
$$D = \sqrt{(4 * A / \pi)}$$

$$\mathbf{D = 3.6704 \text{ m}}$$

Chapter #06

SIMULATION

(6.1) Fired heater:



Fired Heater: FH-100

Design	Rating	Worksheet	Performance	Dynamics	EDR FiredHeater		
Worksheet		Name	feed	air	fuel	feedout	flue
Conditions	Vapour	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
Properties	Temperature [C]	47.7778	25.0000	23.8889	271.1111	276.7373	
Composition	Pressure [kPa]	101.3	101.3	101.3	101.3	101.3	
PF Specs	Molar Flow [kgmole/h]	2653.5426	373.5436	46.3872	2653.5426	420.0040	
	Mass Flow [kg/h]	55407.7417	10776.8335	968.5947	55407.7417	11745.4282	
	LiqVol Flow [m3/h]	134.0150	12.4580	2.3427	134.0150	14.1066	
	Molar Enthalpy [kJ/kgmole]	-9.352e+004	-8.181	-9.437e+004	-8.447e+004	-6.761e+004	
	Molar Entropy [kJ/kgmole-C]	185.6	151.7	182.8	206.7	180.8	
	Heat Flow [kJ/h]	-2.4815e+08	-3.0560e+03	-4.3774e+06	-2.2413e+08	-2.8398e+07	

Fired Heater: FH-100

Design	Rating	Worksheet	Performance	Dynamics	EDR FiredHeater		
Performance							
<div style="border: 1px solid gray; padding: 10px;"> <p>Overall Performance</p> <table border="1" style="width: 100%;"> <tr> <td style="width: 50%;">Duty [kJ/h]</td> <td style="width: 50%; text-align: right;">2.402e+007</td> </tr> </table> </div>						Duty [kJ/h]	2.402e+007
Duty [kJ/h]	2.402e+007						
<ul style="list-style-type: none"> Details Plots Tables Setup 							

(6.2) Absorber

Column: T-100 / COL1 Fluid Pkg: Basis-1 / Acid Gas

Design Parameters Side Ops Rating Worksheet Performance Flowsheet Reactions Dynamics

Performance

Summary
Column Profiles
Feeds / Products
Plots
Cond./Reboiler
Acid Gas

Feeds

	solvent	feed gas
Flow Rate (MMSCFD)	183.7000	25.2000
Methane	0.0000	0.0071
MDEAmine	0.4000	0.0000
CO	0.0000	0.0027
CO2	0.0000	0.1914
Nitrogen	0.0000	0.1996
Hydrogen	0.0000	0.5990
H2O	0.6000	0.0002
H2S	0.0000	0.0000

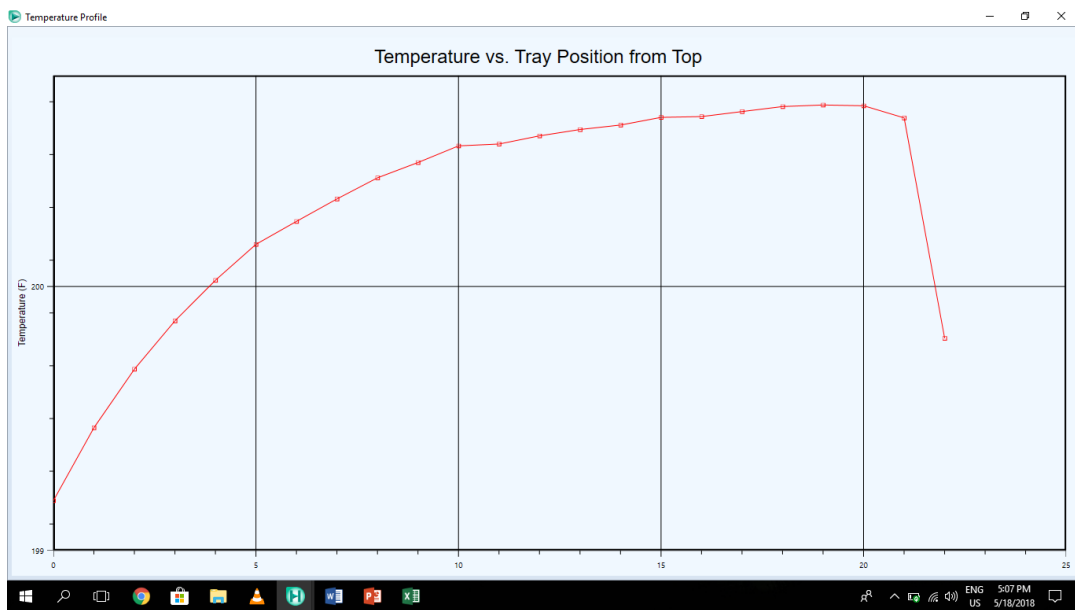
Products

	clean gas	r solvent
Flow Rate (MMSCFD)	183.7440	212.4000
Methane	0.0077	0.0000
MDEAmine	0.0000	0.3950
CO	0.0030	0.0000
CO2	0.0961	0.0141
Nitrogen	0.2192	0.0001
Hydrogen	0.6566	0.0003
H2O	0.0174	0.5904
H2S	0.0000	0.0000

Composition
 Flows
 Recovery
 Molar
 Mass
 Liq Vol

Delete Column Environment... Run Reset Converged Update Outlets Ignored

Windows Taskbar: ENG 5:08 PM US 5/18/2018



Column: T-100 / COL1 Fluid Pkg: Basis-1 / Acid Gas

Design Parameters Side Ops Rating Worksheet Performance Flowsheet Reactions Dynamics

Rating

Tower Sizing

Towers	Tray/Packed Section	Main Tower
Vessels	Uniform Section	<input checked="" type="checkbox"/>
Equipment	Internal Type	Valve
Pressure Drop	Diameter [ft]	4.921
	Tray/Packed Space [ft]	2.000
	Tray/Packed Volume [ft ³]	31.20
	Disable Heat Loss Calcs	<input type="checkbox"/>
	Heat Model	None
	Rating Calculations	<input type="checkbox"/>
	Hold Up [ft ³]	3.120
	Weeping Factor	1.000
	Tray Sizing Analysis for Costing	<empty>

For more detailed and tray by tray information see the individual tower in the Column/Environment.

Delete Column Environment... Run Reset

Converged Update Outlets Ignored

Windows taskbar: 5:05 PM 5/18/2018

Chapter # 07

HAZOP ANALYSIS

The term “HAZOP Analysis” stands for hazard and operability analysis. It is systematic and structured method of examining an already operating system or a system which is planned for operation in future. It is a qualitative analysis of the system which is done by a team of qualified experts to find the locations in system which have high risk potential towards equipment and labour. It is also used to avert the possible adverse effects to process efficiency, which may occur due to the risks at the certain locations. The study is qualitative and also identifies deviations which may potentially result in process inefficiencies. So, this is operability study not to be confused with “Hazard Analysis” which quantitatively measures the risk associated with the process.

(7.1) Procedure:

Certain words are used in identification of steps used for carrying out the hazard and operability analysis (HAZOP). These words along with their brief definitions are given below:

Intention:

This defines how a particular part of the process was intended to operate i.e. the intention of the designer.

Deviations:

These are departures from the designer’s intention. These deviations are detected by the systematic application of the guide words.

Causes:

Here possible reasons for the deviation are considered answers to the questions like how and why are determined. Only the deviation resulting from a realistic cause is treated as meaningful.

Consequences:

The results of a meaningful deviation (now meaningful can be large deviations for some processes and small for processing requiring critical control).

Hazards: Consequences that can potentially cause damage (loss) or injury.

(7.2) Case 1:

An adiabatic pre-reformer is a fixed bed reactor where gases (natural gas + steam) are introduced from top and leave the reactor from bottom. Now looking at various parameters which can deviate various cases are considered.

Intention: Keeping the temperature at pre-reformer inlet at about 525°C.

Deviation: The temperature rises to about 600°C.

Causes: Since the feed to pre-reformer is being heated to its introduction in the pre-reformer. Thus, this depends on the temperature of the flue gases and the heat exchange in coil as a lower heat transfer there (due to whatever reason) would mean higher temperature of the flue gases (in contact with coil) thus leading to higher feed temperature. Also, an increase in primary reformer firing will result in higher flue gases temperature, resulting in higher methane conversion in pre-reformer.

Hazard: This can potentially eliminate methane slip from secondary reformer resulting in oxygen carryover into shift converters poisoning the catalyst there.

(7.3) Case 2:

Intention: Maintaining the pre-reformer conversion at 6%.

Deviation: The conversion increases.

Causes: The increase in conversion can be due to the increase in temperature in the pre-reformer section or pressure loss due to the fouling in pre-reformer feed pre-heating coils.

Consequences: The increase in conversion will disturb the intended methane slip from the pre-reformer.

Hazards: This will cause oxygen carryover to the shift converters causing the poisoning of shift converters catalyst.

(7.4) Case 3:

The feed pre-heater is a fired heater. It is a furnace to heat the feed before it fed to the reformer section.

Intention: To maintain the temperature according to the design/operation requirements.

Deviation: The temperature increases than the design/ operation requirements.

Causes: The increase in the temperature can be due to the increase in the firing and the increase in the burning of fuel.

Consequences: The increase in temperature results in coking and decrease in efficiency of furnace. This will increase the temperature of the feed that will be fed to the reactor.

Hazards: The increased temperature can be threatening to the working environment.

Recommended measures: Controlled burning can control the temperature. It maintains the temperature.

(7.5) Case 4:

Intention: Oxygen content of the fuel gas in the combustion chamber should be maintained.

Deviation: Oxygen content of the fuel gas in the combustion chamber is less than the requirement.

Causes: Premature combustion results in the decrease in the oxygen content. Excessive burning of the fuel is one of the reasons too.

Consequences: When serious, it may lead to heating furnace extinction, flammable gas accumulation into the heating furnace and explosion.

Hazards: It can reduce the combustion in the chamber. It results in the accumulation of the flammable gas in the chamber which results in the explosion.

Recommended measures:

Confirm whether combustion air interruption interlock is set up in the heating furnace to cut off the fuel gas supply.

(7.6) Case 5:

Intention: Pressure of the fuel gas should be maintained.

Deviation: Pressure of the fuel gas is too low.

Consequences: It may lead to flameout, if the fuel gas pressure changes, it leads to the risk of secondary explosion.

Recommended measures: Add high pressure alarm. Confirm whether combustion chamber high pressure interlock is set up in the heating furnace to cut off the fuel gas supply.

Chapter # 08

COSTING

(8.1) Fired Heater Costing

Costing of Furnace (Richardson, 2005)

Absorbed duty = 12289981.99

Bare cost = \$143406.0523;

$F_d = 1$

Furnace cost = \$143406.0523

In 1987 cost is \$143406.0523

Index factor = 2.9

In 2017 = 1987* index factor

Cost of Fired Heater = \$415877.5518

CONCLUSION:

In order to enhance the efficiency of ammonia plant, we first identified key areas to work on. Carbon dioxide removal section was selected on the basis that it gets shutdown 4times annually. Benfield system uses potassium carbonate solution as a solvent for carbon dioxide removal. Potassium carbonate is corrosive and a very dirty solution. It requires large steam quantities for regeneration. Hence a study on solvents supported MDEA, Methyldiethanolamine as the most suitable solvent.

Research suggests that MDEA is non-corrosive and requires very less steam for regeneration. It does not degrade the absorber column as opposed to Benfield solution. To support our study, we designed a carbon dioxide removal section with MDEA as a solvent. Calculations and cost analysis confirm our findings and we were able to make an annual profit of 0.7million Rupees.

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