Incorporation of Adiabatic Pre-reformer at FFC Ammonia-II

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Certificate

This is to certify that work in this thesis has been carried out by **Mr. Muhammad Aqib, Hafiz Muhammad Abdullah and Mirza Muhammad Jamal** and completed under my supervision at school of chemical and materials engineering, National University of Sciences and Technology, H-12, Islamabad, Pakistan.

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Dedication

We would like to dedicate this piece of work to our teachers and family.

Acknowledgement

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

We are grateful to Dr. Arshad Chughtai for his sincere efforts, advice and guidance throughout the project.

We are thankful to our FFC supervisor Mr Farooq (Deputy Manager Engineering) and Ms Anum Shafique, who despite of their tough routine helped us, provided the necessary data and responded promptly to our emails.

We are also thankful to all of our teachers who have helped us and taught us valuable lessons which would go a long way in defining our futures.

ABSTRACT

In this project a solution to a problem at FFC-Ammonia-2 plant has been proposed. The plant is currently being operated at 20% overcapacity, due to which firing in primary reformer has been increased and this has led to excessively high flue gases temperature in the convection section (where heat of flue gases is used to heat various streams) of the primary reformer. This high temperature is harmful for heat exchanging coils and reduces their operational life and poses risks of equipment failure. At the moment extra steam is being injected into the secondary reformer air/steam stream. This steam absorbs the excess heat. This is a makeshift arrangement and our objective was to provide a viable alternative. After literature review, consideration of current industrial practices (especially in ammonia production facilities) and consultation with our both supervisors (industrial and university), incorporation of adiabatic pre-reformer has been proposed as it increases the plant capacity by 20-25% without additional energy cost. This modification has been made at various plants around the world and thus is a reliable technique. The proposed pre-reformer's operating conditions have been specified. The volume of catalyst required has been calculated. Along with heat extraction scheme, a control loop to maintain the process at desired conditions has also been designed. After incorporating the proposed modification it has been concluded that protection steam would no longer be required. Apart from that feed stock flexibility and primary reformer operational flexibility would also be provided. Also the catalyst life of the primary reformer catalyst would be enhanced.

Table of Contents

Nomenclature

k	m^3/kg cat.s	*overall reaction rate constant*
r_i	m^3/kg cat.s	*reaction rate for reaction i, where i can be 1, 2 or 3*
ki	m^3/kg cat.s.$kPa^{0.25}$	*reaction rate constant for reaction i, where i can be 1, 2 or 3*
Pi	kPa	*partial pressure of i, where i can be CH_4, CO_2, CO, H_2 or H_2O*
kpi	kPa^2	*equilibrium constant of reaction i, where i can be 1, 2 or 3*
KCO	kPa^{-1}	*CO absorption coefficient*
KH_2	$kPa^{-2.75}$	*H_2 absorption coefficient*
KH_2O	-	*H_2O absorption coefficient*
P	kPa or bar	*pressure*
T	K or °C	*temperature*
x	-	*conversion*
W	kg	*weight of the catalyst*
F	kgmol/hr	*feed to pre-reformer molar flow rate*
$[CH_4]$	kmol/m^3	*concentration of methane*
$[CH_4']$	kmol/m^3	*initial concentration of methane*
R	kPa.m^3/mol.K	*universal gas constant*
V	m^3	*volume*
ρ	kg/m^3	*density*
H	m	*height*
v_s	m/s	*superficial velocity*
ε	-	*bed voidage*
D_p	m	*surface equivalent particle diameter*
R_{ep}	-	*Reynolds number*
μ	Pa.s	*viscosity*
Sa	m^2	*surface area of particle*
f_p	-	*bed friction factor*

ΔP_c	kPa or bar	*pressure drop due to catalyst bed*
V_p	m^3	*volume of particle*
ΔP_{b1}	kPa or bar	*pressure drop due to inert alumina balls on top*
ΔP_{b2}	kPa or bar	*pressure drop due to inert alumina balls at the bottom*
ΔP_t	kPa or bar	*total pressure drop of catalyst bed and alumina balls*
T_{in}	K or °C	*temperature of feed stream at reactor inlet*
T_{out}	K or °C	*temperature of product stream at reactor outlet*
$-\Delta H_{R1}$	kJ/kmole	*heat of reaction for reaction 1*
$-\Delta H_{R2}$	kJ/kmole	*heat of reaction for reaction 2*
p_m	kmol/m^3	*molar density*
ΔT	K or °C	*temperature difference*
c_p	kJ/kmol °C	*heat capacity*
D_i	m	*internal diameter of shell*
S	bar	*maximum allowable stress at design temperature property of material of construction*
E	-	*weld efficiency*
P_d	bar	*design pressure of vessel*
t_a	m	*shell thickness*
t_b	m	*hemispherical head thickness top and bottom*
m'	kmol/hr	*molar flow rate*
C_c	kJ/hr.K	*product of molar flow rate and heat capacity for cold stream*
C_h	kJ/hr.K	*product of molar flow rate and heat capacity for hot stream*
ε	-	*heat transfer effectiveness*
NTU	-	*number of transfer units*
U	W/m^2.K	*overall heat transfer coefficient*
A_s	m^2	*required surface area*
L_{eff}	m	*effective length of each tube*
d_o	m	*outer diameter of tube*
d_i	m	*inner diameter of tube*
V''	m^3/s	*volumetric flow rate*

A_{st}	m^2	*surface area of each tube*
N	-	*number of tubes*
J_f	-	*friction factor*
N_p	-	*number of passes*

Motivation

Pakistan's economy is still largely based on agricultural sector and fertilizers are essential for good yield of crops. Demand of fertilizers has soared in recent years as more and more farmers are educated about its benefits. Of many products being used one is Urea.

Urea provides nitrogen a primary nutrient which:

- Promotes leaf and stem growth.
- Is an essential component of chlorophyll.
- Regulates the uptake of other key nutrients.

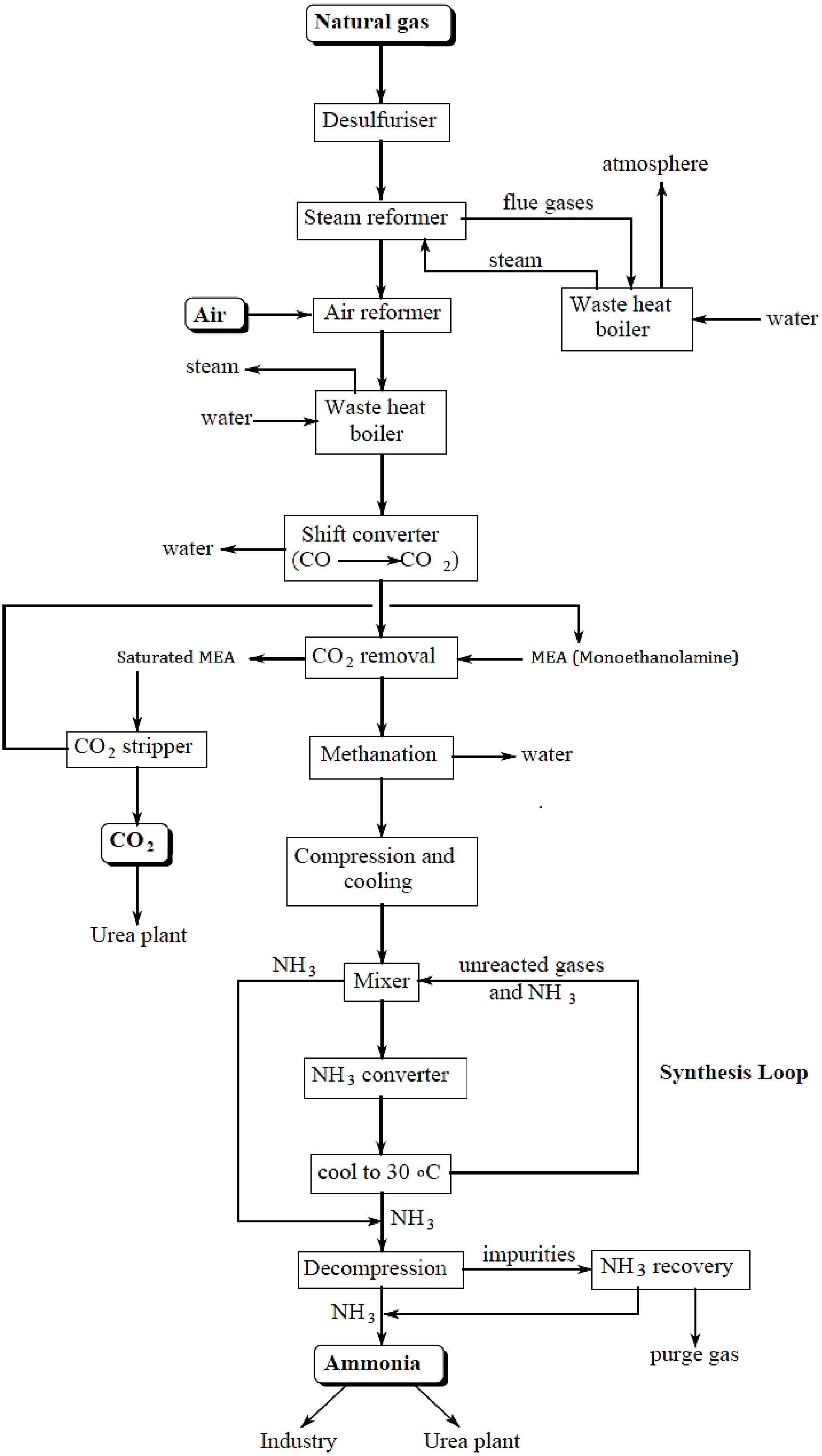
Natural gas is a key raw material used in production of hydrogen which is then combined with nitrogen to produce ammonia. This ammonia is then subsequently combined with carbon di oxide to give urea. Since Pakistan is facing an energy crisis and due to this shortage any modification resulting in reduction of energy consumption or increase in energy efficiency would be beneficial for not only the industry but also the general public as well. Natural gas is the source of energy for fertilizer plants and it is the main feed stock as well.

The production of natural gas currently stands at about 4 MMMSCF a day and the reserves (proven and tapped) are to be exhausted by 2026 as well so it's about time our industry collaborating with government looks to provide solutions and leave no stone unturned in improving process efficiency and this is a little step but surely on in the right direction. Since universities are meant for providing solutions to the problems being faced by the industry and society, but unfortunately in our educational system there is no such collaboration. But this time around at SCME NUST we were assigned industrial projects in which we had a problem and were required to provide a feasible solution to it, this exposed us to the industrial practices and provided us with an excellent opportunity to interact with the industry via our industrial supervisors.

Chapter 1

Ammonia Synthesis

Process Description



Schematic of Ammonia Synthesis

Process Description

Ammonia is produced by Haber's process where hydrogen and nitrogen react over iron catalyst bed to produce ammonia. The process needs air and natural gas as raw materials. Natural gas is obtained from gas fields and is sent directly to ammonia production plant.

Desulfurization

This is the first step in processing of natural gas. Sulphur content of natural gas is reduced to less than 280 μg/m^3 to prevent catalyst poisoning of subsequent reactors. The sulphurous compounds are first converted into hydrogen sulphide (H_2S) by the following reaction.

$$H_2 + RSH \rightarrow RH + H_2S$$

Once converted into inorganic form, the hydrogen sulphide gas (H_2S) is passed over zinc oxide bed at 400 ^{o}C where it is absorbed and removed in the form of solid zinc sulphide.

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

After desulphurization, process steam is mixed in this stream the steam to carbon ratio is 2-3 this will be further discussed later.

Steam Reforming

The desulphurized gas steam stream is heated before being fed into the reformer. Where methane the primary constituent of natural gas reacts with steam on Nickel catalyst to produce a mixture comprising mainly of carbon monoxide, hydrogen and carbon dioxide in minute quantities. This mixture is known as synthesis or syn-gas.

Conventionally a primary reformer followed by secondary reformer are used for this purpose. But advancements in technology and ever increasing emphasis on energy efficiency has led to the development of pre-reformer. Reforming occurs when

Pre-Reforming

In this step all the higher hydrocarbons are converted into methane if the feed stock is naphtha or contains higher concentrations of ethane. In case of predominantly methane feed stock around five to seven percent (depending on the process conditions) methane is reformed.

This step is used to improve energy efficiency of the plant and has the potential to increase the plant throughput by 20-25%. Feedstock mixed with natural gas is introduced at about 450-550°C where reaction takes place over nickel based on alumina catalyst to give product at 400-480°C. The gaseous mixture is heated again before being fed into the primary reformer.

Primary Reformer

This is the reactor where most of the methane conversion takes place (about 70-72 %). This reformer has numerous tubes filled with catalyst (Ni/Al_2O_3). The tubes are fired as, being endothermic reaction's equilibrium is shifted towards right at higher temperature. Following reactions take place in primary reformer.

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

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

$$CO + H_2O \leftrightarrow H_2 + CO_2$$

The feed is introduced at around 550-600°C and leaves the reactor at 800-900°C. The tubes must be able to withstand high temperatures and must have reasonable thermal conductivity so that energy can be transferred effectively.

Tubes of primary reformer not only face very high temperatures but also pressure in excess of 35 bar is experienced which can lead to failure by rupturing if the material of construction is not chosen carefully.

Secondary Reformer

Here the process gas is mixed with compressed pre-heated air at 540 °C. The air is added in such quantity to give a final nitrogen to hydrogen ratio of 1:3 i.e. 3 moles of hydrogen per mole of nitrogen.

The top part of the reformer is the combustion zone and this is followed by a catalyst bed where further conversion of methane takes place. There must be some methane slip from the secondary reformer about 1% so as to confirm that there is no oxygen in the gas stream entering the shift converters, as oxygen is poisonous to the shift converters catalyst.

Shift Conversion

The next step is the shift conversion where the process gases are fed after passing through waste heat recovery section, where the streams temperature is reduced to 350-400 ^oC and fed to high temperature shift converter (HTS). Here carbon monoxide is converted to carbon dioxide by the reaction given below:

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

HTS is filled with chromium oxide initiator and iron oxide catalyst. The gas from the HTS is cooled to 200-250 ^oC before being fed into the low temperature shift converter LTS where carbon monoxide is converted further.

At some instances the gas is first passed through a zinc oxide bed to remove any residual trace amounts of sulphur that would deactivate low temperature shift catalyst. While on other plants low temperature shift converters catalyst is fed in excess to ensure optimum operation of the unit. Final shift gas from this low temperature shift converter at 210^{o}C is cooled to 110°C. Since steam fed is more than the required amount thus there is unreacted steam in the gas stream which is condensed and separated from the gas in a knockout drum (KOD).

Carbon dioxide Removal

The final shift gas contains carbon dioxide, nitrogen and hydrogen. Carbon dioxide is to be separated from this mixture before being further processed this can be done by either of the two methods:

1. Pressure Swing Adsorption (PSA)
2. MEA Absorption.

Pressure Swing Adsorption (PSA)

PSA is the gas separation technology where under pressure and according to the species' affinity and molecular characteristics an adsorbent material, separates a gas from a mixture of gases.

Pressure swing adsorption works due to the fact that gases under high pressure tend to adsorb on solid surfaces. When this pressure is released the adsorbed gases are released. Depending on the differing tendency of the gasses for the media they get adsorbed to different extents and this difference of affinity is used to separate gases.

This technique gives around 99.99% pure Hydrogen in double stage PSA configuration.

MEA Absorption

More than 80% of the ammonia plants use this method to remove CO_2. 15- 30 % solution of MEA in water mixed fortified with corrosion inhibitors. In this method there is an absorption tower in which the gas to be stripped is fed from bottom and lean MEA solution is introduced from top as it falls down the rising gas is stripped of CO_2 as it gets absorbed in the monoethanolamine.

After absorbing the CO_2, the amine-CO_2 solution is preheated and regenerated in a reactivating tower. The reactivating tower removes CO_2 by steam stripping and then by heating. The CO_2 gas (98.5 percent CO_2) is either vented to the atmosphere or used for chemical feedstock in other parts of the plant complex. The regenerated MEA is pumped back to the absorber tower after being cooled in a heat exchanger and solution cooler.

Methanation

Carbon dioxide removal is not hundred percent effective by any process (at least not by the widely employed process) and any trace amount of it in the stream can poison the catalyst of synthesis converter. Catalytic methanation is used to remove this residual carbon dioxide.

Here on a nickel catalyst at 600-800^{o}C and 30 bar pressure methanation takes place following the reactions given below.

$$CO_2 + H_2 \leftrightarrow\ H_2O + CO$$

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

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

The gas at methanator exit contains almost pure one to three ratio of nitrogen to hydrogen, required for the ammonia synthesis.

Ammonia Synthesis:

In this step the gas mixture is firstly compressed to pressures in range of 150-340bar. This is then cooled to 0^oC and mixed with recycle gas. This results in formation of ammonia, ammonia is then separated by condensing it and in liquid-vapour separator and this is then sent to a let-down separator for further separation. Unconverted syn-gas after separation is compressed and heated to 180°C. This compressed and heated syn-gas is then fed to a synthesis converter where, over iron oxide catalyst, conversion to ammonia takes place.

The exit gas is condensed and separated before being sent to a let-down separator where some part of overhead gas is purged. This prevents the buildup of inerts like argon.

Storage:

Ammonia at ambient pressure liquefies at -33^oC. Thus is stored at such conditions in large (20,000 ton) tanks. In most modern facilities these tanks are double walled to minimize chances of leakage and provide insulation.

Chapter 2

Problem at

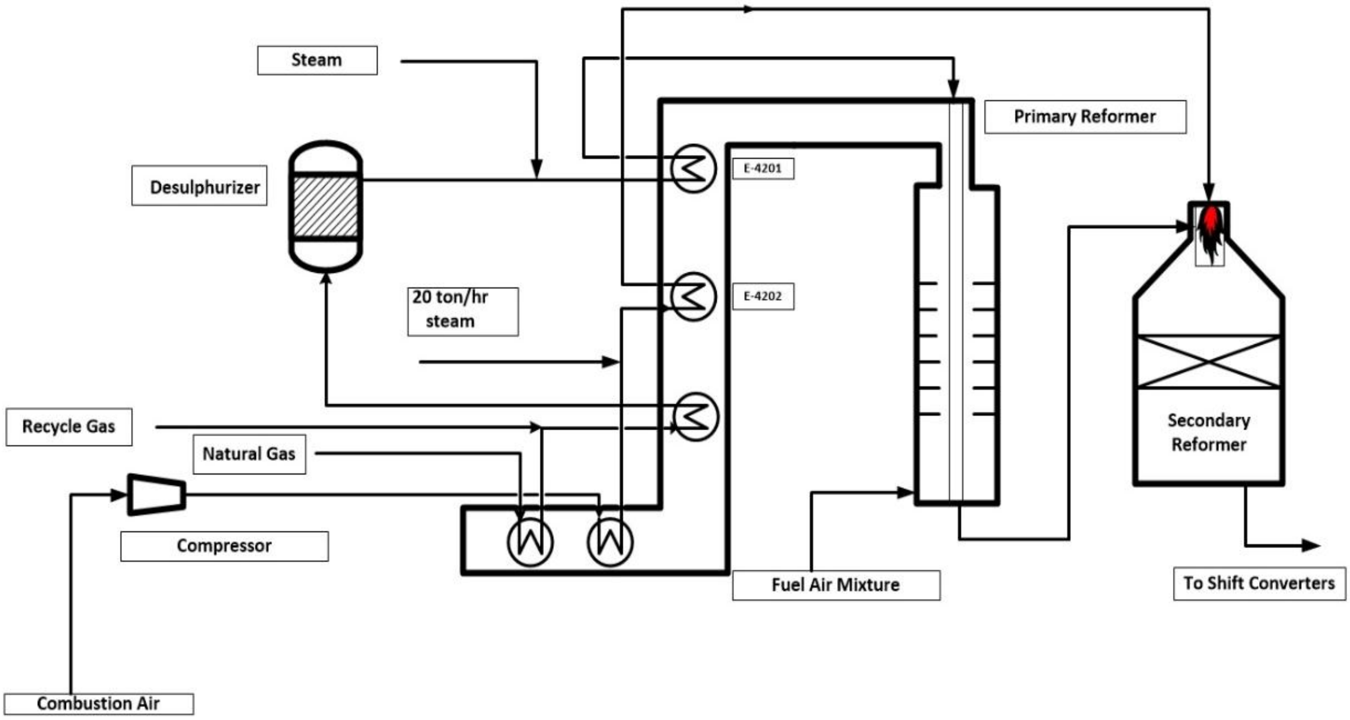
FFC

Ammonia-II

Problem at FFC Ammonia-II

Since the plant is operating at 120% i.e. 20% over capacity. To accommodate for additional throughput, firing in primary reformer has been increased. Due to this increase in firing excessively high (around 1135°C) flue gases temperature downstream in the convection section have been recorded. This exposure of high temperature leads to reduction in safe operational life of the subsequent heat exchanging coils, as the pressures in the exchanger tubes is in excess of 35 bars and this high temperature increases the stresses experienced by coils to very high levels.

The design flue gases temperature is around 1030°C so the current operating temperatures observed are more than 100°C. In order to reduce this temperature to safe limits at the moment 20 tons/hr steam is added in air/steam stream to secondary reformer but this is just a temporary arrangement.



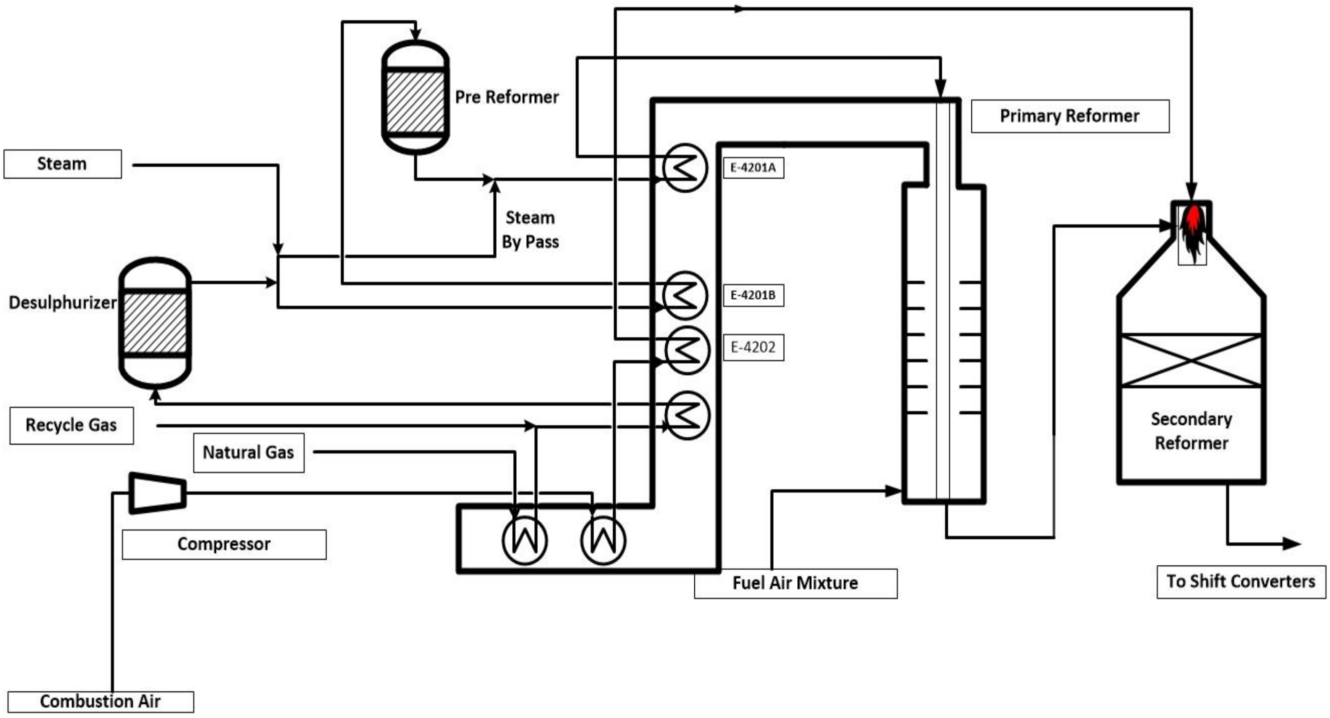
Chapter 3

Proposed Solution & Literature Review

Proposed Solution

The solution proposed to this problem is incorporation of adiabatic pre-reformer. Pre-reformers have been successfully incorporated in already established ammonia plants in India, Netherlands, France and many other countries. In Pakistan Engro fertilizer plant in Daharki, Sindh has also successfully incorporated pre-reformer to increase capacity.

In the modified scheme as shown below, the steam being mixed with the process gas has been split into two one by passing the pre-reformer the reason for this would be explained later. The other steam stream is mixed with natural gas (which has already passed through the desulphurizer), this mixed stream having steam to carbon ratio of 2.4 i.e. S/C = 2.4, at 388^{o}C and 38 bara then passes through coil E-4201B where it is heated to 525^{o}C (the temperature chosen for pre-reformer referring to figure 1). Another consideration regarding pre-reforming temperature is that only so much heat can be extracted from the flue gases so as to reduce its temperature from 1133^{o}C above coil E-4201A to 1026^{o}C this is close to the design temperature of 1030^{o}C, below this the subsequent coils heat transfer operation is disturbed.

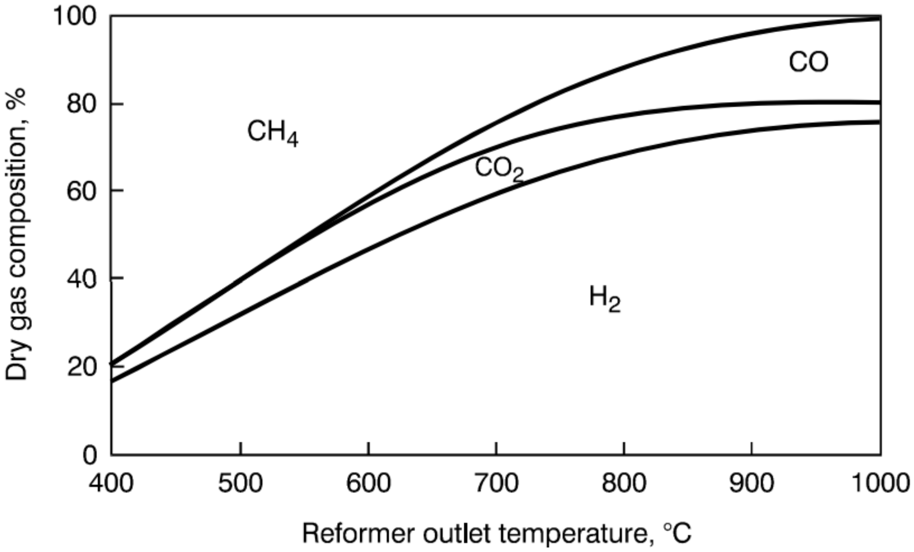


This stream is then fed into the pre-reformer where 6% conversion of methane takes place. Since the reaction is endothermic and the reactor is adiabatic (i.e. no heat in or out of the reactor) the temperature of the products is reduced to 465ºC. This product stream is then passed through coil E-4201A where its temperature is increased from 465ºC to 600ºC. The flue gases temperature at the beginning of convection section is 1133ºC and after exchanging heat with primary reformer feed it is reduced to 1026ºC (well within safe limits), this temperature is further reduced to 920ºC, after heating the pre-reformer feed to 525ºC in coil E-4201B. In this way excess heat is used to pre-heat the pre-reformer feed and then re-heat pre-reformer product as primary reformer feed thus eliminating the need for steam injection into secondary reformer air/steam stream. This will reduce energy consumption to produce steam and save millions in annual energy savings.

Adiabatic Pre-reformer (Literature Review)

Adiabatic pre-reformer is simply a fixed bed reactor, filed with Ni-based on alumina catalyst. If the feed stock is naphtha or higher feed stocks then all the higher hydrocarbons are converted into methane. In case of natural gas feed stock, steam and natural gas mixture is fed at 480-525 ºC, where 5-8% methane conversion takes place in pre-reformer.

Pre-reformer is optimized according to the primary reformer and secondary reformer. Steam to carbon ratio, reforming temperature and conversion are related to each other as follows.



Following reactions occur at equilibrium and are driven towards right side at these operating conditions of 525 ºC and 38 bar pressure to produce a mixture of methane hydrogen carbon dioxide and carbon monoxide.

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

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

$$CO + H_2O \leftrightarrow H_2 + CO_2$$

Steam reforming is highly endothermic reaction and the temperature at reactor outlet is lower than at the inlet as this is an adiabatic process. (Velu, Ke, & Chunshan, 2010)

Factors affecting equilibrium.

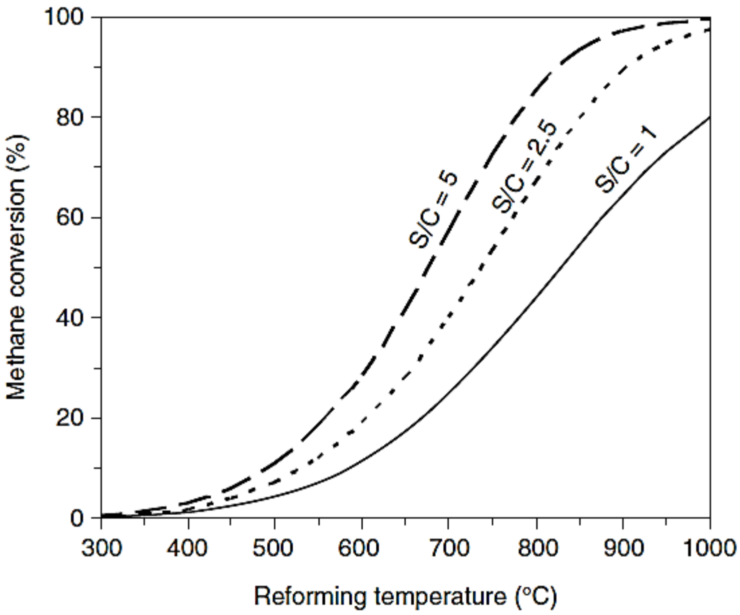
Pressure

Since the reaction produces more moles of product than reactants which means reaction will move in forward direction if the pressure is low i.e. the methane at exit is proportional to the square of pressure i.e. higher the pressure lesser the conversion or more methane at the exit but due to overall economics the operation is carried out at pressures in excess of 30 bar.

Steam to Carbon Ratio

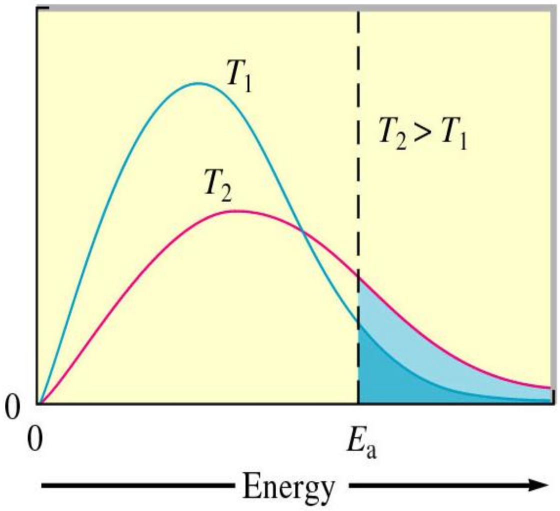
Figure. 1

Although stoichiometry suggests 1:1 ratio of steam and methane but practically S/C ratio is between 2 and 3. At higher temperatures the whisker carbon formation is likely which leads to catalyst deactivation and to prevent this excess steam is added. This ratio is chosen carefully as the cost of steam generation is very high and more steam fed than required would be uneconomical as well.



Temperature

Temperature is another parameter affecting the kinetics and equilibrium. Higher the temperature higher the conversion. According to collision theory this is due to the fact that reaction takes place as a result of effective collisions taking place between the reacting species molecules. Now effective collision is a collision taking place at energy that is higher than the activation energy of the reaction.

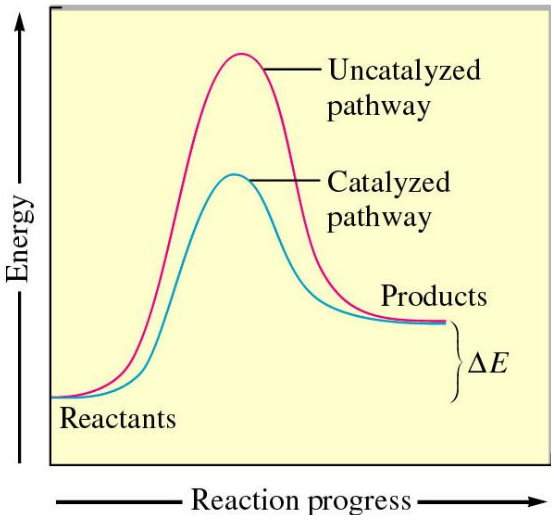


Here in the figure it is shown that at a higher temperature i.e. T_2 the area under the curve shows more effective collisions as compared to lower temperature i.e. T_1. Now this temperature cannot be very high as cost of heating and material properties need to be considered as well.

Catalyst:

A catalyst is a substance which increases the rate of reaction. The catalyst may or may not participate in the reaction and regenerated at the end this essentially means that the catalyst’s mass before and after the reaction is same and is chemically unchanged.

The physical properties of the catalyst may change during the process i.e. from a solid (granular) form to a powder. Catalyst increases the reaction rate by lowering the activation energy of the reaction, with this lower activation energy both the forward and reverse reactions are accelerated and this usually also alters the equilibrium point. As the figure besides shows that the catalysed pathway requires considerably less energy as compared to uncatalyzed pathway. Due to this a large number of molecules collide with enough energy to cross the energy barrier known as “activation energy” thus increasing the number of effective collisions and speeding up the reaction.



While selecting the catalyst following things must be considered:

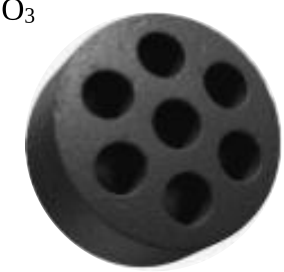
- Catalyst must maintain reasonably high activity for longer time periods and have longer life time.
- Since the operation is carried out at high temperatures it must resist carbon formation which leads to catalyst deactivation if kept unchecked.
- Catalyst deactivation by sintering is also an important parameter to look out for before selection of catalyst.
- High to very high tensile strength value is also desirable to prevent catalyst disintegration during operation which may lead to higher pressure drop.
- The catalyst shape should be such that it causes least pressure drop.

For steam reforming nickel based on alumina (Ni/Al_2O_3) is used although other catalysts have been developed and are used in industries but (Ni/Al_2O_3) is most widely used and has desirable properties.

In this case nickel catalyst on alumina- magnesium base is considered. The catalyst would be provided by Haldor Topsoe it has been specifically designed for adiabatic pre reformer and it has seven holes, which ensure maximum surface area yet cause relatively small pressure drops. Unique preparation method ensures effective activity over wide range of temperature as well as steam to carbon ratio as well.

Catalyst specifications:

Haldor Topsoe pre reforming catalyst AR- 401 Ni based on α-Al_2O_3



Diameter of pellet = 11mm

Diameter of smaller holes = 2mm

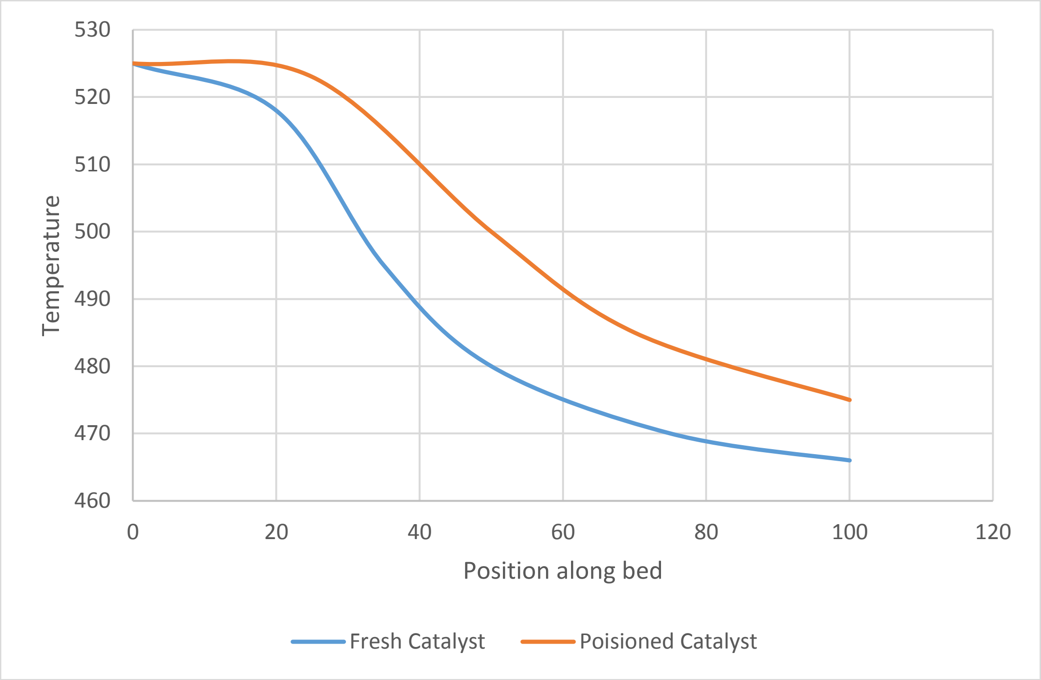
Height of the pallet = 6mm

Life of the catalyst generally is 2-4 years.

Modes of catalyst deactivation

Poisoning:

It refers to the reduction in activity arising due to strong chemisorption of impurities in the feed on the active sites of the catalyst. A poison may reduce the activity by simply blocking the active site or may alter the absorptivity of other species by electronic effect. Poisons may alter the chemical nature of the catalyst or restructure the catalyst surface by reacting and forming another compound. In case of steam reforming H_2S can potentially poison the catalyst. Due to this the desulphurizer unit's efficiency is critical.



Sintering:

At high temperatures the smaller metal crystallites coalesce and agglomerate to form larger particles which have lower surface area to volume ratios. This phenomenon is referred to as sintering and is physical in nature and is a particularly important mode of loss of active surface by structural modification of the catalyst.

Coking:

In catalytic processes involving hydrocarbons (or even carbon oxides), certain side reactions forming carbonaceous residues occur as well. These carbonaceous residues referred to as coke may cover the catalyst surface or block the pores restricting the chemical species' diffusion into pores thus effectively reducing the active surface area of the catalyst.

Chapter 4

Material Balance

Material Balance

Natural Gas from Desulphurizer

Component	Percentage (%)	kgmoles/hr
CH_4	71.03	1537.79
H_2	1.54	33.34
N_2	16.54	358.10
CO	2.15	46.10
CO_2	8.74	187.38
Total	**100**	**2165**

Steam required for S/C ratio of 2.4

Molar Flow Rate of Steam	3690.72

Process Gas (Feed to Pre-reformer)

Component	Percentage (%)	kgmoles/hr
CH_4	26.27	1537.79
H_2	0.57	33.34
N_2	6.12	358.10
CO	0.78	46.10
CO_2	3.20	187.38
H_2O	63.05	3690.72
Total	**100**	**5853.44**

Applying balance on individual species:

Methane conversion is 6% and the major reactions are as given below:

$$CH_4 + H_2O \leftrightarrow 3H_2 + CO \qquad 1$$

$$CO + H_2O \leftrightarrow H_2 + CO_2 \qquad 2$$

Methane in the feed is = 1537.79 kgmoles/hr

Conversion of methane is = 0.06 i.e. 6%

As,

$$Conversion = \frac{Moles\ in\ feed - Moles\ in\ product}{Moles\ in\ feed}$$

Or,

$$Moles\ in\ product = Moles\ in\ feed - Moles\ in\ feed \times Conversion$$

Thus,

Methane in product stream = 1537.79 – 1537.79 × 0.06 kgmoles/hr

Methane in product stream = 1444.78 kgmoles/hr.

Methane converted = 92.27 kgmoles/hr.

From balanced equation 1:

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

1 kgmole of methane requires 1 kgmole of water

Thus, Water consumed = 92.27 kgmoles/hr.

For each mole of methane converted 3 moles of hydrogen are produced.

So,

For 92.27 kgmoles/hr conversion of methane, 276.80 kgmoles/hr of hydrogen are produced.

For each kgmole of methane converted one kgmole of carbon monoxide is produced.

So,

For 92.27 kgmoles/hr conversion of methane, 92.27 kgmoles/hr of carbon monoxide are produced.

At the given conditions the carbon dioxide produced then undergoes water gas shift reaction given by balanced equation 2.

From balanced equation 2:

$$CO + H_2O \leftrightarrow H_2 + CO_2$$

CO converted = 88.65 kgmoles/hr.

For 1 kgmole CO conversion 1 kgmole of water is consumed.

Thus,

Water consumed = 88.65 kgmoles/hr.

For 1 kgmole CO conversion 1 kgmole of hydrogen is produced.

Thus, hydrogen produced = 88.65 kgmoles/hr.

For 1 kgmole CO conversion 1 kgmole of carbon dioxide is produced.

Thus, carbon dioxide produced = 88.65 kgmoles/hr

The net amounts from both reactions are displayed in the form of a table below.

Component	***Produced/Consumed***
CH_4	-92.26
H_2	366.41
CO	3.07
CO_2	88.26
H_2O	-180.53

Now for the product stream composition calculation:

General material balance equation (given below) has been used for individual specie.

$$moles\ of\ i\ in = moles\ of\ i\ out + moles\ of\ i\ consumed + moles\ of\ i\ accumulated$$

Assuming no accumulation and at steady state this equation is reduced to

$$moles\ of\ i\ out = moles\ of\ i\ in - moles\ of\ i\ consumed$$

Since in this case two reactions are taking place we will add or subtract the terms accordingly i.e. the product of one reaction is being consumed in second reaction.

Balance on Methane (CH_4):

Moles of methane out = 1537.79 – 92.26

Moles of methane out = 1445.53 kgmoles/hr.

Balance on Hydrogen (H_2):

Moles of hydrogen out = 33.34 + 276.80 + 88.65

Moles of hydrogen out = 399.75 kgmoles/hr.

Balance on Carbon monoxide (CO):

Moles of carbon monoxide out = 46.10 + 92.27 – 88.26

Moles of carbon monoxide out = 50.11 kgmoles/hr.

Balance on Carbon dioxide (CO_2):

Moles of carbon dioxide out = 187.38 + 88.26

Moles of carbon dioxide out = 275.64 kgmoles/hr.

Balance on water (H_2O):

Moles of water out = 3690.72 – 92.27 – 88.26

Moles of water out = 3510.19 kgmoles/hr.

Total Moles in outlet stream = 1445.53 + 399.75 + 359.10 + 50.11 + 275.64 + 3510.19

Total Moles in outlet stream = 6040.32 kgmoles/ hr.

For the composition of the stream following formula has been used:

$$Percentage\ of\ component\ (i) = \frac{moles\ of\ (i)\ in\ the\ stream}{total\ moles\ of\ the\ stream\ in\ consideration} \times 100$$

$$\text{Percentage of methane} = \frac{1455.53}{6040.32} \times 100$$

Percentage of methane in product = 23.93 %

Similarly,

Percentage of hydrogen in product = 6.61 %

Percentage of nitrogen in product = 5.96 %

Percentage of carbon monoxide in product = 0.83 %

Percentage of carbon dioxide in product = 4.56 %

Percentage of water in product = 58.11 %

The results are tabulated below for the product stream.

Pre-Reformer Outlet (Primary Reformer Feed)

Component	Percentage (%)	kgmoles/hr
CH_4	23.93	1445.53
H_2	6.61	399.75
N_2	5.96	358.10
CO	0.83	50.11
CO_2	4.56	275.64
H_2O	58.11	3510.19
Total	**100**	**6040.32**

Dry basis methane is found by ignoring the water content of the stream i.e.

$$Dry\ basis\ component\ percentage = \frac{moles\ of\ i}{total\ moles\ of\ stream - moles\ of\ water} \times 100$$

$$\text{Dry basis methane at inlet} = \frac{1537.79}{5853.44 - 3690.72} \times 100$$

$$\text{Dry basis methane at inlet} = 71.10\%.$$

Similarly,

$$\text{Dry basis methane at oulet} = \frac{1445.53}{6040.32 - 3510.19} \times 100$$

$$\text{Dry basis methane at oulet} = 57.13\%$$

Dry Basis Methane	*(%)*
Inlet of Pre-Reformer	**71.10**
Outlet of Pre-Reformer	**57.13**

Similar procedure was followed for the primary reformer and the results are shown below:

Primary Reformer (70% methane conversion)

Component	*Produced/Consumed*
CH_4	-1008.64
H_2	3418.27
CO	615.89
CO_2	392.97
H_2O	-1351.57

Primary Reformer Outlet (Secondary Reformer Feed)

Component	Percentage (%)	kgmoles/hr
CH_4	5.35	432.27
H_2	47.23	3818.02
N_2	4.42	356.95
CO	8.23	665.06
CO_2	8.27	668.41
H_2O	26.52	2143.58
Total	**100.00**	**8084.29**

Dry Basis Methane	***(%)***
Inlet of Pre-Reformer	**57.13**
Outlet of Pre-Reformer	**7.28**

Calculation for flue gases

The flue gases composition is calculated by assuming complete combustion of the fuel gas having the following composition.

Component	Percentage (%)	kgmoles/hr
CH_4	71.98	741.39
N_2	20	206.00
CO_2	8.02	82.61

Total	**100**	**1030.00**

Combustion takes place according to the following equation:

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$$

Methane in fuel = 741.39 kgmoles/hr.

So, moles of oxygen required for complete combustion = 1482.78 kgmoles/hr.

Since 5% excess air is fed to ensure complete combustion.

$$Excess\ Air\ (\%) = \frac{Actual\ air\ supplied - Stoichiometric\ air\ required}{Stoichiometric\ air\ required} \times 100$$

Stoichiometric air required = 1482.78/0.2095 kgmoles/hr.

Stoichiometric air required = 7077.78 kgmoles/hr.

For 5% excess air to ensure complete combustion, air supplied = 7431.66 kgmoles/hr

And the balance and composition of the outlet flue gases is:

Component	***Produced/Consumed***
CH_4	-741.39
CO_2	741.39
H_2O	1482.79
O_2	-1482.79

Flue gases outlet from primary reformer (to convection section):

Component	Percentage (%)	kgmoles/hr
N_2	72.50	6080.71
CO_2	9.82	824.00
H_2O	17.68	1482.79
O_2	0.88	74.14
Total	**100.00**	**8387.49**

Chapter 5

Pre-reformer Design

Pre-Reformer Design

Since in pre-reformer having natural gas feedstock, following reactions take place:

1. $CH_4 + H_2O \leftrightarrow 3H_2 + CO$
2. $CO + H_2O \leftrightarrow H_2 + CO_2$
3. $CH_4 + 2H_2O \leftrightarrow 4H_2 + CO_2$

Rate of reaction for methane is given by: (Hou & Hughes, 2001)

$rCH_4 = k[CH_4]$

And,

$k = r_1 + r_3$

Here,

$$\mathrm{r_1} = \frac{\mathrm{k1.}\left(\frac{\mathrm{PCH_4.PH_2O^{0.5}}}{\mathrm{PH_2}^{1.25}}\right).\left(1 - \frac{\mathrm{PCO.PH_2}^{3}}{\mathrm{kp1.PCH_4.PH_2O}}\right)}{\mathrm{Den^2}} \qquad \text{Eq. 1.1}$$

$$\mathrm{r_3} = \frac{\mathrm{k3.}\left(\frac{\mathrm{PCH_4.PH_2O}}{\mathrm{PH_2}^{1.75}}\right).\left(1 - \frac{\mathrm{PCO_2.PH_2}^{4}}{\mathrm{kp3.PCH_4.PH_2O^2}}\right)}{\mathrm{Den^2}} \qquad \text{Eq. 1.2}$$

$$\mathrm{Den} = 1 + \mathrm{KCO}\,.\ \mathrm{PCO} + \mathrm{KH}\,.\ \mathrm{PH_2}^{2.75} + \mathrm{KH_2O}\,.\left(\frac{\mathrm{PH_2O}}{\mathrm{PH_2}}\right) \qquad \text{Eq. 1.3}$$

$$kp1 = 1.198 \times 10^{17} \times e^{(-26830/T)} \qquad \text{Eq. 1.4}$$

$$kp3 = 2.117\ x\ 10^{15}\ \mathrm{x}\ e^{(-22430/T)} \qquad \text{Eq. 1.5}$$

Temperature	**(K)**	798.0000
Pressure	**(kPa)**	3800.000
P_{CH4}	**(kPa)**	998.3020
P_{H2O}	**(kPa)**	2395.924
P_{H2}	**(kPa)**	21.64400
P_{CO}	**(kPa)**	30.02000
P_{CO2}	**(kPa)**	121.6460
k1	**(m^3/Kg cat.s.$(kPa)^{0.25}$)**	0.0000120

k3	**(m^3/Kg cat.s.(kPa)$^{0.25}$)**	0.0000741
KCO	**(kPa^{-1})**	0.005000
KH$_2$	**(kPa$^{-2.75}$)**	0.005500
KH$_2$O	**-**	0.836900

Using equations from 1.1 to 1.5 following values were obtained.

kp$_1$	**(kPa)^2**	299.7860
kp$_3$	**(kPa)^2**	1314.255
Den	**-**	119.6469
r$_1$	**(m^3/Kg cat . s)**	8.76×10^{-7}
r$_1$	**(m^3/Kg cat . h)**	0.003154
r$_3$	**(m^3/Kg cat . s)**	5.69×10^{-5}
r$_3$	**(m^3/Kg cat . h)**	0.205157

Now,

Using,

$$\int_0^x \frac{dx}{r} = \frac{W}{F} \qquad \text{--- eq.1.6}$$

To find required catalyst weight. (Harriot, 2003)

$$r = 0.2072 \times [CH_4] \qquad \left(\frac{kmol}{kg\ cat.h}\right)$$

From figure 1,

6% conversion can be achieved at this temperature, pressure and steam to carbon ratio.

Temperature = 525^{o}C

Pressure = 38 bar.

Steam/Carbon, S/C = 2.4

x = 0.06

Equation 1.6 after putting values becomes

$$\int_0^{0.06} \frac{dx}{0.2072[CH_4]} = \frac{W}{5808}$$

As we know,

$[CH_4] = [CH_4'] \times (1-x)$

Rearranging, for "W" i.e. weight of the catalyst.

$$W = 5808 \times \int_{0}^{0.06} \frac{dx}{0.2072[CH_4'] \times (1-x)}$$

Now, finding $[CH_4']$.

As,

$$PV = nzRT$$

$$[CH_4'] = \frac{n_{CH_4}}{V} = \frac{P_{CH_4}}{zRT}$$

P_{CH4}	**(kPa)**	998.30
z	**-**	0.9988
R	**(kPa.m^3/mol.K)**	0.0083
T	**(K)**	798.00

$[CH_4']$	**(kmol/m^3)**	**0.150562882**

Integrated expression after applying limits becomes,

$$W = \frac{5808}{0.2072 \times 0.1534} \times (-\ln(1-0.06))$$

Weight of Catalyst required, W	**(Kg)**	**11511.09763**

Since,

Density of catalyst	(kg/m^3)	980

$$V = \frac{W}{\rho}$$

Volume	**(m^3)**	**11.74601799**

Pressure Drop Calculation:

Assuming L/D ratio of 2.5

Diameter of bed	**(m)**	1.815
Length of catalyst bed	**(m)**	4.538
Volumetric Flow Rate	**(m^3/s)**	2.828
Flow Area	**(m^2)**	2.589
Bed voidage, Ɛ	**-**	0.500
Superficial velocity	**(m/s)**	2.185

For pressure drop due to catalyst following equations will be used:

$$\Delta P_c = \frac{f_p \times L \times \rho \times {v_s}^2 \times (1-\varepsilon)}{D_p \times \varepsilon^3} \quad ---Eq.2.1$$

$$f_p = \frac{150 \times (1-\varepsilon)}{R_{ep}} + 1.75 \quad ---Eq.2.2$$

$$R_{ep} = \frac{D_p \times v_s \times \rho}{\mu} \quad ---Eq.2.3$$

$$D_p = 6 \times \frac{V}{S_a} \quad ---Eq.2.4$$

Sa	**(m^2)**	0.001630
V	**(m^3)**	1.57×10^{-06}

Using eq. 2.4

D_p	**(m)**	**0.005807631**

ρ	**(Kg/m^3)**	9.769
μ	**(Pa.s)**	0.000025850
Ɛ	-	0.500
v_s (m/s)		2.185

Using eq. 2.3

R_{ep}	**=**	**9588.825**

Using eq. 2.2

f_p	**=**	**1.766**

Using eq. 2.1 i.e.

$$\Delta P_c = \frac{f_p \times L \times \rho \times {v_s}^2 \times (1-\varepsilon)}{D_p \times \varepsilon^3}$$

ΔP_c	**(Pa)**	257277.19
ΔP_c	**(bar)**	**2.53912849**

Inert Alumina Balls

Inert alumina balls must be placed on top of the catalyst to achieve proper distribution of gas and catalyst bed from being disturbed due to consistent high pressure and temperature. Considering the catalyst shape and geometry in view 45 cm thick bed of 4" inert balls will be employed here.

Diameter of bed	**(m)**	1.815
Height of inert balls bed	**(m)**	0.450
Diameter of the particles	**(m)**	0.102
Superficial velocity	**(m/s)**	2.731
Bed voidage, Ɛ	**-**	0.400

Using eq. 2.3

R_{ep}		174738.60
f_p		1.75
ΔP_{b1}	**(Pa)**	5295.47
ΔP_{b1}	**(bar)**	0.05

Similarly for bottom a 30cm bed of inert balls of 4" diameter inert alumina balls are used.

ΔP_{b2}	**(Pa)**	3530.32
ΔP_{b2}	**(bar)**	0.03

Total pressure drop is the sum of the pressure drop due to the inert balls and catalyst.

ΔP_t	**(bar)**	**2.63**

Reformer Outlet Temperature Calculation:

Since, the operation is adiabatic so no energy flows into and out of the system. The energy is consumed or generated only by the reaction. Following reactions take place.

1. $CH_4 + H_2O \leftrightarrow 3H_2 + CO$ $\qquad$ *ΔH = +205kJ/mol*
2. $CO + H_2O \leftrightarrow H_2 + CO_2$ $\qquad$ *ΔH = -41.0kJ/mol*

In order to determine the outlet stream temperature we made following assumptions:

- The heat capacity of the stream remains constant throughout the reactor.
- Heat losses through the reactor walls are minimal.
- The relative reaction extents are fixed for the reactor conditions.

Using following relation.

$$T_{out} = T_{in} + \left(\frac{((-\Delta H_{R1}) \times \in_1 + (-\Delta H_{R2}) \times \in_2) \times C_{CH_4{}'} \times x_{CH_4}}{\rho_m \times C_p} \right) \quad \text{--- eq. 3.1}$$

c_p	**(kJ/kg-mol-C)**	45.375
Conversion (x)		0.06
ε_1		1
ε_2		0.96
$(-\Delta H_{r1})$	**(kJ/kmol)**	-209000
$(-\Delta H_{r2})$	**(kJ/kmol)**	41000
ρ_m	**(kg-mol/m^3)**	0.573
$C_{CH4'}$	**(kg-mol/m^3)**	0.151
T_{in}	**(K)**	798

Putting values in equation 3.1

We get,

T_{out}	**(K)**	738.42
T_{out}	**(^oC)**	465.42
ΔT		59.58

Shell and Head thickness

There are two kinds of thicknesses that would be calculated while finding out vessel thickness. Since this is an internally pressurised vessel and can be considered thin walled vertical vessel. We would find minimum required shell thickness at circular seam weld and minimum required shell thickness at longitudinal seam weld.

1. Minimum required thickness at longitudinal seam weld (t_a) is given by:

$$t_a = \frac{P \times D_i}{S \times E - 0.6 \times P} \qquad \text{--- eq. 4.1}$$

2. Minimum required thickness at circular seam weld (t_b) is given by:

$$t_b = \frac{P \times D_i}{2 \times S \times E + 0.4 \times P} \qquad \text{---eq. 4.2}$$

Values required for this calculation are given below:

D: Internal diameter of shell	**(m)**	1.815
P: Design internal pressure (max)	**(bar)**	50.00
E: Weld efficiency		0.850
S: Allow able stress at 600 ^0C	**(bar)**	2300

Putting the values in equations 4.1 & 4.2,

From eq. 4.1:

$$t_a = \frac{50 \times 100000 \times 1.815}{2300 \times 100000 \times 0.85 - 0.6 \times 50 \times 100000}$$

From eq. 4.2:

$$t_b = \frac{50 \times 100000 \times 1.815}{2 \times 2300 \times 100000 \times 0.85 + 0.4 \times 50 \times 100000}$$

Solving these we get:

t_a	**(mm)**	47.30
t_b	**(mm)**	23.20

Minimum shell thickness	**(mm)**	47.30
Corrosion Allowance	**(mm)**	4.000
Shell Thickness	**(mm)**	51.30
Head Thickness (Top)	**(mm)**	27.20
Head Thickness (Bottom)	**(mm)**	27.20

Chapter 5

Control System

Automatic Process Control

Introduction:

Automatic process control is a system that is applied to different processes in order to control process conditions, parameters and variables and to keep them at the required values. Automatic process control can maintain temperature, pressure, process variables, flow, compositions and some other parameters at the desired value.

For any process, control is very important for many reasons. Based on the industrial experiences the most important reasons for applying a process control system include:

- It can prevent accidents or any kind of damage to plant or any personal. It keeps the process conditions under control and within the safety limits.
- It can protect environment by reducing waste and minimizing emissions.
- Safety is the most important consideration while applying control system so it ensures that the plant is always working under safety limits.
- It can maintain the product quality, colour, composition and purity etc.
- It can reduce the cost of production and ensure smooth operation.

So the automatic control systems are applied to maintain product quality, increase production rate, reduce the need of labour and at the same time provide the safe environment.

Drawbacks of poorly performing control loops:

Operability and profitability of industrial plants can be effected by the control loop. Considering the importance of control loops, one would expect that they always perform at their peak, but this is not the case. In fact, several studies have shown that roughly one third of industrial control loops perform poorly. Poorly performing control loops can make a plant difficult to operate and may have several costly side-effects, including:

1. Reduced production rate
2. Increased emissions
3. Lower efficiency

4. Plant trips following process upsets
5. Poor product quality
6. Slower start up and transition times
7. More off-spec product or rework
8. Premature equipment wear

For these reasons, control loop performance should always be kept at the highest possible level. This is achieved through continuously monitoring loop performance and taking the appropriate corrective actions when sub-optimal performance is detected.

To effectively manage, improve, or sustain control loop performance, it is important to monitor how well loops perform. Loop performance monitoring can provide valuable feedback on the success of control optimization projects; it helps maintaining a high standard of loop performance in the long run; and it can be used to pinpoint offending control loops for corrective action.

Basic Components:

There are three basic components of any control system

- Sensor-Transmitter
- Controller
- Final Control Element

Sensor and Transmitter:

A sensor element measures a process variable: flow rate, temperature, pressure, level, pH, density, composition, etc. Much of the time, the measurement is inferred from a second variable: flow and level are often computed from pressure measurements, composition from temperature measurements.

These are the primary and secondary devices respectively.

Sensor:

Sensor also called primary element. Its purpose is to detect changes in the environment and provide the resultant output. It is a device that is used to measure the process variables for example temperature, pressure, flow rate etc.

Different sensors for different purposes include:

Measurement	Sensors
Temperature	Thermocouple, Thermistors, RTD's, IC
Pressure	Fibre optic, elastic liquid based manometers, Vacuum, electronic.
Flow	Differential pressure, Electromagnetic, Thermal mass, Positional Displacement.
Level	Thermal displacement, ultrasonic radio frequency, Differential pressure.

Sensor Selection:

Sensors/ transmitters are selected based upon certain features including:

- **Range:** It is the measurement limit of the sensor
- **Calibration:** It is the extent to which the sensor measures the correct value.
- **Accuracy:** It is how precisely a sensor measures a value.
- **Resolution:** It is the smallest change in value that is measured by the sensor.
- **Response:** It is the time that a sensor takes to reach the true value when there occurs a change in the value which it is measuring.
- **Sensitivity:** It is one of the most important parameters while selecting a sensor. It is the change in output of the sensor per unit change in the value it is measuring.
- **Cost:** It depends on how much expensive sensor can one afford. It is increased with increased accuracy, better precision, longer range and longer life.

The sensor that is being used in this system for temperature measurement is thermocouple.

Thermocouple:

Thermocouple is a temperature measuring device. Thermocouples contain two legs of wires that are made up of different materials. Both of these wires are weld together on a single end, this will create a junction. This is the junction where temperature is measured. When the junction experiences a change in temperature, a voltage is created.

Transducers:

A transducer is a device that receives a signal and retransmits it in a different form. For example, we've discussed me /P transducers that convert a current signal to pneumatic form. It is also called convertor. After a value is measured by the sensor, it is the received by the transducer. Transducer is a device that receive a signal from the sensor and retransmit in a different form that is readable by the controller.

There are many different types of transducers including:

Transducer I/P: It will convert electric signals in to pneumatic ones. Input may be 4 to 20 mA and output is in 3 to 15 psig.

Signals:

There are three basic types of singles that are used in any industry:

Pneumatic Signal: Pneumatic signal is also called air pressure. It normally ranges between 3 - 15 psig. It is usually represented in diagram as,

Electric Signal: It is also the most common type of signal used. It ranges between 4 – 20 mA, 10 – 50 mA, 1 – 5 V or 0 – 10 V etc. It is usually represented as dash lines.

— — — — — — —

Digital signal: Third type is digital, or discrete signal. It is normally comprised of 0 & 1 combinations. In diagram it is represented as,

<u>Controller:</u>

Controller is a device that takes an input from the transmitter and perform accordingly to maintain the controlled variable at the set point. In this system we are using a feedback controller.

Feed Back Controller:

A feedback controller is a type of controller that takes input from the system's outlet. It compensates for all the disturbances due to which the controlled variable deviates from the set point. A feedback controller decide what to do to maintain the controlled variable at the set point. (Smith, 2005)

A feedback controller comes to a decision by solving an equation based upon the difference between controlled variable and the set point. This difference is called an error and is given by an equation:

$$e(t) = r(t) - c(t)$$

Where,

$c(t)$ = Controlled variable

$r(t)$ = Reference Point or set point

$e(t)$ = error, %TO (Transmitter Output)

In deviation variable form:

$$E(t) = R(t) - C(t)$$

Where,

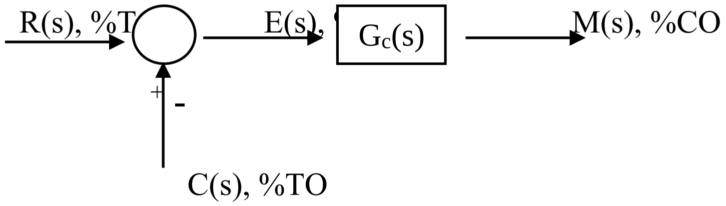
$E(t) = e(t) - 0$

$R(t) = r(t) - r$, where r is the initial steady state vale of the reference or the set point.

After taking Laplace Transformation it becomes,

$$E(s) = R(s) - C(s)$$

The basic block diagram of a feedback controller is shown as,



There are many different types of feedback controllers.

In this system a PID controller is used to achieve the required objectives.

PID Controller:

It is a proportional–integral–derivative controller (PID controller). A feedback control loop mechanism is used in industrial control systems. It calculates the error value continuously as the difference between the set point value and the controlled variable. The purpose of the controller is to minimize the error by controlling the final control element that is the control valve. It will either close or open the valve accordingly to adjust the value. (Smith, 2005)

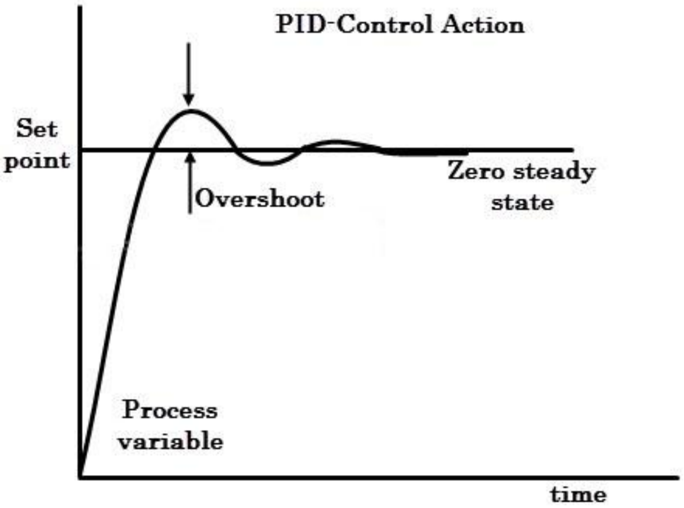
Equation:

$$m(t) = \overline{m} + K_c e(t) + \frac{K_c}{\tau_I} \int e(t)\, dt + K_c \tau_D \frac{de(t)}{dt}$$

This is the equation for PID controller. The first term shows the value of proportional action, second is of integral which completely eliminates the offset but the oscillations are increased. So, the third term derivative is added, which reduces the number of oscillations.

Purpose of selection: The PID is selected because temperature is inherently slow variable, and it takes time to develop it. So, PID is suitable for this as this will reduce the noise and the integral will remove or reduce the offset.

So the main purpose of PID is to give fast response, remove offset and less oscillations. The response is made faster by proportional part. If gain is increased the response is faster. The offset is removed by integral part. And the derivative action will control the over shoot and unnecessary oscillations.



Final Control Element:

This is the third main component of the control system. It takes an input from the controller and adjust accordingly to keep the controlled variable maintained at set point.

The valve that is used for our system is the glob valve.

GLOBE VALVE

In spite of the fact that globe valves in the past had the circular bodies which gave them their name, numerous advanced globe valves don't have a lot of a round shape. In any case, the term globe valve is still frequently utilized for valves that have such an inward component. In pipes, valves with such a component are additionally regularly called stop valves since they don't have the worldwide appearance, however the term stop valve may allude to valves which are utilized to stop stream notwithstanding when they have different instruments or outlines.

Globe valves are utilized for applications requiring throttling and regular operation. For instance, globe valves or valves with a comparable system might be utilized as inspecting valves, which are usually utilized when fluid specimens are being taken. Since the confound limits stream, they are not suggested where full, unhindered stream is required.

Globe valves are reverse seated by nature.

The body is the primary weight containing structure of the valve and the most effectively recognized as it structures the mass of the valve. It contains the majority of the valve's inside parts that will interact with the substance being controlled by the valve. The cap is associated with the body and gives the regulation of the liquid, gas, or slurry that is being controlled.

Globe valves are normally two-port valves, albeit three port valves are likewise created generally in straight-stream design. Ports are openings in the body for liquid streaming in or out. The two ports might be situated straight opposite each other or anyplace on the body, or arranged at an edge, (for example, a 90°). Globe valves with ports at such an edge are called edge globe valves. Globe valves are chiefly utilized for destructive or high thick liquids which harden at room temperature. This is on the grounds that straight valves are composed so that the outlet channel is in accordance with the gulf funnel and the liquid has a decent risk of staying there on account of level channelling. On account of edge valves, the outlet channel is coordinated towards the base. This permits the liquid to deplete off. Thus, this counteracts obstructing and/or consumption of the valve segments over a timeframe. A globe valve can likewise have a body fit as a fiddle of a "Y". This will permit the development of the valve to be directly at the base instead of the traditional pot sort development (to organize base seat) if there should be an occurrence of different valves. This will again permit the liquid to go through without trouble and minimizes liquid stopping up/consumption in the long haul.

Important Characteristics of selected valve:

- Globe valve with stem angled perpendicular to the main steam.
- It is equal percentage valve.
- In fail safe action the valve is in Fail Close position.
- Provides maximum controllability on the flow.

Calculation for Valve Sizing:

Valve used: Globe Valve.

$\text{Temperature: } 380°\text{C} = 716°\text{F}$

$\text{Optimum Pressure} = 38 \text{ bar} = 551.143 \text{ psia}$

$\text{Allowable Pressure Drop} = 10 \text{ psia}$

$\text{Downstream Pressure} = 541 \text{ psia}$

$\text{Flowrate} = 1690.72 \frac{\text{kgmol}}{\text{hr}} = 66.433 \frac{\text{m}^3}{\text{hr}} = 66432.96 \frac{\text{kg}}{\text{hr}} = 146459.606 \frac{\text{lb}}{\text{hr}}$

$\text{Valve Action} = \text{Air to to open} = \text{Fail close}$

$\text{Pressure} = 3206 \text{ psia}$

$\text{Temperature} = 1165.1 \text{ R} = 708.5°\text{F}$

$\text{Molecular Weight} = 18.02\,\frac{\text{g}}{\text{mole}}$

$\text{Ratio of Specific Heat} = 1.33$

$\text{Flow Characteristics} = \text{Equal Percentage}$

$\text{Specific Volume} = \frac{0.075\ \text{m}^3}{\text{kg}}$

$\text{Flow rate} = 19.3F_P C_v P_1 Y \sqrt{\frac{rM_w}{T_1 Z}}$

$\text{Pressure Drop Ratio} = X_T = 0.75$

$F_K = \frac{k}{1.40} = \frac{1.33}{1.40} = 0.95$

$$\text{Ratio of pressure drop to absolute relief pressure} = \frac{\Delta P}{P_1} = X = \frac{10.125}{651.143}$$

$$= 0.0184 \approx 0.02$$

$F_K X_T = 0.95 \times 0.75 = 0.71$

$$Y = \text{Expansion Factor} = 1 - \frac{x}{3 \times F_K X_T} = 1 - \frac{0.02}{3 \times 0.71} = 0.9906$$

Compressibility Factor

$$P_R = \frac{P}{P_C} = \frac{551.143}{3208} = 0.1718$$

$$T_R = \frac{T}{T_C} = \frac{839.667R}{1165.1} = 0.72$$

$\text{Using Figure 3.4} = Z = 0.9394$

$$C_v = \frac{w}{F_P P_1 Y}\sqrt{\frac{T_1 Z}{xMw}} = \frac{146459.606}{19.3 \times 551.143 \times 0.9906} \times \sqrt{\frac{839.67 \times 0.9394}{0.02 \times 18.02}}$$

$$= 13.899 \times 46.783 = 650.23$$

Valve size = 8"

So our control valve is 8 inches in size with equal percentage characteristics and fail safe action as fail close.

Important Parameters to Maintain:

In pre-reformer around 6% conversion is required, this is the important parameter that we need to control. If conversion inside the pre-reformer is increased beyond 6% then there won't be any methane slip in the secondary reformer. On the other hand if the conversion is less than 6%, then the pre-reformer outlet temperature will increase beyond specified limit and as a result high temperature process gas will be introduced in coil E4201-A. This high temperature introduction of process gas will not be able to absorb sufficient amount of energy from the coil and as a result coil may get damage. So following is the list of parameters that we want to control:

- Conversion around 6%.
- Outlet temperature of pre-reformer around 460 ^{o}C.
- Required amount of heat absorption in both newly installed coils.
- Temperature around 600 oC around primary reformer inlet.
- Maintain steam to carbon ratio such that to avoid cocking.

So here comes the need of a control system. Our system is in need of such a control system that not only meet these objectives but also be able to keep the system running smoothly and efficiently in a stable environment.

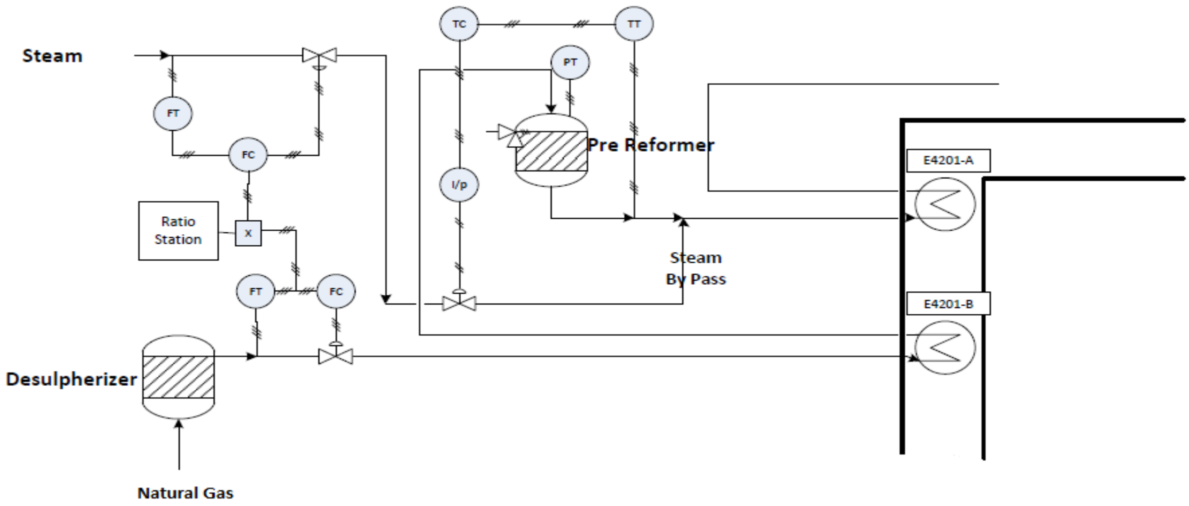
Applying Control System:

A control system is developed to address the problems faced and to meet the required objective.

Since the pre-reformer is to be introduced in already running plant, so system is required not to affect the flow rates of the system to avoid any effect on the overall production so a bypass is introduced in the system.

Control:

Now if our conversion varies we control in by varying the flow rate of steam. If conversion is increased by any other mean for example by the rise of inlet temperature then the controller will decrease the flow of steam in the main stream by opening the bypass valve. Similarly if conversion is reduced the valve is closed down and flow rate in the main stream is increased which will in turn increase the percentage of conversion pre-reformer.



If outlet temperature is increased, controller will increase the flow of steam in the main stream and conversion is increased which will result in decrease of outlet temperature.

Calculation of controller's working equation:

There are two main equations. First shows the relation between inlet outlet temperature and conversion of pre-reformer. The second gives relation between conversion and the feed flow rates. These are given as follows:

$$\int_{0}^{0.06} \frac{dx}{0.2072[CH_4]} = \frac{W}{5808}$$

$$T_{out} = T_{in} + \left(\frac{((-\Delta H_{R1}) \times \in_1 + (-\Delta H_{R2}) \times \in_2) \times C_{CH_4}{}' \times x_{CH_4}}{\rho_m \times C_p} \right)$$

Solving first equation and finding relation between steam flow and conversion

$$\int_{0}^{0.06} \frac{dx}{0.2072[CH_4]} = \frac{W}{S + 2155.95}$$

Where s is the flow rate of steam.

Substituting the known values we get:

$$S = \frac{288105.691}{(Tin)\ln(1-x)} - 2155.95 \quad --\text{A}$$

Using second equation and after substituting the known values, we get

$$x = \frac{Tin - Tout}{1186.88}$$

From here T_{in} will be comes out to be

$$Tin = 1186.88\,.x - Tout$$

Substitution these values in equation a resultant equation is:

$$S = \frac{288105.691}{(1186.88\,.x - Tout)\ln(1-x)} - 2155.95$$

Taking x = 6% we get,

$$\boldsymbol{S = \frac{288105.691}{Tout + 344.2128} - 2155.95}$$

Where T_{out} is in °C.

Controller’s action:

Since the valve that is installed is fail close and when the input to the controller i.e. outlet temperature is increased, the controller has to close the valve to increase the flow rate of steam in the main stream. So controller installed is reverse acting controller.

Working of controller is shown on the lab view. We have to main situation:

- What controller should do when the outlet temperature is increased
- What controller should do when the outlet temperature is reduced

Case 1:

When the outlet temperature is increased, then according to the derived equation,

$$\mathbf{S} = \frac{\mathbf{288105.691}}{\mathbf{Tout + 344.2128}} - \mathbf{2155.95}$$

The steam flow is less than the required value to keep the temperature at the set point, so controller will increase the steam flow rate. As a result the conversion in pre-reformer is increased and the outlet temperature come back down to the set point. As shown below.

Case 2:

When the outlet temperature is decreased, then according to the derived equation,

$$\mathbf{S} = \frac{\mathbf{288105.691}}{\mathbf{Tout + 344.2128}} - \mathbf{2155.95}$$

The steam flow is more than the required value to keep the temperature at the set point, so controller will decrease the steam flow rate. As a result the conversion in pre-reformer is decreased and the outlet temperature come back up to the set point. As shown below.

Chapter 5

Heat Exchanging Coils Sizing

Heat Exchanging Coils Calculation:

Two coils need to be sized namely E-4201A and E-4201B.

For the calculation of required surface area following assumptions are made:

1. Overall heat transfer coefficient remains constant.
2. Heat capacity of both the streams remains constant.
3. Isothermal phase change.
4. Adiabatic operation i.e. no energy lost from the exchanging coils.

Coil E-4201A:

As,

$$Q = m' C_p \Delta T \qquad \text{---eq. 5.1}$$

For cold stream, i.e. Pre-reformer product

T_{in}	**(K)**	739.15
T_{out}	**(K)**	873
m'	**(kg-mol/hr)**	6039.32
C_p	**(kJ/kg-mol.K)**	43.89

Using eq. 5.1,

$Q = 6039.32 \times 43.89 \times (873 - 739.15)$

$Q = 3.548 \times 10^7$ (kJ/hr)

For hot stream, i.e. flue gases from primary reformer

T_{in}	**(K)**	1406
T_{out}	**(K)**	-
m'	**(kg-mol/hr)**	8387.49
C_p	**(kJ/kg-mol.K)**	39.94

Using equation 5.1 in slightly modified from to calculate the outlet temperature of the hot stream.

$$T_{out} = T_{in} + \frac{-Q}{m' \times C_p} \text{---eq. 5.2}$$

Putting values in equation 5.2, it becomes

$$\mathrm{T_{out}} = 1406 - \frac{3.548 \times 10^7}{8387.49 \times 39.94}$$

$\mathrm{T_{out}} = 1300.09$ °C

Now,

$$C = m' \times C_p \text{ ---eq. 5.3}$$

$\mathrm{C_c} = 6039.32 \times 43.89$

$\mathrm{C_c} = 2.65 \times 10^5$ (kJ/hr.K)

Similarly,

$\mathrm{C_h} = 8387.49 \times 39.94$

$\mathrm{C_h} = 3.35 \times 10^5$ (kJ/hr.K)

As,

$$c'' = \frac{C_{min}}{C_{max}} \text{ ---eq. 5.4}$$

Putting values in above equation,

$$\mathrm{c''} = \frac{2.65 \times 10^5}{3.35 \times 10^5}$$

$\mathrm{c''} = 0.791$

Effectiveness is found by the relation given below.

$$\epsilon = \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \text{ ---eq. 5.5}$$

Putting values in equation 5.5

$$\epsilon = \frac{873 - 739}{1406 - 739}$$

$\epsilon = 0.201$

For cross flow arrangement NTU is given by:

$$\mathrm{NTU} = -\ln\left(1 + \frac{ln(1 - \epsilon \times c'')}{c''}\right) \text{---eq. 5.6}$$

Putting values in equation 5.6

$$\mathrm{NTU} = -\ln\left(1 + \frac{ln(1 - 0.201 \times 0.791)}{0.791}\right)$$

$\mathrm{NTU} = 0.247$

Now finding the surface area required:

$$NTU = \frac{U \times A_s}{C_{min}}$$

Putting the values in equation above:

U	**(W/m^2.K)**	93.20
C_{min}	**(kJ/hr.K)**	2.65×10^5

$$\mathrm{A_s} = \frac{\mathrm{NTU} \times \mathrm{C_{min}}}{\mathrm{U}}$$

$$\mathrm{A_s} = \frac{0.247 \times 2.65 \times 10^5 \times 1000}{93.20 \times 3600}$$

$$\mathrm{A_s} = 195.84\ \mathrm{m}^2$$

Now finding the number of tubes required:

Effective length, L_{eff} = 9.2 m

Tube diameter d_0 = 0.114 m

Since area is given by:

$A_{st} = \pi \times d \times L$

Surface area of each tube = 3.298 m^2

Number of tubes are given by:

$$N = \frac{A_s}{A_{st}}$$

Number of tubes, N = 60

Tube diameter d_0 = 0.114 m

For schedule 80 and given d_0,

Internal diameter of tube, d_i = 0.104 m

Since cross sectional area is given by:

$$A_{cross\ sectional} = \frac{\pi}{4} \times {d_i}^2$$

$A_{cross\ sectional}$ = 0.00846 m^2

$A_{\text{cross sectional (total)}} = 0.508\ m^2$

Volumetric flow rate = 10641 m^3/hr

Volumetric flow rate = 2.95 m^3/hr.

As velocity is given by:

$$u = \frac{Volumetric\ flowrate}{A_{cross\ (t)}}$$

Velocity, u = 5.818 m/s

Density	**(kg/m^3)**	9.18
Viscosity	**(Pa.s)**	0.000024
d_i	**(m)**	0.104

As,

$$N_{Re} = \frac{\rho \times d_i \times v}{\mu}$$

$N_{Re} =$ 230869.64

From Coulson Richardson Vol. 6

J_f = 0.0035

No of passes = 4

Now using equation, given below to find pressure drop.

$$\Delta Pt = Np\left(8J_f\left(\frac{L}{D}\right))(\frac{\mu}{\mu w})^{-m} + 2.5\right)\left(\frac{\rho v^2}{2}\right)$$

Considering

$(\mu/\ \mu_w) = 1$

We get:

ΔPt = 3096.340 Pa

ΔPt = 0.03096 bar

Coil E-4201B:

As,

$$Q = m'C_p\Delta T$$ ---eq. 5.1

For cold stream, i.e. Pre-reformer Feed

T_{in}	**(K)**	661
T_{out}	**(K)**	798
m'	**(kg-mol/hr)**	5853.43
C_p	**(kJ/kg-mol.K)**	43.60

Using eq. 5.1,

$Q = 5853.43 \times 43.60 \times (798 - 661)$

$Q = 3.496 \times 10^7$ (kJ/hr)

For hot stream, i.e. flue gases from primary reformer

T_{in}	**(K)**	1300.09394
T_{out}	**(K)**	-
m'	**(kg-mol/hr)**	8387.49
C_p	**(kJ/kg-mol.K)**	39.05

Using equation 5.1 in slightly modified from to calculate the outlet temperature of the hot stream.

$$T_{out} = T_{in} + \frac{-Q}{m' \times C_p}$$ ---eq. 5.2

Putting values in equation 5.2, it becomes

$$T_{out} = 1300.09 - \frac{3.496 \times 10^7}{8387.49 \times 39.05}$$

T_{out} =1193.34 °C

Now,

$$C = m' \times C_p \quad \text{Eq. 5.3}$$

$C_c = 5853.43 \times 43.60$

$\mathrm{C_c} = 2.552 \times 10^5$ (kJ/hr.K)

Similarly,

$\mathrm{C_h} = 8387.49 \times 39.05$

$\mathrm{C_h} = 3.275 \times 10^5$ (kJ/hr.K)

As,

$$c'' = \frac{C_{min}}{C_{max}} \quad \text{Eq. 5.4}$$

Putting values in above equation,

$$\mathrm{c''} = \frac{2.552 \times 10^5}{3.275 \times 10^5}$$

$\mathrm{c''} = 0.779$

Effectiveness is found by the relation given below.

$$\epsilon = \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \quad \text{Eq. 5.5}$$

Putting values in equation 5.5

$$\epsilon = \frac{798 - 661}{1300 - 661}$$

$\epsilon = 0.214$

For cross flow arrangement NTU is given by:

$$\mathrm{NTU} = -\ln\left(1 + \frac{\ln(1 - \epsilon \times c'')}{c''}\right) \quad \text{Eq. 5.6}$$

Putting values in equation 5.6

$$\mathrm{NTU} = -\ln\left(1 + \frac{ln(1 - 0.214 \times 0.779)}{0.779}\right)$$

$\mathrm{NTU} = 0.267$

Now finding the surface area required:

$$NTU = \frac{U \times A_s}{C_{min}}$$

Putting the values in equation above:

U	**(W/m²·K)**	**90**
C_{min}	**(kJ/hr.K)**	2.552×10^5
As,		

$$A_s = \frac{\mathrm{NTU}\times C_{\min}}{U}$$

$$A_s = \frac{0.214\times 2.552\times 10^5\times 1000}{90\times 3600}$$

A_s =210.54 m^2

Now finding the number of tubes required:

Effective length, L_{eff} = 9.2 m

Tube diameter d_0 = 0.114 m

Since area is given by:

$A_{st} = \pi\times d\times L$

Surface area of each tube = 3.298 m^2

Number of tubes are given by:

$$N = \frac{A_s}{A_{st}}$$

Number of tubes, N = 64

Now for pressure drop calculation:

Tube diameter d_0 = 0.114 m

For schedule 80 and given d_0,

Internal diameter of tube, d_i = 0.104 m

Since cross sectional area is given by:

$$A_{cross\ sectional} = \frac{\pi}{4}\times {d_i}^2$$

$A_{cross\ sectional}$ = 0.00846 m^2

$A_{cross\ sectional\ (total)}$ = 0.508 m^2

Volumetric flow rate = 8153 m^3/hr

Volumetric flow rate = 2.265 m^3/hr.

As velocity is given by:

$$u = \frac{Volumetric\ flowrate}{A_{cross\ (t)}}$$

Velocity, u = 4.195 m/s

Density	**(kg/m^3)**	8.89
Viscosity	**(Pa.s)**	0.000021
d_i	**(m)**	0.104

As,

$$N_{Re} = \frac{\rho \times d_i \times v}{\mu}$$

N_{Re} = 184242.26

From Coulson Richardson Vol. 6

J_f = 0.0031

No of passes = 4

Now using equation, given below to find pressure drop.

$$\Delta Pt = Np\left(8J_f\left(\frac{L}{D}\right))(\frac{\mu}{\mu w})^{-m} + 2.5\right)\left(\frac{\rho v^2}{2}\right)$$

Considering

(μ/ μ_w) = 1

We get:

ΔPt = 1470.235 Pa

ΔPt = 0.01470 bar

Chapter 6

Material of Construction

Material of Construction

While selecting the material of construction we have to consider many factors. The material should be economical and easy to work on. It should not contaminate the product during the process and should have good mechanical properties at the prevailing operating conditions. It should be corrosion resistant and should maintain its properties at high temperature for resistance to oxidation, carburization, nitriding, sulfidation and other type of high temperature corrosions.

The most important characteristics to be considered when selecting a material of construction are;

1. Mechanical Properties
    a) Tensile strength
    b) Elastic modulus /Young's modulus
    c) Fracture resistance
    d) Wear resistance
2. The effect of high and low temperatures on the mechanical properties i.e. corrosion resistance.
3. Any special properties required, such as;
    a) Thermal conductivity
    b) Electrical resistance
    c) Magnetic properties
    d) Welding
    e) Casting availability in standard sizes-plates
    f) Cost

In order to select the exact material of construction, the process environment to which the material will be exposed must be evidently clear. Additional to main corrosive chemicals present, the following factors must be considered:

- Temperature (affects corrosion rate and mechanical properties.
- Pressure

- Presence of trace impurities
- The amount of aeration/differential oxidation cells
- Stream velocity and agitation / erosion-corrosion
- Heat transfer rates / differential temperatures
- The conditions that may ascend during abnormal operation, such as at start-up and shutdown, must be considered, in addition to normal, steady state operation.

Material of construction for coils:

Incolloy 800H alloy is an austenitic, solid-solution alloy. It is suitable for the application because it can maintain its chemical properties at higher temperatures and shows significant resistance to carburization and other corrosions occurring at higher temperatures. It has greater creep and rupture strength making it suitable for high pressure cyclic loads at higher temperature for extended period of time this is due to the fact that it has higher carbon content.

It requires a carbon range of 0.05 to 0.10%. It is used in manufacturing of furnace components, pigtails, headers and cracker tubes of petrochemical furnace and sheathing of electric heating elements.

Composition

Component	Mass Percentage (%)
Nickel	30-35
Chromium	19-23
Iron	39.5
Carbon	0.05-0.10
Aluminium	0.15-0.60
Titanium	0.15-0.60

For pipe/tube manufacturing the average grain size required is ASTM 5. The physical and thermal properties of the metal are shown below in the table.

Physical properties[1]

Density	g/cm^3	7.94
Melting range	^{o}C	1357-1385
Specific heat	J/Kg.^{o}C	460
Curie temperature	^{o}C	-115
Tensile Modulus	GPa	141.3
Shear Modulus	GPa	50.7
Poisson's ratio	GPa	0.394

Thermal properties:

Electrical resistivity	µΩ.m	1.291
Coefficient of thermal expansion	µm/m.K	18.00
Thermal conductivity	W/m.K	31.90

Corrosion resistance:

It possess strength at high temperature and it is suitable for the environments which involve reactions at high temperature like oxidation and carburization. The corrosions concerned due to the particular environment of the convection section especially due to the composition of the flue gases are explained below.

Oxidation:

Nickel and chromium content is high which makes the Incolloy 800h's resistance excellent towards oxidation. The chromium present in the alloy forms a protective layer of surface oxide and nickel increase protection during cyclic exposures to high temperatures. The problem in nickel chromium alloys may occur when environment is reducing towards nickel and tends to oxidize to chromium. This can cause internal oxidation of a nickel chromium alloy. The results are extensive embrittlement of the alloy and is signified by severe oxidation of chromium leaving the remaining alloy magnetically strong. Increase in iron content from 39 to 46% can prevent the problem.

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[1] The properties are calculated at 1000 ^{o}C

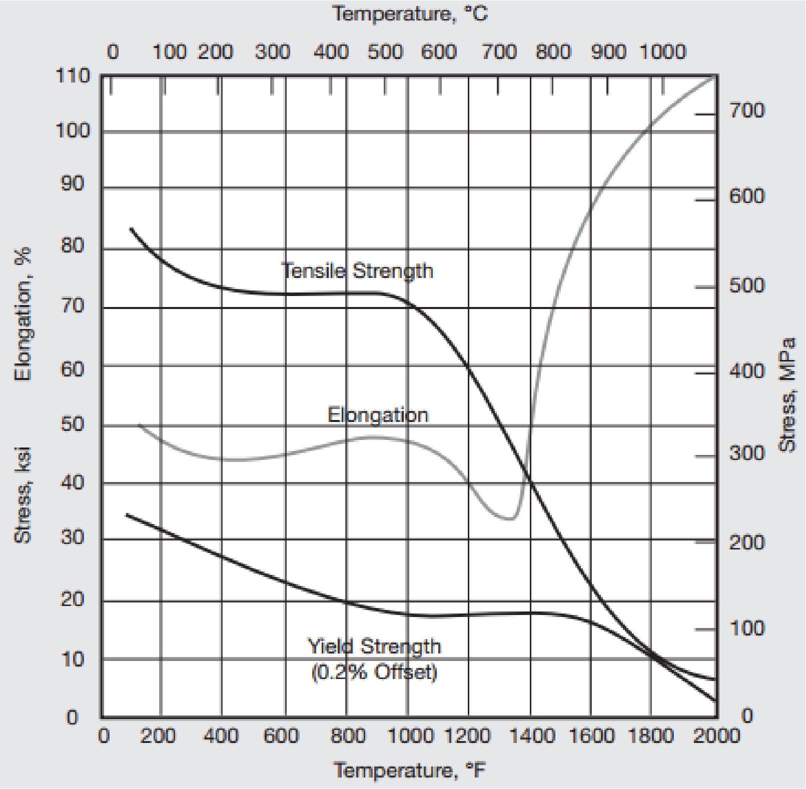
Carburization:

In this phenomenon metal carbides form in the alloy which reduces its mechanical properties and also result in depletion and waste of material. The carburization-corrosion process is also known as "**Metal Dusting"** which consequently results in formation of powdery matter due to surface damage. The environments containing hydrocarbon gases and CO are specifically hazardous for the piping. The primary barrier in the formation of surface carbides is Cr_2O_3 which forms a layer on alloy surface. Secondly the Nickel metal present in the alloy resists the ingression of carbides which try to diffuse in the material.

Nitriding:

Nitriding happens when nitrogen is absorbed in metal at higher temperatures causing the metal to **case harden**, a condition in which the layer containing nitrogen becomes harder than the remaining metal. The resistance towards Nitriding in an alloy increases with the increase in nickel content of alloy as it decrease the permeability of the alloy.

Tensile strength and yield strength curve for the material



Material of construction for Pre-reformer:

Stainless steel type 316 is an austenitic CR-Ni stainless steel which also has molybdenum. It is also known as **inox** and it contains 10.5% chromium by mass. For harsher environment 13-26% chromium is used to resist oxidation. This addition improves its corrosion resistant properties and makes it more reliable than type 304 at elevated temperatures. It shows excellent pitting resistance.

Chromium forms a passivation layer of chromium oxide which prevents the metal beneath from oxidation. This layer is impervious to air and water and reforms when scratched from the surface. Type 316 cannot be hardened by heat treatment.

It can be easily formed and drawn. Stainless steel 316 is highly resistant to acid attack, effects of basic substances and is suitable for handling organic fluids. Its typical applications are furnace equipment, heat exchangers, pressure vessels, exhaust manifolds, pharmaceutical equipment, photographic equipment, chemical equipment, digesters, tanks, evaporators and pulp and paper processing equipment. The percentage composition of S.S 316 is shown in the table below:

Composition

Component	Mass Percentage (%)
Carbon	0.02
Manganese	2
Sulphur	0.045
Phosphorus	0.030
Silicon	0.750
Chromium	16-18
Nickel	10-14
Molybdenum	2-3
Nitrogen	0.10
Iron	balance

Physical & thermal properties[2]:

Density	g/cm^3	7.99
Melting range	°C	1371-1399
Specific heat	KJ/kg.°C	0.50
Modulus of elasticity	GPa	28
Tensile strength	MPa	84
Yield strength	GPa	42
Coefficient of thermal expansion	µm/m.K	20
Thermal Conductivity	W/m.K	21.4
Electrical resistivity	µΩ.m	1.489

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[2] The properties are reported for 1000°C.

Chapter 7

HAZOP Analysis

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 risks associated with the process.

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.

Since 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.

Case 1:

Intention: Keeping the temperature at pre-reformer inlet at about 525^{o}C.

Deviation: The temperature rises to about 600^{o}C.

Causes: Since the feed to pre-reformer is being heated in coil E-4201B prior to its introduction in the pre-reformer. Thus this depends on the temperature of the flue gases and the heat exchange in coil E-4201A as a lower heat transfer there (due to whatever reason) would mean higher temperature of the flue gases (in contact with coil E-4201B) thus leading to higher feed temperature. Also an increase in primary reformer firing will result in higher flue gases temperature, this would also increase the temperature of the feed.

Consequences: Although 600^{o}C is within the safe operational limits but operation at this temperature would result in higher conversion than intended. Since the reaction taking place in pre-reformer is endothermic and its equilibrium is shifted towards right at higher 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.

Case 2:

Intention: Maintaining the steam to carbon ratio (S/C) at 2.4.

Deviation: Steam to carbon ratio falls below 2.0

Causes: Problem in steam generation system upstream or control system maintaining the ratio malfunctions. Bypass control valve failure may also lead to this deviation.

Consequences: As steam to carbon ratio is critical in determining the conversion achievable in the pre-reformer and a lower value of S/C would lead to lower conversion in the reformer this would lead to higher methane slip from the reforming section. Also steam to carbon ratio is critical to prevent carbon formation and deposition as at lower steam flow rates methane produces carbon by the following reaction.

$$CH_4 \rightarrow C + 2H_2$$

Hazard: Carbon deposition on the catalyst surface would render it in effective. This can cause serious economic loss. Also higher methane slip would be uneconomical.

Case 3:

Intention: Maintaining the pressure at a certain range (28 to 32 bar).

Deviation: The pressure exceeds that range.

Causes: The increase in pressure is due to the fouling in exchanger tubes, preheating the pre reformer product. This will cause pressure build up in the vessel. Another reason could be the malfunctioning of pressure control valve.

Consequence: The pressure build up will render the reaction kinetics. It will also cause less conversion as the reaction requires low pressure. Less conversion in the pre reformer will increase the temperature in the vessel.

Hazards: The increase in temperature will cause coking in the pre reformer causing the deactivation of the catalyst. This will result in serious economic loss. Also the pressure build up can rupture the tubes or the vessel.

Case 4:

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.

Chapter 8

Economic Analysis

Economic Analysis

For economic analysis we have used Lang factor method. **(Sinnott, 2005)**

Reactor:

For bare vessel cost we will use the vessel dimensions i.e.

Length	1.875 m
Diameter	6.950 m

Purchase cost = Cost from figure × Material factor × Pressure factor

Cost factor = $18,000

Material factor = 2.0

Pressure factor = 1.8

Purchase cost = 18,000×2×1.8

Purchase cost = $64,800

Catalyst Cost = $47.8/lit

Transport cost from the vendor = 5 % of the capital.

Catalyst delivered cost = $50.20

Since catalyst requirement for this application is = 11750 lit

So total cost of catalyst = $589,750

This will be treated as annual maintenance cost considering the life of catalyst to be 3 years.

Misc. Vessel internals cost = 0.25×bare vessel cost

Misc. vessel internals cost = 0.25×64800

Misc. vessel internals cost = $16,200

Reactor cost total = $81,000

Exchanger coils:

The total cost of heat exchanger is given by:

$$C = f_f \times f_m \times f_p \times C_b \qquad \text{Eq. 5.1}$$

Where,

f_f = fin factor.

f_m = material factor.

f_p = pressure factor.

C_b = tube bundle cost factor.

C = total cost of finned tube heat exchange coil.

For Coil E-4201A:

Area used in these calculations is in sq. ft.

Since,

Area = 195 m^2

Area = 2097 ft^2

Fin factor:

$$f_f = \exp(-1.1156 + 0.0906 \times ln(A)) \qquad \text{Eq. 5.2}$$

Putting values in above equation:

$f_f = \exp(-1.1156 + 0.0906 \times \ln(2097))$

$f_f = 0.655$

Material factor:

$$f_m = g_1 + g_2 \times \ln(A) \qquad \text{Eq. 5.3}$$

For Incolloy 800H,

$g_1 = 1.1854$

$g_2 = 0.49706$

Putting values in Eq.5.3

$f_m = 1.1854 + 0.49706 \times \ln(2097)$

$f_m = 4.987$

Pressure factor:

For 20-42 bar pressure range

$$f_p = 1.0305 + 0.07140 \times \ln(A) \quad \text{Eq.} 5.4$$

Putting values in E.q.5.4

$f_p = 1.0305 + 0.07140 \times \ln(2097)$

$f_p = 1.5799$

Tube bundle cost:

$$C_b = \exp(8.58 - 0.3058 \times ln(A) + 0.0681 \times ln(A)^2) \qquad \text{Eq.} 5.5$$

Putting values in Eq.5.5

$C_b = \exp(8.58 - 0.3058 \times \ln(2097) + 0.0642 \times \ln(2097)^2)$

$C_b = 27000$

Now putting all values in Eq. 5.1

$C = 0.655 \times 4.987 \times 1.5799 \times 27000$

C = \$139,500

For Coil E-4201B:

Area used in these calculations is in sq. ft.

Since,

Area = 210 m^2

Area = 2260 ft^2

Fin factor:

Putting values in Eq.5.2:

$f_f = \exp(-1.1156 + 0.0906 \times \ln(2260))$

$f_f = 0.659$

Material factor:

For Incolloy 800H,

$g_1 = 1.1854$

$g_2 = 0.49706$

Putting values in Eq.5.3

$f_m = 1.1854 + 0.49706 \times \ln(2260)$

$f_m = 5.024$

Pressure factor:

Putting values in E.q.5.4

$f_p = 1.0305 + 0.07140 \times \ln(2260)$

$f_p = 1.581$

Tube bundle cost:

Putting values in Eq.5.5

$C_b = \exp(8.58 - 0.3058 \times \ln(2260) + 0.0642 \times \ln(2260)^2)$

$C_b = 29150$

Now putting all values in Eq. 5.1

$C = 0.659 \times 5.024 \times 1.581 \times 29150$

C = \$152,500

Now that major equipment cost has been calculated, now installed equipment cost would be determined.

$$ICE = (1 + f_1 + f_2 + f_3 + f_4) \times PCE \qquad \text{Eq.5.6}$$

ICE = Installed cost of equipment.

PCE = Purchased cost of equipment.

f_1, Equipment erection

$f_1 = 0.40$

f_2, Piping

$f_2 = 0.70$

f_3, Instrumentation

$f_3 = 0.20$

f_4, Electrical

$f_4 = 0.10$

Putting values in Eq.5.6

$ICE = (1 + 0.4 + 0.7 + 0.2 + 0.1) \times 373000$

$ICE = \$895{,}200$

Now for fixed capital:

$$Fixed\ capital\ cost = (1 + f_5 + f_6 + f_7) \times ICE \quad \text{Eq.5.7}$$

f_5, Design and engineering

$f_5 = 0.3$

f_6, Contractor fee

$f_6 = 0.05$

f_7, Allied charges

$f_7 = 0.05$

Putting values in equation 5.7

$Fixed\ capital\ cost = (1 + 0.3 + 0.05 + 0.05) \times 895200$

$Fixed\ capital\ cost = \$1{,}254{,}000$

Operating Cost:

Now for operating cost i.e. annually recurring costs:

Category	Percentage of Fixed capital (%)
Maintenance	6.5
Capital charges	10
Insurance	1
Royalties	1.5

Operating cost = \$238,300

Cost of catalyst would also be treated as annual i.e. considering the catalyst life time of 2 years.

Since total cost of catalyst = \$589,750

Annual cost of catalyst = \$295,000

Total Operating Cost = 295000+238300

Total Operating Cost = \$533,300

Since after the modification protection steam would no longer be needed thus saving the cost of steam generation.

Assuming 90% plant attainment.

Operational days = 365-37

Operational days = 328

Steam generation cost = \$13.2/ton

Annual cost savings = \$13.2×20×24×328

Annual cost savings = \$2,078,208

Payback period:

$\text{Initial investment} = \ \$1{,}254{,}000$

$\text{Net annual cash flow for year 1} = \ \$2{,}078{,}208 - \ \$533{,}300$

$\text{Net annual cash flow for year 1} = \$1{,}544{,}908$

Since this value is greater than the initial investment thus to find payback period we need to simply divide the investment and net cash flow. (Leland & Anthony, 2002)

$$\text{Payback Period} = \frac{1254000}{1544908}$$

$$\text{Payback Period} = \ 0.81\ years$$

Conclusion

Conclusion

Such a short payback period suggests that this project should be pursued and it is hereby concluded that any such modification would not only save millions for the industry but will also reduce energy consumption, since our country is in dire need of energy and along with production of more energy from various sources we have to optimize existing processes to properly utilize what we have at the moment.

Apart from energy efficiency improvement this modification has various other benefits which include:

1. Overall plant capacity increased by 25%.
2. Primary reformer catalyst life enhanced.
3. Flexibility in terms of feed stock.
4. Flexibility of primary reformer operation.

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