USE OF ALTERNATE GAS FOR FFC UREA PLANT IN PLACE OF MARI GAS

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CERTIFICATE

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DEDICATION

TO OUR PARENTS

Without whom none of this would have been possible and for their support throughout our lives.

AND TEACHERS

For inspiring us and supporting us throughout the entirety of this project.

ACKNOWLEDGEMENTS

First of all, we would like to express our utmost gratitude to **Almighty Allah**. Without His mercy, we would never been able to complete our project.

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ABSTRACT

Mari gas is a depleting non-renewable resource and needs to be replaced as a raw material for manufacturing of Urea. The project aims to anticipate the impacts of using a mixture of 70% Mari gas and 30% RLNG in a pre-existing ammonia plant. Material and energy balance applied on the system reviled the hydrogen production and thermal demand of the system increased. The HYSYS simulation coincided with our findings. The primary and secondary reformers, HTS and LTS reactors, Methanator and Ammonia reactor are all redesigned with some main heat exchanges. The cost analysis and HAZOP analysis of the plant is also carried out.

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CHAPTER 1

INTRODUCTION

1.1 Scope of the Project:

The Fauji Fertilizer Company (FFC) is the one of growing manufacturer of urea in Pakistan with a total capacity of 2 million tonnes per annum [1]. FFC has three urea plants under its flag, two of them (Plant-I and Plant-II) in Goth Machhi and one (Plant-III) is present in Mirpur Mathelo. The ammonia and urea capacity of each plant is given in the following table [2]:

Sr. No.	Plant Name	Plant Location	Ammonia Production (MT/Year)	Urea Production (MT/Year)
1.	Base Unit-Goth Machhi	Goth Machhi	403,000	695,000
2.	Expansion Unit-Goth Machhi	Goth Machhi	363,000	635,000
3.	Mirpur Mathelo Unit	Mirpur Mathelo	413,000	718,000

Table 1 FFC Ammonia and Urea Production Stats

The scope of this project covers the FFC urea plants located in Goth Machhi. Ammonia is used as a raw material for the production of urea. The FFC manufactures ammonia which is used in the production of urea by the reforming of Mari gas. Mari gas is a non-renewable natural resource which is depleting due to high consumption rates. FFC plant 1 and 2 utilize 184 (MMSCFD) of Mari gas [3]. The depletion of Mari reserves at Daharki has cause FFC to look for alternatives for its raw material from Mari gas to something else. Meanwhile the load shedding of Mari gas to the production plants continues. The accessible alternative to deal with the shortfall of the process gas is to import re-gasified liquified natural gas (RLNG) and use a mixture of Mari gas and LNG as the process gas for production of ammonia and urea subsequently. But introduction of RLNG to the feed gas changes its composition hence changing the operational requirements of the process.

Mari gas supplied to FFC plant 1 and 2 from Daharki has following composition:

Mari gas Components	Moles
Nitrogen	337
Carbon dioxide	149.73
Methane	1494.5
Ethane	3.75
Propane	0
Butane	0

Table 2 Mari gas Composition

The Mari gas supplied to FFC urea plant is utilized as process gas as well as fuel. The focus of this research project is to determine a ratio of Mari gas and RLNG as a process gas which shall produce profitable urea without major alterations to the standing Ammonia-Urea plant.

The project entails a detailed study and analysis of using mixture of RLNG and Mari gas as a feed stream to the plant providing a cost and safety analysis to anticipate the feasibility of using RLNG as a partial substitute of Mari gas. The focus of the project is to investigate and anticipate the impact of changing feed on the plant and how will it affect the production of urea in terms of plant's operating conditions and economics.

1.2 Introduction to Ammonia:

Ammonia as a chemical compound having no colour with a piquant odourl and alkaline in nature. Naturally, ammonia is very scarce and is produced as a result of degradation of organic molecules like proteins present in plants and animal wastes. Ammonia is consistently and constantly is being produced in soil based on the rate of microbial activity and conditions of the soil [4]. Artificially ammonia is produced majorly using the Haber's process. Ammonia is the among the top ten large scale commercially produced chemicals and around 130 million tonnes of ammonia are produced annually [5]. Majorly ammonia is used to produce fertilizers as it is too volatile to be directly introduced to plants as a nitrogen substitute.

1.2.1 Chemical and Physical Properties of Ammonia:

Ammonia is a polar covalent compound which has hydrogen bonding among its molecules. Ammonia is a good organic solvent due to its lower dielectric constant. Following are the chemical and physical properties of ammonia:

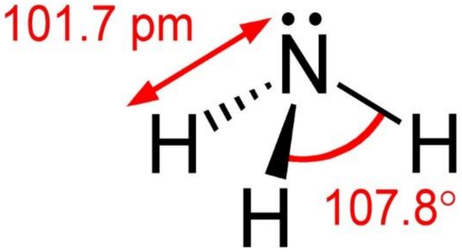
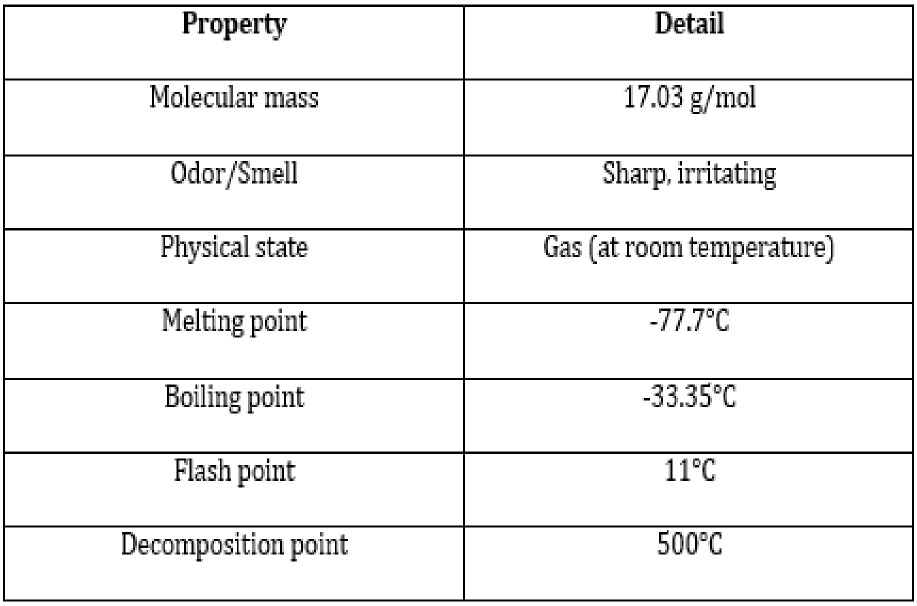
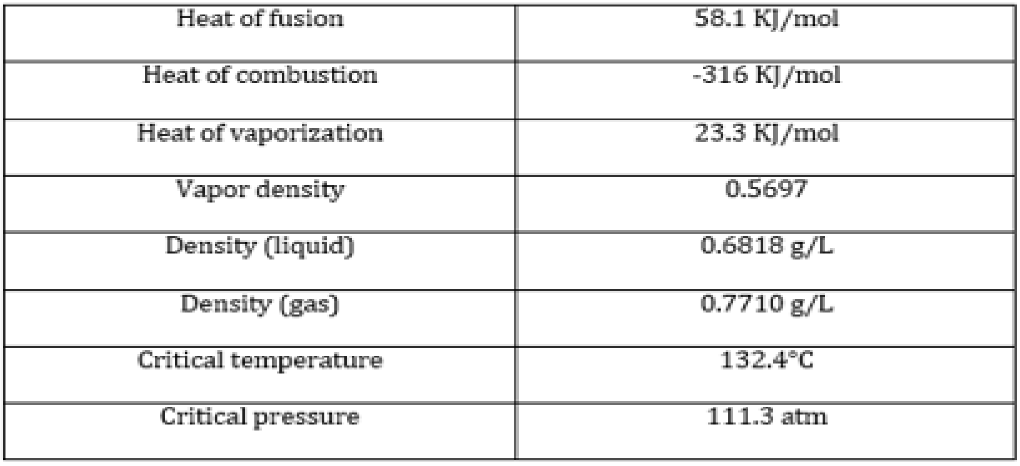


Figure 1 Structure of Ammonia molecule



Property	Detail
Molecular mass	17.03 g/mol
Odor/Smell	Sharp, irritating
Physical state	Gas (at room temperature)
Melting point	-77.7°C
Boiling point	-33.35°C
Flash point	11°C
Decomposition point	500°C



Heat of fusion	58.1 KJ/mol
Heat of combustion	-316 KJ/mol
Heat of vaporization	23.3 KJ/mol
Vapor density	0.5697
Density (liquid)	0.6818 g/L
Density (gas)	0.7710 g/L
Critical temperature	132.4°C
Critical pressure	111.3 atm

Table 3 Physical Properties of Ammonia

Property	**Details**
Chemical formula	NH_3
Reactivity	Quite reactive
Type of base	Weak
Affinity (water)	High
Oxidation power	Strong reducing agent
Volatility	Increases with increase in pH
Corrosiveness	Corrosive to some metal

Table 4 Chemical Properties of Ammonia

1.2.2 Chemical Reactivity of Ammonia:

1.2.2.1 Combustion:

Combustion of Ammonia is difficult as it requires a lot of energy to produce N_2 gas and water.

$$4NH_3 + 3O_2 + \text{heat} \rightarrow 2N_2 + 6H_2O$$

However, in the presence of catalyst NOx's are produced

1.2.2.2 Reaction with water:

Ammonia on it own dissolve in water with release of heat.

$$NH_3 + H_2O \rightleftharpoons NH_4^+ + OH^-$$

NH_4OH is basically aqueous solution of ammonia and is basic in nature. In equilibrium

Molar solution of 1.0 of NH3 produces 4.2 milli moles of hydroxide ion.

1.2.2.3 Self-dissociation Reaction:

Ammonia undergoes auto ionization like water to its acid and base conjugate.

$$2NH_3\ (\text{aq}) \rightleftharpoons NH_4^+\ (\text{aq}) + NH_2^-\ (\text{aq})$$

NH_3 is a weak base.

1.2.3 Applications and Uses of Ammonia:

Following are the major applications of ammonia:

1.2.3.1 Fertilizer Production:

Mainly ammonia is used in the manufacturing and production of fertilizers. It serves as a best nitrogen based nutrient supply for the soil which helps in production of food crops. Ammonia cannot be directly introduced to the soil due to its volatility but is 1.2.3.2 Water Purification:

Ammonia is used in purification of water supplies. It is used to produce mono-chloro-amine which is used as a disinfectant and helps in purifying water thus reducing the risk of water borne diseases.

1.2.3.3 Cleaner purposes:

Ammonia is used as cleaning purposes in household which is usually ammonium hydroxide. It is mostly used in cleaning of stainless steel, ovens and soaking things. The ammonia used in household usually consist of 5 to 10 % of ammonia.

1.2.3.4 Fermentation process:

In fermentation industry the solution of ammonia (having 16% to 24% ammonia) is used in industry of fermentation. They are the source of N_2 for microorganism and vary pH level during fermentation process,

1.2.3.5 Refrigeration:

It is used as a refrigerant gas and used in ACs because it can adsorb huge amount of heat from outside refrigerator.

1.2.3.6 Textile:

Ammonia can also be used in the making of synthetic fibre such rayon and nylon fabric. It can be used in dyeing and scouring of silk, cotton, and wool.

1.2.3.7 Other Industrial Uses:

Many products are made of ammonia which includes plastics, explosives, fabrics, dyes, and pesticides. It is used as stabilizer to carry out several functions. It is used in the manufacturing of compounds like nitric acid, Hydrogen cyanide, Ammonium carbonate and many other compounds.

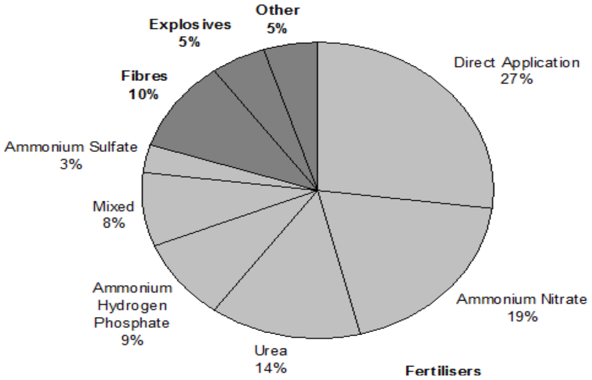


Figure 2 Applications of Ammonia

1.2.3.8 Ammonia as Energy Transporter:

Ammonia due to its flexible nature and being carbon free carrier offers good energy density as compared to compressed air. These properties of ammonia suggest that it can be useful in the future for the transportation purposed and power houses.

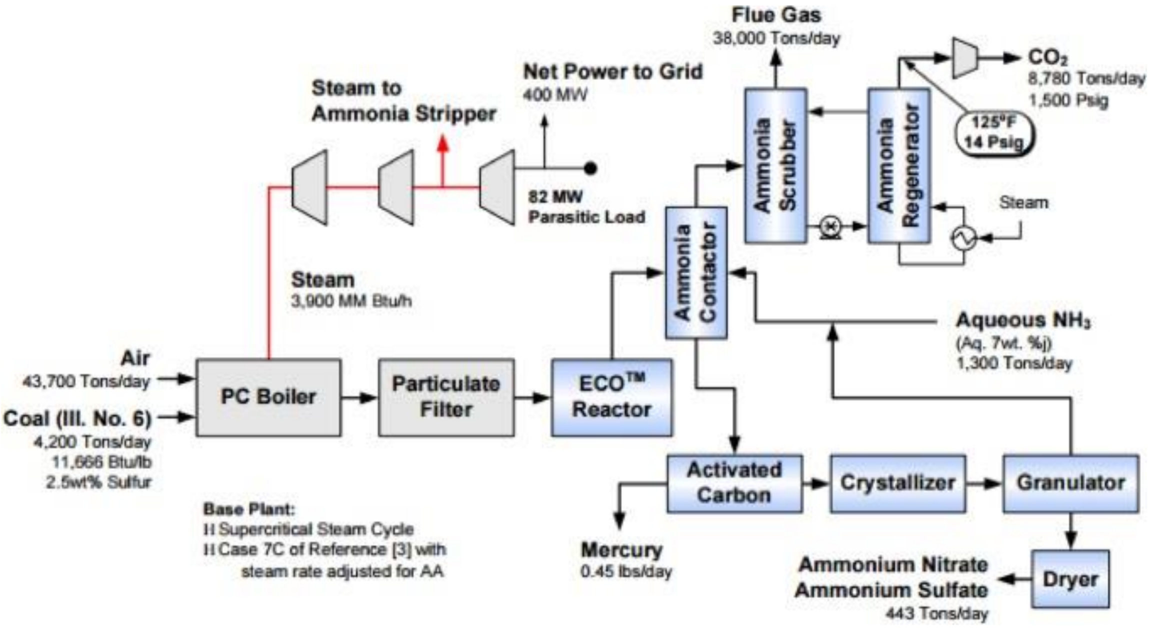


Figure 3 Power plant using ammonia as energy carrier

CHAPTER 2

LITERATURE REVIEW

Ammonia is made of two elements nitrogen and hydrogen, formed from a gas phase mixture of the aforementioned in a ratio of almost 1 to 3, with the formula NH_3. It is a stationary binary hydride, and the most simplistic pnictogen hydride, Ammonia has no color of its gas and a pungent odour. Either advertently or indirectly, ammonia considered as the basic component material for the integration of many pharmaceutical goods.

Before the Haber-Bosch Process, NH_3 was accumulated from the process of dry distillation of nitrogen compound vegetable. Reduction of nitrous acid is done by distillation or nitrites with hydrogen. It was also obtained when coal was distilled or when ammonium salts were decomposed by alkaline hydroxides.

2.1 Haber Bosch Process:

Following are some of the typical processes used in the manufacture of NH_3 by Haber Bosch process; steam reforming in primary reformer, air in secondary reformer, Water gas shift, removal of carbon oxides, syntheses of Ammonia and finally discarding of un-necessory gases. This process is even more cheap, in terms of consumption of energy when excess air in the secondary reforming is rescued and when the reforming gas is treated with different hydrogen rich steam and returns that steam to the synthesis gas. After carbon monoxide is removed, the reformer product gas is then compressed by, at the least, 50 per cent, if no more. The same process consequently can also occur at a plant which has a gas circulator, even if the plant does not have an air compressor. The synthesis gas is kept at a pressure of about 40 to 80 bar abs. The temperature of about range 350 to 430 C is usually provided or maintained at the outlet, in a presence of catalyst consisting of cobalt, iron and other reaction promoting compounds. From other stream of the reformer product gas, the hydrogen-rich stream is recovered.

2.2 Ammonia Production Using Naphtha:

Ammonia is produced in plants, which can be used as feed for Naphtha acts as a production of hydrogen, while the required amount of N_2 is extracted from the air. Naphtha feed, which tends to have a higher carbon to hydrogen ratio (CHR), when is worked with, the generated CO_2 tends to be excessive. In a R-LNG the CO_2, that is generated, is quite less, naptha has higher CRH then ammonia. The ammonia production is a process which can be divided into 2 sections: Syngas Production (preparation of feedstock) and Synthesis of ammonia.

2.2.1 Syngas Production

This part of the process includes desulphurisation, steam reforming which is done in primary reformer, air reforming (secondary reforming), methanation, water gas shift reactions of carbon monoxide, and removal of carbon dioxide.

The process of desulphurisation is then further divided into 2 parts: pre-desulphurisation, or PDS, and final desulphurization, or FDS. In Pre-desulphurisation, sulphur compounds, which are organic removal is done by converting them to hydrogen sulphides, with the use of a catalyst. This then is separated from the inlet by distillation.

2.2.2 Ammonia Synthesis

In this process the nitrogen and hydrogen reacts in a ratio of 1 and 3 to form ammonia, referred to as synthesis gas or syngas. For that, the reaction of reformation is split into 2 sections: primary reformer and secondary reformer.

In primary reforming, where there is steam, hydrocarbon reforming takes place. This is a controlled reaction, in order to maintain a percentage of 10.5% of the gas which is present in product stream, that is actually needed to perform overall air reforming which is done in secondary reformer. But when using natural gas, the gas now sulphur free, is fed to steam and endothermic reaction in nature takes place and metallic nickel is used as a catalyst in the primary reformer.

In secondary reformer, the nitrogen, by air reforming, is obtained for ammonia synthesis. By combustion of molecules of the mixture of gases in air, heat is supplied here. The gas being burnt then provides the heat which is then used for the reforming process or what is left of it. Whereas after the reforming section is done, the product stream consists of H_2, N_2, CO, CO_2, and 0.3 mol% of unreacted hydrocarbons. And Since only pure nitrogen and pure hydrogen gases are needed for ammonia synthesis, any carbon oxide is removed, as it is also necessary, from the raw Syngas. For the secondary reformer using natural gas, the processed gas exits the primary reformer and then enters a secondary reformer. Compressed and processed air from the atmosphere is then heated in the convection zone. An air gas reaction then occurs at the upper available space of the secondary reformer. The reformer output stream may contain carbon monoxide and some varying amounts of carbon Dioxide, but that amount is dependent upon the process technology and feedstock. Carbon oxides are removed from the equation using shift converters and Carbon Dioxide absorbers.

In a Water gas shift converter, the CO is the reducing agent for water to produce hydrogen and carbon dioxide gases. The carbon Monoxide is thus converted promptly into removable carbon dioxide. The carbon Dioxide is then removed by absorbing in 0.31 mol % MDEA (methyl di-ethanol-amine.) Even after a shift reaction and the removal of carbon dioxide from the stream, the output stream still consists of some traces of carbon monoxide and carbon dioxide. Whereas, when it comes to using natural gas, the gases that are exiting the secondary reformer after the reaction are extremely hot. The heat that is obtained from these gases is then utilised to produce a high-pressure stream. This high-pressure stream then is used to run a high-pressure turbine. Here the process of the shift converter can be divided further into 2 parts: High temperature shift converter and low temperature shift converter.

In a high temperature shift converter, the gases from the secondary reformer are cooled down by passing them through a waste heat reboiler. The reformed gas from the secondary converter now contains some carbon monoxide. The carbon monoxide is then converted into carbon dioxide by oxidising it by reacting carbon monoxide with steam. Iron oxide is already present in the gas and is used as a catalyst here to speed and help the oxidation reaction. On the other hand, in a low temperature shift converter, the gas is cooled down by passing it through a waste heat re-boiler. The gas, which has now been cooled, is fed to an LTS and here it reacts with steam. Copper (II) Oxide is used as a catalyst and passing the gas through the LTS helps to dilute the concentration of carbon monoxide in the gas. The gas still contains carbon dioxide and carbon monoxide. Carbon dioxide is removed from the gas with a process known as "Benfield System." During this process, Carbon dioxide is absorbed by the Benfield solution, which consists of hot potassium carbonate K_2CO_3, potassium bicarbonate $KHCO_3$, diethanolamine (DEA) as a promoter and potassium metavanadate KVO_3 as a corrosion inhibitor. Packed bed absorbers containing Pall Rings absorb carbon dioxide. The gas injected from the bottom is then sprayed with the Benfield solution from the top. The solution then flows down and absorbs carbon dioxide from the gas and is then called Rich carbon dioxide solution. This solution is then passed onto the stripping tower from an outlet at the bottom and is regenerated into Benfield solution again. Of all the processes involved that tend to remove carbon oxides from the stream, methanation might be the simplest one, but that still does not mean it is easy. This method is used to lower the carbon dioxide concentration in the output stream to below 10 ppm. The stream outputting after methanation has a composition of nitrogen gas to hydrogen gas of 1 to 3 and is referred to as syngas. Even for the methanation of the stream, natural gas is required as that stream too has carbon monoxide and carbon dioxide. Here carbon Monoxide and carbon Dioxide are converted into methane gas. This is a reverse process to hydrogenation reaction and methane is not a catalyst poisoning agent.

2.3 Alternatives to Haber-Bosch process:

Ammonia has a great importance in production of hydrogen in future as it is a good source of hydrogen. Even though ammonia, for a large percentage of global production, is used for the production of fertilisers, it has also been used as a fuel for space heating and in transport vehicles. Ammonia is produced through Haber-Bosch process, a well-known and a very energy and capital taking process at the present, and is considered to be an good energy storage source, moreover, the infrastructure for its transport is readily available and its distribution is also readily available in many countries throughout the world.

There are many new process which are under way, in order to search for the more efficient and cheap processes of excellent ammonia production growth forecast. The electrochemical tracks, among these methods, have the potential to extensively decrease the energy feed. It also has the potential to easy operate the reactor plans and to decrease the difficulty and expenses of the balance of plants as contrasted to the traditional ammonia generation techniques.

2.3.1 R-LNG (as compared to other fuels):

Process of Naphtha reforming in presence of steam and usage of Naphtha as a fuel, the ammonia plant was designed. Whereas another type of plant of ammonia is being designed for natural gas as the fuel. The plant allows the usage of R-LNG feed, Naphtha feed or mixed feed. The R-LNG can also be used as a fuel for the feed re-heaters or the primary reformer. Although, in order for the R-LNG to be used, new and additional equipment is needed to be incorporated. Some alterations or modifications are also required to some of the processes or process parameters in order to accommodate the overall changes in the whole process and overall reforming process and reaction. The industry is looking to make a switch from Naphtha or natural gas to R-LNG when it comes to the ammonia production. This is a plausible effort by the industry folks to utilise cleaner and cheaper feed stocks or fuel. This helps the industry to achieve a reduction in consumption of energy in the syngas unit varily. Using a Life Cycle Assessment or an LCA, different methods for the production of ammonia can be compared and evaluated. The first ammonia production system consists of an electrolyser for the production of hydrogen gas and a Haber-Bosch plant for the process of ammonia synthesis. Many resources can be utilised to fulfil the energy requirements of the system, hydropower sector, nuclear power, biomass, or waste of munciple. The LCA, or the check the cycle of life, methodology is utilised to identify the quality of environmental effects in increasing earth temperature potential, abiotic depletion, or toxicity of human. Each type of category is assessed during the LCA for each method. Whilst the unfamiliar production of ammonia options are carried competitively in terms of environmental effects, and energy and energy efficiency, both are treated as important criteria for practical application of ammonia production. It can be inferred from the check the life cycle, or the LCA results that municipal waste incineration plants, or the hydropower-based ammonia production methods have the least environmental impact as compared to other considered methods. The greenhouse gas emission which prediction is done already, or the GHG, is calculated for carbon dioxide emission per production of kg of ammonia. The index of sustainability values are also calculated according to the comparatively evaluated energy and energy efficiencies of the system, which also serves as an indicator of any potential improvement.

Using glucose as a substrate, the production of bio-hydrogen was investigated for the ammonia inhibition-I batch and continuous flow reactors. The rate of production of bio-hydrogen was highly dependent on the pH and the concentration of ammonia. In batch tests, the yield of hydrogen remained relatively constant, but in continuous flow tests, yields of hydrogen and rate of production of hydrogen, both were adversely affected by ammonia.

There has been a proposal of using hydrogen as a source of fuel cells thanks to the catalytic decomposition of ammonia. However, the catalyst required is ruthenium, which is quite costly. Even though there are many connections for the catalyst activity that imply that there are many potential and useful alternatives, but the particular candidate differs.

Changes to Traditional Plant to Accommodate the Usage of R-LNG

LNG Preheater

In order to preheat the natural gas a heater is required. Cylindrical draft type heaters are being used by using fuel which is natural gas.

New Sulphur Absorbers

The PDS, or pre-desulphurisation, is eliminated from the overall process as the sulphur content in the feed stream is low. Only the FDS, or the final desulphurisation, section remains a part of the process in the flow line. In a single sulphur absorber, the accessible volume of sulphur absorption catalyst is not enough to give a consistent catalyst continuance. A new sulphur absorber is placed upstream, as it is necessary, and as such the existing comes at the downstream position. Here, the absorber tends to act as a backup absorber in a case when the new absorber is removed or is not in service for catalyst replacement, or if sulphur breakthroughs from the new absorber. Now, desulphurisation is referred to as hydro-desulfurization and now consists of 2 processes; first, the organic contents of sulphur are being converted into hydrogen sulphide in a hydrogenator and then the hydrogen sulphide is removed by absorbing it in the sulphur absorbers. Each has its own catalyst bed consisting of a high-density zinc oxide as the catalyst. The benefit of uniquely tailored fluid units, its robustness and dependability are the working temperature.

Modification & Rerouting of Piping around HDS

After the new reactor is installed, some alterations or modifications are required. An additional upstream inlet is required for water for the existing hydrogenation reactor, as R-LNG is being used as feedstock, for suppression of all reactions taking place in reactor that involves hydrocracking. It is because there is no moisture in R-LNG.

Reformer Burners

The traditional ammonia production plants use primary reformers designed for the vaporised Naphtha combustion along with off gasses that ae present in nozzle, but volumetric flow of the fuel is increased when the feedstock is switched to R-LNG from commonly used Naptha, and there is need to install new burners that have dual nozzel. Moreover, fuel headers size is not sufficient for the massive volumetric flow in case of R-LNG feedstock. If the current fuel headers are to be used, this can result in maldistribution of the fuel present in burner as there will be a great drop in the pressure. A fuel header is needed, suitable or compatible with the usage for R-LNG and the fuel headers can be used for off gas combustion as well as for Naptha.

Fuel Preheater for LNG

Normally supply of Regasified Natural Gas is at a battery limit, the depressurisation of the pressure of fuel gas will result in a notable decrease in temperature. There is will be suitable to install a new fuel preheater

Mixing Point of Feed Stream

As plant has capacity to continues its operation both for LNG as well as Naptha, it is mandatory to have an accurate method for measuring amount of carbon present in feed so control of steam to carbon ratio will become easy. A vortex flow meter is installed for this very purpose along with flow orifice already present in the bypass over the existing feed control valve. On basis of large range of flows measurements, determination of feed's molecular weight has become easy and straightforward.

2.4 Carbon Deposition in Steam Reforming and Methanation

2.4.1 Structure of Carbon & Coke

Carbon is a outlet of carbon monoxide incongruity while coking is manufactured from breakdown or reduction of feed on metals. Actually the coke forms, nevertheless, varies from high molecular weight hydrocarbons to carbon, depends upon the condition under which the coking started and forged or carbon or coke species are seen in reforming with steam; whisker like carbon formed at higher temperatures and including hydrocarbon films made by polymerization or pyrolytic carbon from cracking of hydrocarbons. Recent studies of the a very good nickel having a bonding and structure of carbon on it and covers by diffraction (LEED) and Auger electron spectroscopy (AES) confirm the presence of "dispersed" or atomic carbon stable and polymerized carbon stable.

2.4.2 Removal

One of the major and severe problems in conventional steam reforming using R-LNG is the formation of coke. The usual molar ratio of steam to methane that is done in steam reformer or the initial composition, with or feed is with oxygen and it is well distributed in the tubes of primary reformer reactor. It is quite clear that adding the oxygen will help move the system far from the boundary because O_2 aids to prevent the formation of the coke deposits inside the tubes of primary reformer, as its importantly know all the process take place under catalyst. Extracting hydrogen gas leads to the travel of the reactant side operation point along a line directed away. Extraction of hydrogen might very well, by permeation, not lead any greater chances of deposition of carbon may actually reduce the formation of the coke. This is a very crucial search with practical demonstration, and it is also notable that there have not been any idea of for formation of coke in the FDMR reactor, as usually hydrogen removal, like decreasing the amount of water will also increase the formation of carbon on tubes of reformer. Although it can reasonably be inferred that there is still a great need for confirmation that if it will be this way for the longer duration of the processes or operations that the approximated hours of the experimentatio

CHAPTER 3

AMMONIA PRODUCTION PROCESS AT FFC

In Pakistan, Ammonia preparation plants extract H_2 from a hydrocarbon in the form of natural gas. And as with all Ammonia preparation plants, N_2 is distilled from atmospheric air. Some of the processes of utmost importance, when it comes to preparation of Ammonia, are mentioned below:

Hydrogen (H_2) and Nitrogen (N_2) are mixed, the resultant compound is called Ammonia. The H_2 to N_2 ratio is 3 to 1, respectively. The mixture might also contain, to a limited degree, some inert gases, i.e. Argon or Methane.

In the desulphurization section, the process gas is utterly desulphurized.

The Hydrocarbon, now desulphurized, is now converted to raw synthesis gas, collectively with steam and air. This gas contains Hydrogen, Nitrogen, Carbon Dioxide, and Carbon Monoxide. This reformation happens at approx. 30 Kg/cm^2.

During Gas Purification, Carbon Monoxide is converted into Carbon Dioxide and Hydrogen, and then Carbon Dioxide is removed by absorbing it into The Benfield Solution.

In Ammonia Synthesis, using a catalytic reaction, the purified synthesis gas is then converted into Ammonia, after being compressed to a pressure of 260 Kg/cm^2.

3.1 Desulphurization:

The natural gas that is supplied to the plant, is kept at a pressure between 38-45 Kg/cm^2, as a pressure of approx. 40 Kg/cm^2 is required for the process. The natural gas is then heated in a preheater and passed through a de-sulphizer, where the sulphur in the natural gas is removed. But it is normally bypassed since no sulphur is present in current feed.

3.2 Primary Steam Reforming

After then the feed which is sulphur-free gas with steam. The following is the resultant reaction of Methane and Steam.

Steam reforming reaction.

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

$$CO + H_2O \rightarrow CO2 + H_2$$

This highly endothermic reaction contains metallic Nickel as the catalyst in the primary reformer.

3.3 Secondary Reformer:

Processed gas, from the primary reformer, now enters a secondary reformer. In the convection zone, compressed, processed air from the atmosphere is heated.

In the upper, available space of secondary reformer, an air gas reaction occurs. There is a High gas temperature at the bottom and front of the catalyst bed, because of the combustion of air. 950 ^oC is the approx. exit temperature, whereas the reaction, in contact with the catalyst, will occur at approx. 1100-1200 ^oC.

3.4 Shift Conversion:

The gases, exiting secondary reformer, are extremely hot and tend to have a temperature of more than 1000 ^oC. The heat from these gases is then utilized in the production of high-pressure steam, which tends to be of approx. 34 Kg/cm^2 pressure. This high-pressure steam is then used to run a high-pressure turbine.

The process of shift conversion is divided into two steps.

3.4.1 High-Temperature Shift Converter:

Hot gases, from the secondary reformer, are then cooled down to a temperature of 336 ^oC, after passing them through a waste heat re-boiler. The now reformed gas, from the secondary reformer, contains about 11.9% CO. This CO is oxidized into CO_2 by reaction with steam, using Iron Oxide as the catalyst, that is already present in the gas.

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

3.4.2 Low-Temperature Shift Converter:

The gas, with the temperature of up to 430 ^oC, is cooled to a temperature of 203 ^oC by passing it through waste heat re-boiler. Now cooled gas is fed to LTS, where it again reacts with steam and CuO is used as the catalyst. After passing through LTS, CO's concentration is lessened to approx. 0.18%.

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

3.5 Carbon Dioxide Removal:

The gas now contains 18.5% CO_2 and 0.18% CO. The CO_2 is removed from the gas by a process called "Benfield System."

During this process, CO_2 is absorbed by the Benfield Solution, that is the Pot. Carbonate Solution.

Packed bed absorbers, containing Pall Rings, absorb CO_2. Gas injected from the bottom is sprayed with The Benfield Solution, from the top. The solution flows down and is called Rich CO_2 solution.

The gas, consisting of 0.1% CO_2, is released from the top, and the CO_2 Rich solution is passed on to the stripping tower, from the bottom, where it is regenerated into Benfield Solution again.

3.5.1 Regeneration of the Benfield Solution:

Regeneration of The Benfield Solution occurs in the stripper, in which, CO_2 is divested from the Rich CO_2 Benfield Solution.

The operation conditions for stripper are:

1. High temperature
2. Low pressure

The solution enters from the top of the stripper and passes through many plates and sieves. The stripper operates at a low and high temperature. The CO_2-free solution is then sent back to the absorber from the bottom. LS provides the heat in the stripper. The excess CO_2 is then transported to a Urea Preparation Plant.

3.6 Methanator:

The gas still consisting of CO and CO_2. CO and CO_2 are then converted into Methane. The process is an inverse of a reforming process and is an exothermic hydrogenation reaction. Methane is not a catalyst poisoning agent.

3.7 Ammonia Synthesis & Refrigeration:

The synthesised gas, after being compressed to a pressure of approx. 250 Kg/cm$_2$, is passed to an axial radial flow converter at a temperature of about 380-520 ^oC. The gas is passed over Iron which acts as a catalyst, and about 33% of the gas is converted into Ammonia. The reaction effluent is cooled Ammonia, which liquified and separated. The un-reacted gas is recycled.

Main steps involved are:

1. Synthesis gas compression and recycle of unreacted gas.
2. Separation of ammonia and refrigeration.
3. Ammonia storage.

3.8 Compression of Synthesized Gas & Recycling of Unreacted Gas:

Synthesized gas is compressed from a pressure of 26 Kg/cm^2 to a pressure of 250 Kg/cm^2, with a 4-case centrifugal compressor. The last casing contains a re-circulator, with an impeller and adjustable inlet guide vanes, and re-circulator raises the pressure of the recycled gas, to the extent of pressure drop in the synthesis loop. A set of 2 turbines runs the compressor, extracting and condensing the steam. These turbines are provided with 100 Kg/cm^2 steam, produced in The Ammonia Plant. Steam is extracted at approx. 37 Kg/cm^2 and is used for the turbines.

3.9 Synthesis Loop Cooling & Purging:

A synthesis loop consists of an axial radial converter. The gas flows through 3 catalyst beds, in the inward direction.

The gas is inserted in the loop, after compression, in between 2 Ammonia chillers. This removes the minute traces of CO_2 and H_2O vapours in the methanator. Failure to remove these can result in the formation of Carbamate, which is corrosive to the catalyst. In exchanger, the mixture is chilled, and Ammonia is liquified and extracted in an Ammonia Separator. The gas, from the separator, is heated in the exchanger. The gas's pressure is raised by passing it through re-circulator. The gas is further heated in an exchanger tube side and admitted to the converter.

3.10 Separation & Refrigeration of Ammonia:

Converter effluent, consisting of Synthesized Ammonia, unreacted gas (after cooling), is purged and mixed with make-up synthesis gas and is further cooled in the II chiller. Liquid Ammonia is separated in a separator. Liquid Ammonia is then inserted, from a separator, in a down Vessel, and its pressure is reduced to approx. 25 Kg/cm^2. Ammonia is then supplied to a Urea Plant directly from this vessel. The excessive liquid Ammonia, from Knock-out drums, is depressurized to 3.5 Kg/cm^2 and is sent to Ammonia Storage. Some Ammonia from that storage is recycled for refrigeration loop,that consists of a compressor, condensers, and a Knock-out vessel.

3.11 Process Flow Diagram (PFD):

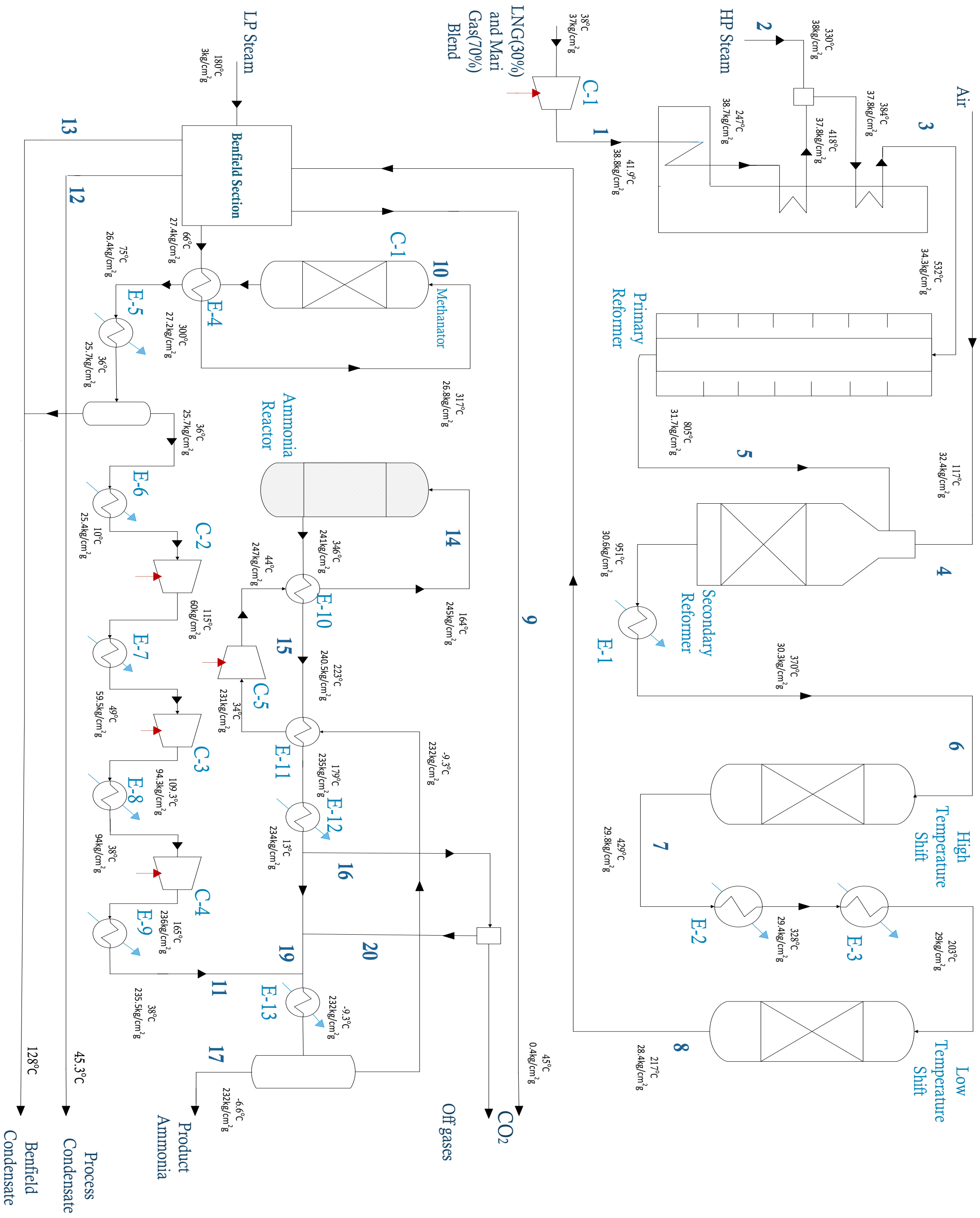


Figure 4 Process Flow Diagram for ammonia production

CHAPTER 4

MATERIAL BALANCE

In this chapter, we did material balance to find out the impact of new feed on the plant. The manual calculations were done by hand and also on excel sheet for further improvements. Material balance on reactors and separation columns and Benfield section was applied as they are the main units installed.

1 kgmol/day was taken as the basis of the material balance and the plant was assumed to be at steady state, implying there is no accumulation. Process flow diagrams are also been added for reference and to understand process. Furthermore, equations and formulas are given where required and tables are been formed to show input and output streams of raw materials and products.

Material Balance is important because it tells us about flow rates and how the equipment is handling and doing the required job. Usually we apply law of conservation of mass for material balance that says that mass can neither be created nor destroyed.

4.1 Material balance on Reactors:

To apply material balance on the reactors following steps were followed:

1. The material balance was applied on basis of one hour.
2. Steady state process was assumed meaning no accumulation.
   In – out = Generation - Consumption + Accumulation
   Applying steady state process condition (Accumulation=0)
   In – out = Generation - Consumption
3. All the units were converted into kgmol
4. Conversions of each reactor occurring in the reactor were calculated
   $$\text{Conversion (\%)} = \frac{\text{Moles in feed stream} - \text{Moles in product stream}}{\text{Moles in feed stream}} * 100$$
5. These conversions were then used to find the moles of reactant of each reaction.

4.1.1 Primary Reformer:

Following are the reactions and their conversions that occur in primary reformer:

$CH_4 + H_2O \rightarrow CO + 3H_2$ Conversion=72%

$CO + H_2O \rightarrow CO_2 + H_2$ Conversion=48%

$C_2H_6 + 4H_2O \rightarrow 2CO_2 + 7H_2$ Conversion=100%

$C_2H_8 + 6H_2O \rightarrow 3CO_2 + 10H_2$ Conversion=100%

$C_4H_{10} + 8H_2O \rightarrow 4CO_2 + 13H_2$ Conversion=100%

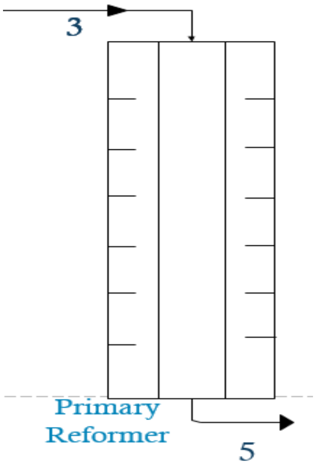


Figure 5 Primary Reformer

Primary Reformer (PR)		
Components	**In**	**Out (5)**
	(3)	
Nitrogen	235.85	235.87
Carbon dioxide	104.52	747.98
Methane	1604.68	449.32
Ethane	29.61	0
Propane	6.50	0
Butane	2.47	0
Water	5775	3887.72
Hydrogen	0	4324.45
Carbon monoxide	0	600.8
Total moles	**7759**	**10246.1**

Table 5 Primary reformer material balance

4.1.2 Secondary Reformer:

$2H_2 + O_2 \rightarrow 2H_2O$ Conversion=100%

$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$ Conversion=100%

$2CO + O_2 \rightarrow 2CO_2$ Conversion=100%

$CH_4 + H_2O \rightarrow CO + 3H_2$ Conversion=90%

$CO_2 + H_2 \rightarrow CO + H_2O$ Conversion=100%

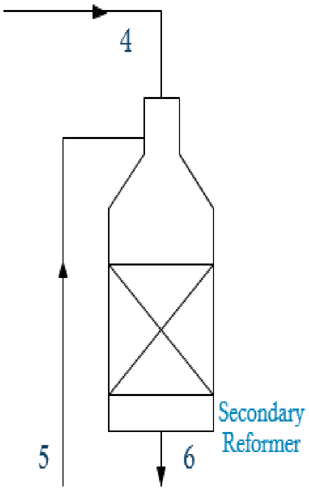


Figure 6 Secondary Reformer

Secondary Reformer (SR)			
Components	**In**		**Out (6)**
	PR Outlet (5)	**Air (4)**	
Nitrogen	235.87	1229.85	1465.7
Carbon dioxide	747.98	0.49	747.98
Methane	449.32	0	36.64
Water	3887.72	7.827	4144.6
Hydrogen	4324.45	0	4900.8
Carbon monoxide	600.8	0	1013.45
Oxygen	0	330.8	0
Argon	0	14.85	14.85
Total moles	**10246.4**	**1583.8**	**12324.0**

Table 6 Material balance of Secondary Reformer

4.1.3 High Temperature Shift Reactor:

In the high temperature shift reactor CO is converted into CO_2 in the presence of iron oxide catalyst.

$CO + H_2O \rightarrow CO_2 + H_2$ Conversion=48%

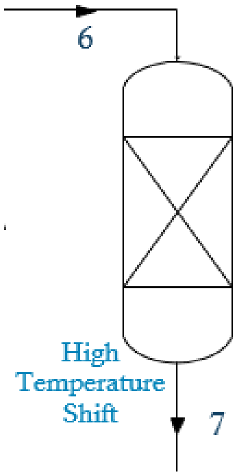


Figure 7 HTS Reactor

High temperature Shift		
Components	**In (6)**	**Out (7)**
	SR Outlet	**To LTS**
Nitrogen	1465.7	1465.7
Carbon dioxide	747.98	1528.33
Methane	36.64	36.64
Water	4144.6	3364.24
Hydrogen	4900.8	5681.15
Carbon monoxide	1013.45	233.093
Argon	14.85	14.85
Total moles	**12324.2**	**12324.1**

Table 7 Material balance of HTS reactor

4.1.4 Low Temperature Shift Reactor:

In the LTS reactor remaining CO is being converted into CO_2 in the presence of CuO catalyst. The material balance of the reactor is as follows:

$$CO + H_2O \rightarrow CO_2 + H_2 \qquad \text{Conversion = 92\%}$$

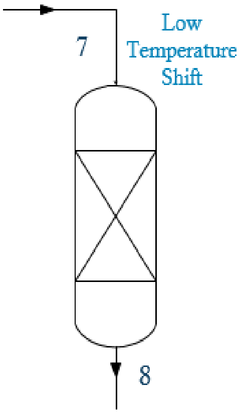


Figure 8 LTS Reactor

Low Temperature Shift		
Components	**In**	**Out**
Nitrogen	1465.7	1465.7
Carbon dioxide	1528.33	1742.78
Methane	36.64	36.64
Water	3364.24	3149.79
Hydrogen	5681.15	5895.60
Carbon monoxide	233.09	18.64
Argon	14.85	14.85
Total moles	**12324.02**	**12324.01**

Table 8 Material balance of LTS

4.1.5 Methanator:

In the Methanator, CO and CO_2 are removed from the process stream by converting them into methane and water.

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O \qquad \text{Conversion=100\%}$$

$$CO + 3H_2 \rightarrow CH_4 + H_2O \qquad \text{Conversion=100\%}$$

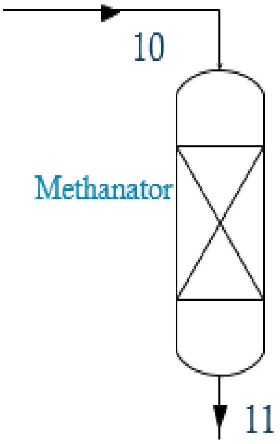


Figure 9 Methanator Reactor

Methanator		
Components	**IN (10)**	**OUT (11)**
Nitrogen	1425.902	1425.90
Carbon dioxide	7.413	0
Methane	35.76	61.38
Water	41.49	74.5
Hydrogen	5744.6	5660.3
Carbon monoxide	18.207	0
Argon	12.903	12.90
Total moles	**7286.3**	**7235.1**

Table 9 Material balance of methanator

4.1.6 Ammonia Reactor:

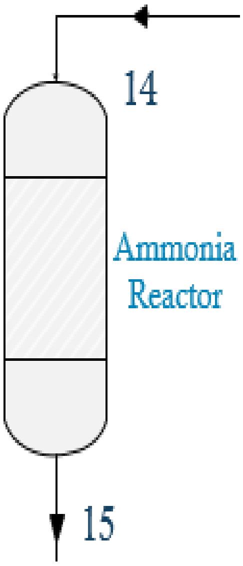


Figure 10 Ammonia Reactor

Ammonia Reactor		
Components	**IN (14)**	**OUT (15)**
Nitrogen	4645.45	3275.04
Methane	3218.55	3218.55
Hydrogen	102651.1	98539.92
Ammonia	3413.597	6154.418
Argon	625.55	625.55
Total moles	**114554.3**	**207078.3**

Table 10 Material balance of ammonia reactor

$$N_2 + 3H_2 \rightarrow 2NH_3 + N_2 \text{ Conversion=29.3\%}$$

4.2 Material balance on Benfield Section:

Material balance of the Benfield section, in which CO_2 is stripped off from the process stream before it moves to the ammonia reactor, is as follows:

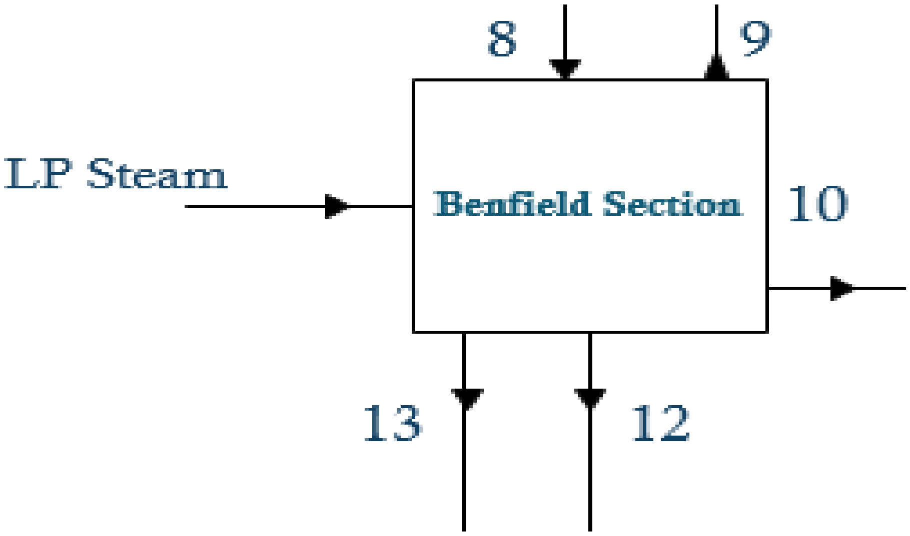


Figure 11 Benfield Section

Benfield Section						
Components	**In**		**Out**			
	From	**Steam**	**CO2 to Urea**	**To Methanator**	**Benfield Condensate**	**Condensate**
Nitrogen	1465.7	0	4.67	1460.86	0.164	0
Carbon dioxide	1742.78	0	1731.7	7.595	3.4856	0
Methane	36.64	0	0	36.64	0	0
Water	3149.79	691	95.214	42.517	715.81	2987.228
Hydrogen	5895.6	0	9.216	5885.497	0.9778	0
Carbon monoxide	18.647	0	0	18.647	0	0
Argon	14.85	0	1.6305	13.219	0	0
Total	**12324**	**691**	**1842.4305**	**7464.975**	**720.4374**	**2987.228**

Table 11 Material balance of Benfield Section

4.3 Material balance on Separator:

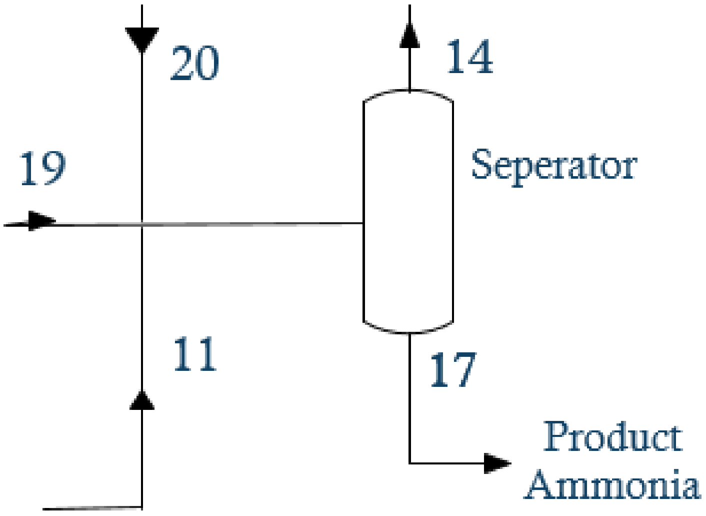


Figure 12 Separator

Separator					
Components	**IN**			**Out**	
	11	**20**	**19**	**14**	**17**
Nitrogen	1425.9	0.0072	3220.74	4645.45	1.17
Methane	61.38	0.1198	3170.63	3218.55	13.57
hydrogen	5660.4	0.3568	97032.34	102651.1	41.93
argon	12.9	0.0376	616.88	625.55	4.27
Ammonia	0	21.4122	6056.38	3413.59	2664.5
Total moles	**7160.6**	**21.9**	**110096.9**	**114554.3**	**2725.4**

Table 12 Material balance of separator

CHAPTER 5

ENERGY BALANCE

In this chapter, energy balance of on the ammonia plant using mixed feed of LNG and Mari gas was applied. The manual calculations were done by hand and also on excel sheet for further improvements. Energy balance on furnace, reactor, compressors, and multiple heat exchangers was applied as they are the main units impacted by the changed feed.

The main equation used in this balance was the law of conservation of energy i.e. energy of input streams will be equal to outlet streams if there is no accumulation or generation of energy. Equations and formulas are given where required and tables are been formed to show input and output streams of raw materials and products. Energy balance is very important because it provides information about heat input, output, accumulation, and temperatures of streams etc.

Values of different factors such as specific heats were either taken from Perry's chemical engineering or by using Aspen HYSYS (As it gives values of C_p, C_v, average temperatures etc.). Temperatures of different streams were taken from the literature review or from different research papers.

5.1 Assumptions and formulas used in energy balance:

Following assumptions were made for the energy balance of the plant:

1. The process is considered as steady state process which implies:

$$\Delta U + \Delta E_k + \Delta E_p = Q - W_s + W_{fl}$$

$$\Delta E_k = \Delta E_p = 0$$

$$\Delta H = Q + W_{fl}$$

2. For energy balance of a component of a plant involving a reaction following equation is used:

$$\Delta H_{r,t} = \Delta H_P - \Delta H_R + \Delta H_{reaction} \text{ (kJ/hr)}$$

3. Enthalpy of each stream was calculated using following formula:

$$H_T = \int_{Td}^{T} Cp * dT \qquad \text{(kJ /hr)}$$

4. To calculate C_p of a mixed stream, containing more than one chemical species following formula was used:

$$Cp\ (mixture) = x_a\, C_pa + x_b\, C_pb + x_c\, C_pc + \ldots\ldots\ldots$$

5. Calculation for the power of the compressor was done by calculating the work done by the compressor and then dividing it by isentropic efficiency.

$$P = \frac{P_s}{E} \quad (MW)$$

5.2 Energy Balance on furnace and Reactors:

5.2.1 Furnace:

In furnace the feed gas is pre-heated to achieve around 532 ^{0}C of temperature. This is done in order to achieve better heat efficiency of the process. If the furnace step is skipped more thermal energy will be required in the primary reformer which is quite hard to achieve. The furnace heats the feed gas in three passes rather in a single pass. This is done to make the pre-heating process form effective and efficient.

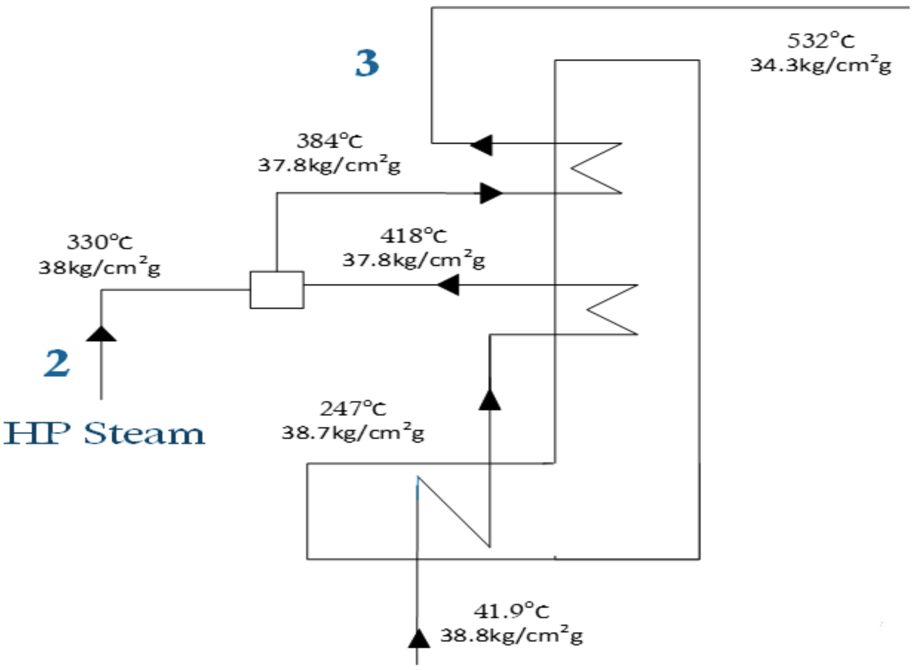


Figure 13 Furnace

Furnace						
Parameters	**First Pass**		**Second Pass**		**Third Pass**	
	In (1)	**Out (1)**	**In (1)**	**Out (1)**	**In (3)**	**Out (3)**
Temperature (C)	41.9	247	247	418	384	532
Pressure(Kg/cm^2)	38.8	38.7	38.7	37.8	37.8	34.3
Cp (KJ/mole*^{o}C)	39.86	46.98	46.98	55.48	43.53	45.51
Moles	1984	1984	1984	1984	7759	7759
Enthalpy(kJ/hr)	3.30E+06	2.30E+07	2.30E+07	4.60E+07	1.30E+08	1.80E+08
Q (KJ/hr)	**1.98E+07**		**2.20E+07**		**5.70E+07**	

Table 13 Energy balance on Furnace

5.2.2 Primary Reformer:

The energy balance of the primary reformer is as follows:

Enthalpy = H = m C_p dT

H_{in} = 7759 * 45.51 * (372)

H_{in} = 1.80 * 10^8 kJ/hr

Similarly, the enthalpy of the outlet stream is calculated, and it comes to be:

H_{out} = 3.20 * 10^8 kJ /hr

Heat of reaction is calculated by combining the all reactions taking place in primary reformer as mentioned in material balance.

Heat of reaction = 262237.8 kJ /hr

$\Delta H_{r,t} = \Delta H_P - \Delta H_R + \Delta H_{reaction}$

$\Delta H_{r,t} = 1.4 * 10^8$ kJ/hr

As flow work is zero in this case of reactor hence Q = $1.4 * 10^8$ kJ /hr.

Primary Reformer		
Parameters	**In (3)**	**Out (5)**
Temperature (^{o}C)	532	805
Pressure (Kg/cm^{2})	34.3	31.7
Cp (KJ/mole*^{o}C)	45.51	39.08
Moles	7759	10246
Enthalpy(kJ/hr)	1.80E+08	3.20E+08
Heat of Reaction	262237.8	
Q(KJ/hr)	**1.40E+08**	

Table 14 Energy balance of Primary reformer

5.2.3 Secondary Reformer:

Similar calculations for energy balance are done in secondary reformer as were done on the primary reformer. Following is the table of energy balance calculations of secondary reformer:

Secondary Reformer			
Parameters	**In**		**Out**
	5	**4**	**6**
Temperature (^{o}C)	805	117	951
Pressure (Kg/cm^{2})	31.7	32.4	30.6
Cp (KJ/mole*^{o}C)	39.18	30.69	37.73
Moles	1.025e00+4	1584	1.23E+04
Enthalpy(kJ/hr)	8.60E+06		4.40E+08
Heat of Reaction	-1.00E+05		
Q(KJ/hr)	**4.30E+08**		

Table 15 Energy balance of Secondary reformer

5.2.4 High Temperature Shift Reactor:

The energy balance of the HTS is as follows:

High Temperature Shift (HTS)		
Parameters	**In (6)**	**Out (7)**
Temperature (°C)	370	429
Pressure (Kg/cm^2)	30.3	29.8
Cp (KJ/mole*°C)	34	34.86
Moles	1.23E+04	
Enthalpy(kJ/hr)	1.50E+08	1.80E+08
Heat of Reaction	-3.24E+04	
Q (KJ/hr)	**2.9e00+7**	

Table 16 HTS energy balance

5.2.5 Low Temperature Shift Reactor:

Following is the summary of energy balance of LTS.

Low Temperature Shift (LTS)		
Parameters	**In**	**Out**
Temperature (°C)	203	217
Pressure (Kg/cm^2)	29	28.4
Cp(KJ/mole*°C)	33.61	33.7
Heat of Reaction	-8.90E+03	
Moles	1.23E+04	
Enthalpy(kJ/hr)	8.40E+07	9.00E+07
Q (KJ/hr)	**6.10E+06**	

Table 17 LTS energy balance

5.2.6 Methanator:

Energy balance on the methanator is as follow

Methanator		
Parameters	**In**	**Out**
Temperature (^{o}C)	300	317
Pressure (Kg/cm^2)	27.2	26.8
Cp (KJ/mole*^{o}C)	29.65	29.8
Moles	7465	7465
Enthalpy(kJ/hr)	6.60E+07	6.83E+07
Heat of Reaction	-4977	
Q(KJ/hr)	2.29E+06	

5.2.7 Ammonia reactor:

Ammonia Reactor		
Parameters	**In (14)**	**Out (15)**
Temperature (^{o}C)	164	346
Pressure (Kg/cm^2)	245	241
Cp (KJ/mole*^{o}C)	30.39	31.38
Moles	1.15E+05	111813.5
Enthalpy(kJ/hr)	5.70E+08	1.20E+09
Heat of Reaction	-269144.01	
Q(KJ/hr)	**7.06E+08**	

Table 18 Ammonia reactor energy balance

5.3 Energy balance on Compressors:

5.3.1 Compressor 1:

To calculate power required for the operation of compressor 1 first work done by it was calculated and the dividing it by isentropic efficiency of the compressor.

$$\Delta U + \Delta E_k + \Delta E_p = Q - W_s + W_{fl}$$

$$\Delta E_k = \Delta E_p = 0$$

$$\Delta H = Q + W_{fl}$$

In case of compressors W= ΔH so:

Enthalpy = H = m C_p dT

H_{in} = 1.98 * 10 ^{3} * 39.05 * (41.9-38)

H_{in} = 2.90 * 10 ^{6} kJ/hr

Similarly, the enthalpy of the outlet stream is calculated, and it comes to be:

H_{out} = 3.30 * 10 ^{6} kJ/hr

Power of compressor 1= (4 * 10 ^{5})/ (0.75) = 1.36 MW

C-1		
Parameters	**In**	**Out**
Temperature (^{o}C)	38	41.9
Pressure (Kg/cm^{2})	37	38
Cp (KJ/mole*^{o}C)	39.05	39.89
Moles	1.98E+03	
Enthalpy(kJ/hr)	2.90E+06	3.30E+06
Efficiency	75%	
Power (MW)	1.36	

Table 19 Compressor 1 energy balance

5.3.2 Compressor 2:

C- 2		
Parameters	**In**	**Out**
Temperature (^oC)	10	115.4
Pressure (Kg/cm^2)	25.4	60
Cp(KJ/mole*^oC)	28.95	29.33
Moles	7.16E+03	
Enthalpy(kJ/hr)	2.07E+06	2.40E+07
Efficiency	75%	
Power (MW)	8.48	

Table 20 Compressor 2 energy balance

5.3.3 Compressor 3:

C-3		
		Out
Temperature (^oC)	49	109.3
Pressure (Kg/cm^2)	59.5	94.3
Cp (KJ/mole*^oC)	29.32	29.54
Moles	7.16E+03	
Enthalpy(kJ/hr)	1.03E+07	2.31E+07
Efficiency	75%	
Power (MW)	4.7	

Table 21 Compressor 3 energy balance

5.3.4 Compressor 4:

C- 4		
Parameters	**In**	**Out**
Temperature (°C)	38	165
Pressure (Kg/cm^2)	94	236
Cp (KJ/mole*°C)	29	30.07
Moles	7.16E+03	
Enthalpy (Kj/hr)	8.04E+06	8.04E+06
Efficiency	**75%**	
Power (MW)	**9.9**	

Table 22 Compressor 4 energy balance

5.3.5 Compressor 5:

C- 5		
Parameters	**In**	**Out**
Temperature (°C)	34	44.1
Pressure (Kg/cm^2)	231	247
Cp (KJ/mole*°C)	30.6	30.6
Moles	1.15E+05	
Enthalpy (Kj/hr)	1.19E+08	1.38E+08
Efficiency	75%	
Power (MW)	7.037	

Table 24 Compressor 5

5.4 Energy balance on Heat Exchangers:

E-1		
Tube Side (Gas)		
Parameters	**In**	**Out**
Temperature (^{o}C)	951	370
Pressure(Kg/cm^{2})	30.6	30.3
Cp(KJ/mole*C)	37.73	34.01
Moles	1.23E+04	
Enthalpy	4.40E+08	1.50E+08
Q(KJ/hr)	**-2.87E+08**	
Shell Side (BFW)		
Parameters	**In**	**Out**
TEMP diff	45	
Cp (KJ/mole*C)	75.38	
Moles	8608	
Q(KJ/hr)	**2.90E+08**	

Table 25 E1 energy balance

E-2		
Tube Side(Gas)		
Parameters	**In**	**Out**
Temperature (^{o}C)	429	328
Pressure(Kg/cm^{2})	29.8	29.4
Cp(KJ/mole*C)	34.86	34.21
Moles	1.23E+04	
Enthalpy	1.80E+07	1.38E+08
Q(KJ/hr)	**-4.60E+07**	
Shell Side(BFW)		
Parameters	**In**	**Out**
TEMP diff	45	
Cp (KJ/mole*C)	75.38	
Moles	13584	
Q(KJ/hr)	**4.53E+07**	

Table 26 E2 energy balance

E-3		
Tube Side (Gas)		
Parameters	**In**	**Out**
Temperature (^{o}C)	75	36
Pressure(Kg/cm^{2})	26.4	25.7
Cp(KJ/mole*C)	29.8	29.35
Moles	7.24E+03	
Enthalpy	1.68E+07	7.60E+06
Q(KJ/hr)	**-2.87E+08**	
Shell Side (CW)		
TEMP diff	10	
Cp (KJ/mole*C)	75.38	
Moles	12100	
Q(KJ/hr)	**2.90E+08**	

Table 27 E-3 energy balance

E-4		
Tube Side		
Parameters	**In**	**Out**
Temperature (^{o}C)	66	300
Pressure(Kg/cm^{2})	27.4	27.2
Cp(KJ/mole*C)	29.05	29.65
Moles	7465	
Enthalpy	1.40E+07	6.60E+07
Q(KJ/hr)	**5.20E+07**	
Shell Side		
Temperature (C)	317	75
Pressure(Kg/cm2)	26.8	26.4
Cp(KJ/mole*C)	29.8	29.8
Moles	7235	
Enthalpy	6.83E+07	1.68E+07
Q(KJ/hr)	**-5.17E+07**	

Table 28 E-4 Energy balance

E-5		
Tube Side(Gas)		
Parameters	**In**	**Out**
Temperature (°C)	115.4	49
Pressure(Kg/cm²)	60	59.5
Cp(KJ/mole*C)	29.33	29.32
Moles	7.16E+03	
Enthalpy	2.40E+07	1.03E+07
Q(KJ/hr)	**-1.37E+07**	
Shell Side(BFW)		
TEMP diff	10	
Cp(KJ/mole*C)	75.38	
Moles	18260	
Q(KJ/hr)	**1.35E+07**	

Table 29 E-5 energy balance

E-6		
Tube Side(Gas)		
Parameters	**In**	**Out**
Temperature (°C)	36	10
Pressure(Kg/cm²)	25.7	25.4
Cp(KJ/mole*C)	28.97	28.95
Moles	7.16E+03	
Enthalpy	7.40E+06	2.07E+06
Q(KJ/hr)	**-5.30E+06**	
Shell Side(BFW)		
TEMP diff	39	
Cp(KJ/mole*C)	29.1	
Moles	4611	
Q(KJ/hr)	**5.23E+06**	

Table 30 E-6 energy balance

E-7		
Tube Side(Gas)		
Parameters	**In**	**Out**
Temperature (°C)	109.3	38
Pressure(Kg/cm²)	94.3	94
Cp(KJ/mole*C)	29.54	29.66
Moles	7.16E+03	
Enthalpy	2.31E+07	8.04E+06
Q(KJ/hr)	**-1.40E+07**	
Shell Side(CW)		
TEMP diff	15	
Cp(KJ/mole*C)	75.38	
Moles	12381.2	
Q(KJ/hr)	**1.39E+07**	

Table 31 E-7 energy balance

E-8		
Tube Side(Gas)		
Parameters	**In**	**Out**
Temperature (°C)	165.1	38
Pressure(Kg/cm²)	236	235.5
Cp(KJ/mole*C)	30.07	30.6
Moles	7.16E+03	
Enthalpy	3.50E+07	8.30E+06
Q(KJ/hr)	**-2.66E+07**	
Shell Side (CW)		
TEMP diff	10	
Cp(KJ/mole*C)	75.38	
Moles	35000	
Q(KJ/hr)	**2.63E+07**	

Table 32 E-8 energy balance

E-9		
Shell Side		
Parameters	**In**	**Out**
Temperature (°C)	44.1	164
Pressure(Kg/cm^2)	247	245
Cp(KJ/mole*C)	30.68	30.39
Moles	1.46E+05	
Enthalpy	1.38E+08	5.70E+08
Q(KJ/hr)	**4.30E+08**	
Tube Side		
Temperature (C)	346	223
Pressure(Kg/cm2)	241	240.5
Cp(KJ/mole*C)	31.36	31.36
Moles	1.15E+05	
Enthalpy	1.24E+09	7.83E+08
Q(KJ/hr)	**-4.25E+08**	

Table 33 E-9 energy balance

E-10		
Shell Side		
Parameters	**In**	**Out**
Temperature (°C)	179.65	13
Pressure(Kg/cm^2)	235	234
Cp(KJ/mole*C)	31.36	31.49
Moles	6.29E+08	4.50E+07
Enthalpy	1.12E+05	
Q(KJ/hr)	**-5.80E+08**	
Shell Side (BFW)		
Parameters	**In**	**Out**
TEMP diff	150	
Cp(KJ/mole*C)	75.38	
Moles	50000	
Q(KJ/hr)	**5.75E+08**	

Table 35 E-10 energy balance

E-11		
Tube Side		
Parameters	**In**	**Out**
Temperature (°C)	223	179.65
Pressure(Kg/cm^2)	240.5	235
Cp(KJ/mole*C)	31.36	31.36
Moles	1.18E+05	
Enthalpy	7.83E+08	6.20E+08
Q(KJ/hr)	**-1.50E+08**	
Shell Side		
Temperature (C)	-9.3	34
Pressure(Kg/cm2)	232	231
Cp(KJ/mole*C)	31.11	30.67
Moles	1.15E+05	
Enthalpy	3.30E+07	1.19E+08
Q(KJ/hr)	**-1.52E+08**	

Table 36 E-11 energy balance

E-12		
Tube Side		
Parameters	**In**	**Out**
Temperature (°C)	12.3	-9.3
Pressure(Kg/cm^2)	234	232
Cp(KJ/mole*C)	30.07	30.6
Moles	7.16E+03	
Enthalpy	3.50E+07	8.30E+06
Q(KJ/hr)	9.20E+07	
Shell Side		
Temperature (C)	45	45
Pressure(Kg/cm2)	29.01	29.01
Cp(KJ/mole*C)	21030	21030
Moles	**1.70E+08**	
Enthalpy	45	45
Q(KJ/hr)	29.01	

Table 37 E-12 energy balance

CHAPTER 6

SIMULATION

Based on real time industrial data simulations are performed for the process using ASPEN HYSYS V 8.8. Ammonia, oxygen, nitrogen, carbon dioxide, carbon monoxide, hydrogen, methane, ethane, propane, butane, and argon were selected as the components for simulation. For the process, Peng Robinson was selected as the fluid package. The process gas is preheated in a furnace and then send to the primary reformer reactor [12] [13].

6.1 ASPEN HYSYS Model of the plant:

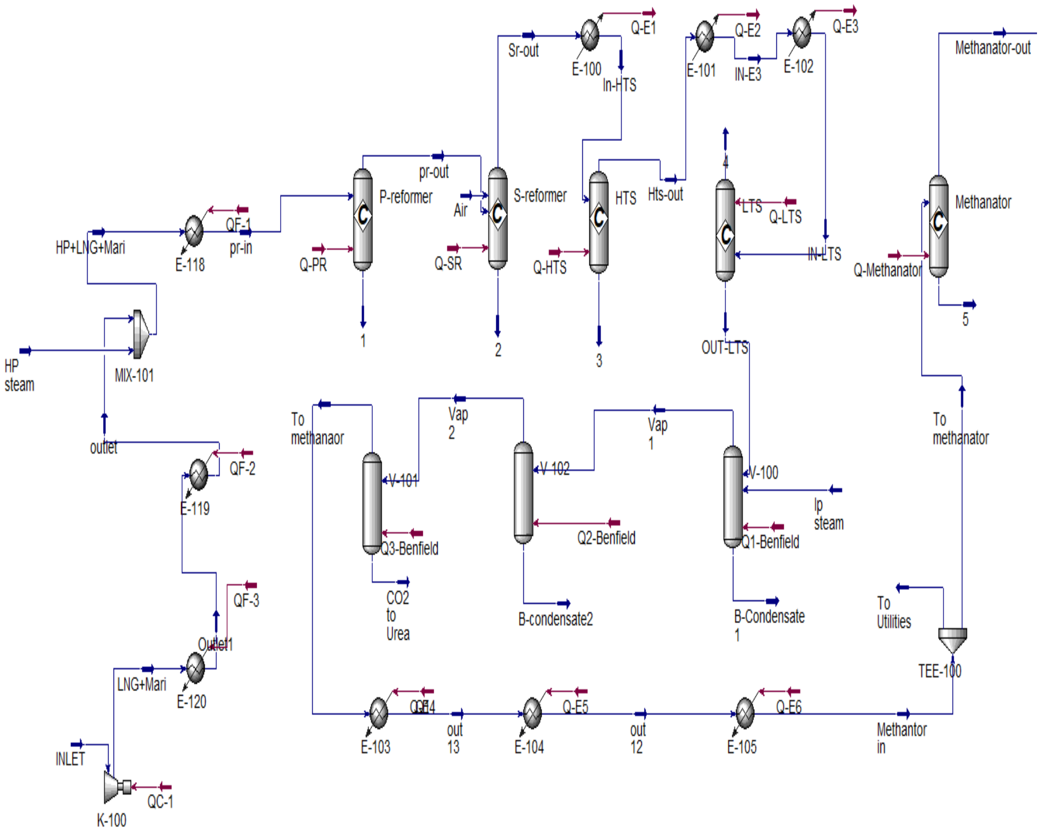


Figure 14 HYSYS Simulation Model A

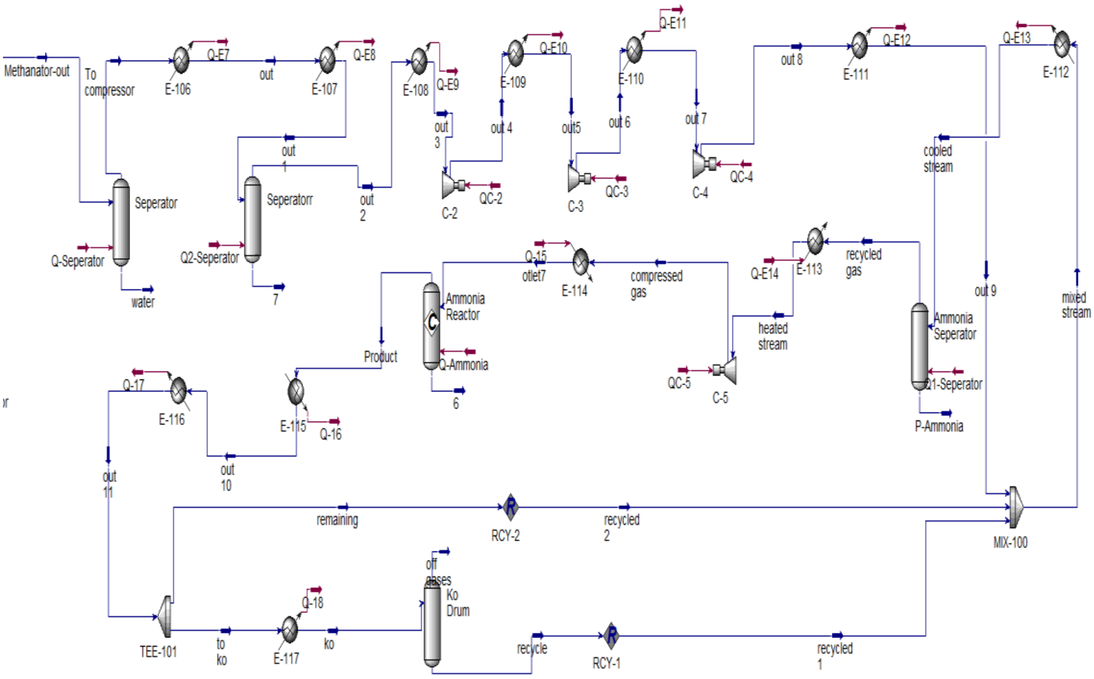
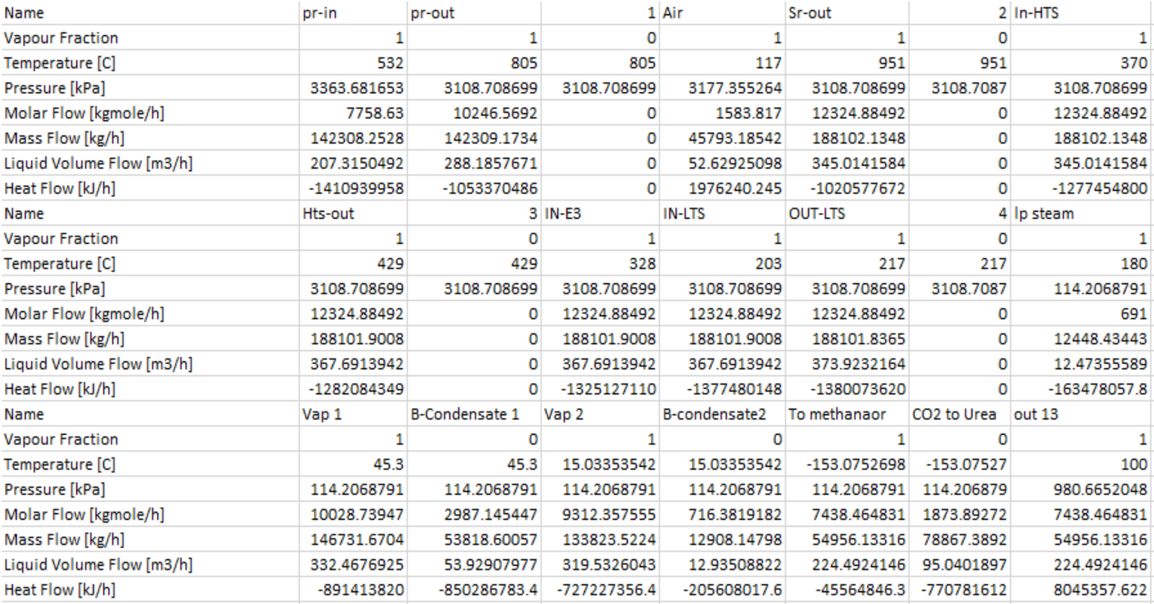


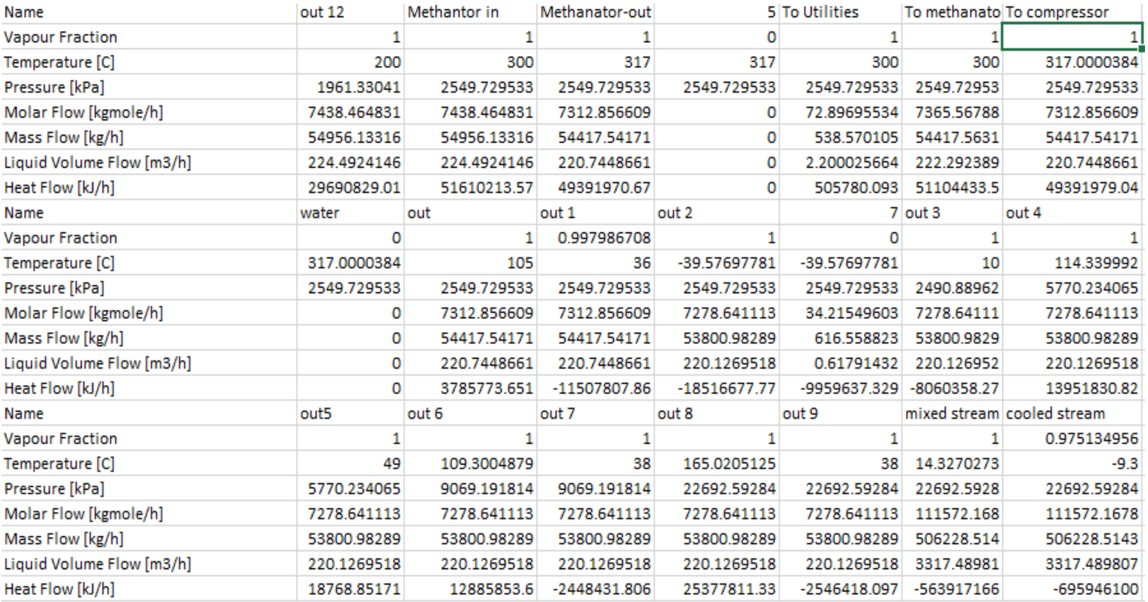
Figure 15 HYSYS Simulation Model B

6.2 Material balance Using HYSYS:



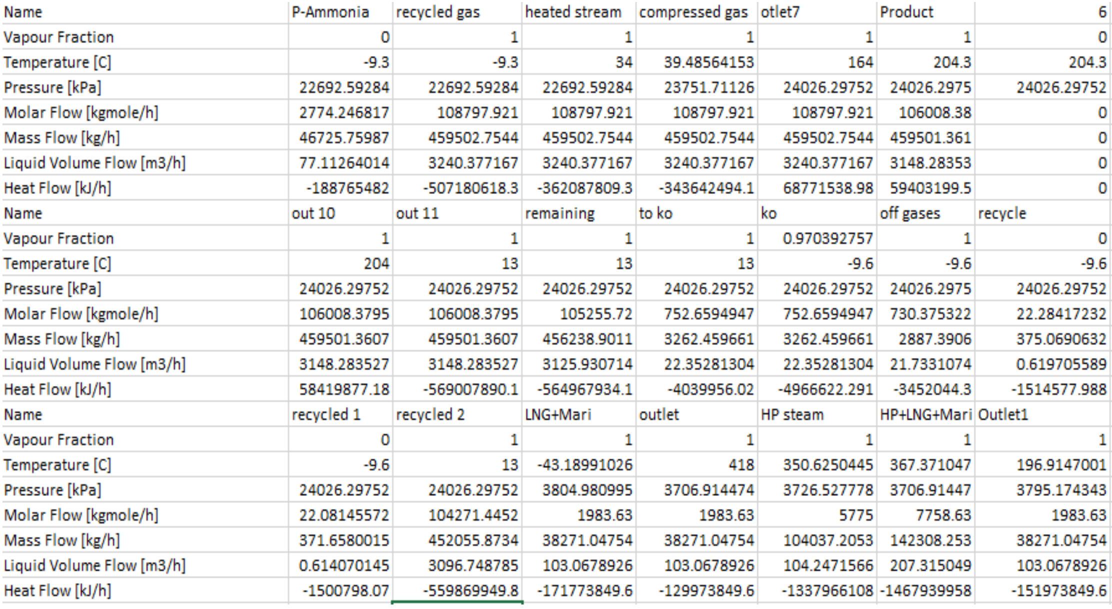
Name	pr-in	pr-out	1	Air	Sr-out	2	In-HTS
Vapour Fraction	1	1	0	1	1	0	1
Temperature [C]	532	805	805	117	951	951	370
Pressure [kPa]	3363.681653	3108.708699	3108.708699	3177.355264	3108.708699	3108.7087	3108.708699
Molar Flow [kgmole/h]	7758.63	10246.5692	0	1583.817	12324.88492	0	12324.88492
Mass Flow [kg/h]	142308.2528	142309.1734	0	45793.18542	188102.1348	0	188102.1348
Liquid Volume Flow [m3/h]	207.3150492	288.1857671	0	52.62925098	345.0141584	0	345.0141584
Heat Flow [kJ/h]	-1410939958	-1053370486	0	1976240.245	-1020577672	0	-1277454800
Name	Hts-out	3	IN-E3	IN-LTS	OUT-LTS	4	lp steam
Vapour Fraction	1	0	1	1	1	0	1
Temperature [C]	429	429	328	203	217	217	180
Pressure [kPa]	3108.708699	3108.708699	3108.708699	3108.708699	3108.708699	3108.7087	114.2068791
Molar Flow [kgmole/h]	12324.88492	0	12324.88492	12324.88492	12324.88492	0	691
Mass Flow [kg/h]	188101.9008	0	188101.9008	188101.9008	188101.8365	0	12448.43443
Liquid Volume Flow [m3/h]	367.6913942	0	367.6913942	367.6913942	373.9232164	0	12.47355589
Heat Flow [kJ/h]	-1282084349	0	-1325127110	-1377480148	-1380073620	0	-163478057.8
Name	Vap 1	B-Condensate 1	Vap 2	B-condensate2	To methanaor	CO2 to Urea	out 13
Vapour Fraction	1	0	1	0	1	0	1
Temperature [C]	45.3	45.3	15.03353542	15.03353542	-153.0752698	-153.07527	100
Pressure [kPa]	114.2068791	114.2068791	114.2068791	114.2068791	114.2068791	114.206879	980.6652048
Molar Flow [kgmole/h]	10028.73947	2987.145447	9312.357555	716.3819182	7438.464831	1873.89272	7438.464831
Mass Flow [kg/h]	146731.6704	53818.60057	133823.5224	12908.14798	54956.13316	78867.3892	54956.13316
Liquid Volume Flow [m3/h]	332.4676925	53.92907977	319.5326043	12.93508822	224.4924146	95.0401897	224.4924146
Heat Flow [kJ/h]	-891413820	-850286783.4	-727227356.4	-205608017.6	-45564846.3	-770781612	8045357.622

Figure 16 Screenshot 1



Name	out 12	Methantor in	Methanator-out	5	To Utilities	To methanato	To compressor
Vapour Fraction	1	1	1	0	1	1	1
Temperature [C]	200	300	317	317	300	300	317.0000384
Pressure [kPa]	1961.33041	2549.729533	2549.729533	2549.729533	2549.729533	2549.72953	2549.729533
Molar Flow [kgmole/h]	7438.464831	7438.464831	7312.856609	0	72.89695534	7365.56788	7312.856609
Mass Flow [kg/h]	54956.13316	54956.13316	54417.54171	0	538.570105	54417.5631	54417.54171
Liquid Volume Flow [m3/h]	224.4924146	224.4924146	220.7448661	0	2.200025664	222.292389	220.7448661
Heat Flow [kJ/h]	29690829.01	51610213.57	49391970.67	0	505780.093	51104433.5	49391979.04
Name	water	out	out 1	out 2	7	out 3	out 4
Vapour Fraction	0	1	0.997986708	1	0	1	1
Temperature [C]	317.0000384	105	36	-39.57697781	-39.57697781	10	114.339992
Pressure [kPa]	2549.729533	2549.729533	2549.729533	2549.729533	2549.729533	2490.88962	5770.234065
Molar Flow [kgmole/h]	0	7312.856609	7312.856609	7278.641113	34.21549603	7278.64111	7278.641113
Mass Flow [kg/h]	0	54417.54171	54417.54171	53800.98289	616.558823	53800.9829	53800.98289
Liquid Volume Flow [m3/h]	0	220.7448661	220.7448661	220.1269518	0.61791432	220.126952	220.1269518
Heat Flow [kJ/h]	0	3785773.651	-11507807.86	-18516677.77	-9959637.329	-8060358.27	13951830.82
Name	out5	out 6	out 7	out 8	out 9	mixed stream	cooled stream
Vapour Fraction	1	1	1	1	1	1	0.975134956
Temperature [C]	49	109.3004879	38	165.0205125	38	14.3270273	-9.3
Pressure [kPa]	5770.234065	9069.191814	9069.191814	22692.59284	22692.59284	22692.5928	22692.59284
Molar Flow [kgmole/h]	7278.641113	7278.641113	7278.641113	7278.641113	7278.641113	111572.168	111572.1678
Mass Flow [kg/h]	53800.98289	53800.98289	53800.98289	53800.98289	53800.98289	506228.514	506228.5143
Liquid Volume Flow [m3/h]	220.1269518	220.1269518	220.1269518	220.1269518	220.1269518	3317.48981	3317.489807
Heat Flow [kJ/h]	18768.85171	12885853.6	-2448431.806	25377811.33	-2546418.097	-563917166	-695946100

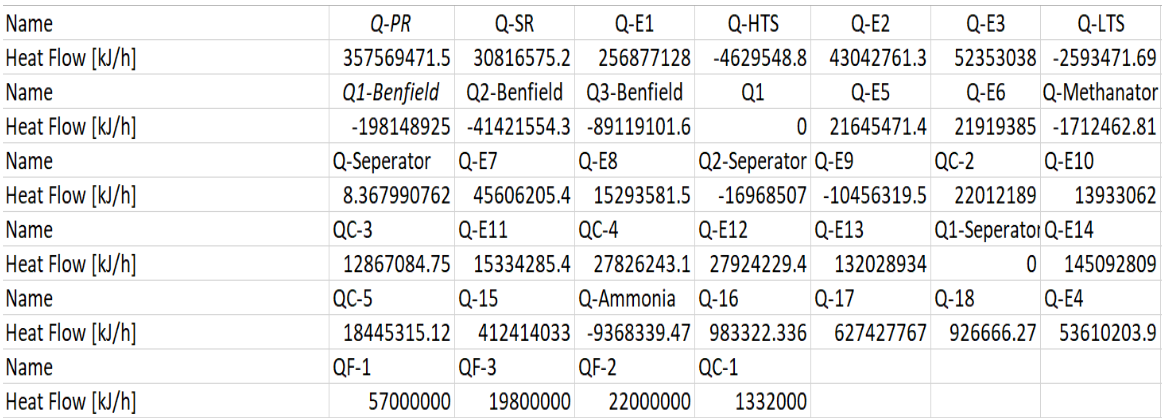
Figure 17 screenshot 2



Name	P-Ammonia	recycled gas	heated stream	compressed gas	otlet7	Product	6
Vapour Fraction	0	1	1	1	1	1	0
Temperature [C]	-9.3	-9.3	34	39.48564153	164	204.3	204.3
Pressure [kPa]	22692.59284	22692.59284	22692.59284	23751.71126	24026.29752	24026.2975	24026.29752
Molar Flow [kgmole/h]	2774.246817	108797.921	108797.921	108797.921	108797.921	106008.38	0
Mass Flow [kg/h]	46725.75987	459502.7544	459502.7544	459502.7544	459502.7544	459501.361	0
Liquid Volume Flow [m3/h]	77.11264014	3240.377167	3240.377167	3240.377167	3240.377167	3148.28353	0
Heat Flow [kJ/h]	-188765482	-507180618.3	-362087809.3	-343642494.1	68771538.98	59403199.5	0
Name	out 10	out 11	remaining	to ko	ko	off gases	recycle
Vapour Fraction	1	1	1	1	0.970392757	1	0
Temperature [C]	204	13	13	13	-9.6	-9.6	-9.6
Pressure [kPa]	24026.29752	24026.29752	24026.29752	24026.29752	24026.29752	24026.2975	24026.29752
Molar Flow [kgmole/h]	106008.3795	106008.3795	105255.72	752.6594947	752.6594947	730.375322	22.28417232
Mass Flow [kg/h]	459501.3607	459501.3607	456238.9011	3262.459661	3262.459661	2887.3906	375.0690632
Liquid Volume Flow [m3/h]	3148.283527	3148.283527	3125.930714	22.35281304	22.35281304	21.7331074	0.619705589
Heat Flow [kJ/h]	58419877.18	-569007890.1	-564967934.1	-4039956.02	-4966622.291	-3452044.3	-1514577.988
Name	recycled 1	recycled 2	LNG+Mari	outlet	HP steam	HP+LNG+Mari	Outlet1
Vapour Fraction	0	1	1	1	1	1	1
Temperature [C]	-9.6	13	-43.18991026	418	350.6250445	367.371047	196.9147001
Pressure [kPa]	24026.29752	24026.29752	3804.980995	3706.914474	3726.527778	3706.91447	3795.174343
Molar Flow [kgmole/h]	22.08145572	104271.4452	1983.63	1983.63	5775	7758.63	1983.63
Mass Flow [kg/h]	371.6580015	452055.8734	38271.04754	38271.04754	104037.2053	142308.253	38271.04754
Liquid Volume Flow [m3/h]	0.614070145	3096.748785	103.0678926	103.0678926	104.2471566	207.315049	103.0678926
Heat Flow [kJ/h]	-1500798.07	-559869949.8	-171773849.6	-129973849.6	-1337966108	-1467939958	-151973849.6

Figure 18 screenshot 3

6.3 Energy balance using HYSYS:



Name	*Q-PR*	Q-SR	Q-E1	Q-HTS	Q-E2	Q-E3	Q-LTS
Heat Flow [kJ/h]	357569471.5	30816575.2	256877128	-4629548.8	43042761.3	52353038	-2593471.69
Name	*Q1-Benfield*	Q2-Benfield	Q3-Benfield	Q1	Q-E5	Q-E6	Q-Methanator
Heat Flow [kJ/h]	-198148925	-41421554.3	-89119101.6	0	21645471.4	21919385	-1712462.81
Name	Q-Seperator	Q-E7	Q-E8	Q2-Seperator	Q-E9	QC-2	Q-E10
Heat Flow [kJ/h]	8.367990762	45606205.4	15293581.5	-16968507	-10456319.5	22012189	13933062
Name	QC-3	Q-E11	QC-4	Q-E12	Q-E13	Q1-Seperator	Q-E14
Heat Flow [kJ/h]	12867084.75	15334285.4	27826243.1	27924229.4	132028934	0	145092809
Name	QC-5	Q-15	Q-Ammonia	Q-16	Q-17	Q-18	Q-E4
Heat Flow [kJ/h]	18445315.12	412414033	-9368339.47	983322.336	627427767	926666.27	53610203.9
Name	QF-1	QF-3	QF-2	QC-1			
Heat Flow [kJ/h]	57000000	19800000	22000000	1332000			

Figure 19Energy balance screenshot

CHAPTER 7

SELECTION AND EQUIPMENT DESIGN

7.1 Design Calculations of Reactors:

Q = Heat Flow	V_r = Volume of reactor
V_b = Volume of bed	N= No of tubes
E = Void Fraction	Q_{avg} = Average heat flux
L = Length of tube	I.D = Internal Diameter of tube
W = Weight of catalyst	V_c = Volume of catalyst

7.1.1 Primary Reformer R-101:

Selection:

Primary reformer is a type of furnace having tubes in it. Reactions are taking place in these tubes having catalyst in it and they are heated by the burners inside the walls of furnace. For designing we chose it to be a Foster Wheelar side fired reformer due to reduced methane slip and it can bear small amount of coking because of propane and butane in our feed. Tubes and Outer body are made of stainless steel which are bearable for the reactor pressure. Design calculations are given below [14].

No of tubes:

$$N = Q/\pi * I.D * L * Q_{avg} \qquad \textbf{(i)}$$

Q= 8.545x10^7 kcal/hr

I.D = 0.113m

L = 11m

Qavg= 65800 kcal/m²hr

By putting these values in above Equation (i) we get no of tubes,

N= (8.545x10^7)/(π)*(0.113) *(11) *(65800)

N= 332 tubes

Volume of reactor:

Volume of reactor = N.π.ID^2.L/4

V$_r$=(332)*(3.14)*(0.113)2*.11/4

V$_r$ = 36.6 m^3

Volume of Catalyst:

Catalyst used R-67R-7H (Ni 15%, MgO25%, Al2O360%)

Bulk density of catalyst = 970 kg/m^3

Voidage = E = 0.131

E = (V$_r$ – V$_c$)/V$_r$

As we know E and Vr, by putting values we get V$_c$. (Volume of Catalyst)

V$_c$ = 31.8 m^3

Weight of Catalyst:

Weight of Catalyst = (Bulk density) *(Volume of catalyst)

Weight of catalyst = (970) *(31.8)

Weight of catalyst = 30,846 kg

Space Velocity:

Space Velocity = Volumetric Feed Flow rate/ Volume of reactor

Space velocity = 173801/36.6*3600

Space velocity = 1.31 sec^{-1}

Comparison:

By comparing the design calculation, we can clearly see the impact of feed on primary reformer specifications. As the duty is increased the weight of catalyst and volume of reactor also increased for the reforming reaction to take place. No. of tubes of primary reformer also increased.

Specifications	100% Mari Gas	70% Mari Gas and 30% LNG
Duty	7.39x10^7 kcal/hr	8.545x10^7 kcal/hr
Internal Dia of tube	0.113m	0.113m
Length of tube	11m	11m
Average heat flux	65800 kcal/m^{2}hr	65800 kcal/m^{2}hr
Volume of Catalyst:	27.6m^3	31.8m^3
Volume of reactor	31.8m^3	36.6m^3
Voidage	0.131	0.131
Weight of Catalyst:	26,543 kg	30,846 kg
No. of tubes	288	332

Table 23 R- 101 Specification comparison

7.1.2 Secondary Reformer R-102:

There are two types of reactions that are involved in primary reformer, three of them are combustion reactions and then another type is reforming reaction which is involved in the reactor bed [15].

We will take into account reforming reaction which is taking place in the bed for the design of this reactor.

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

Selection:

Is this reactor, we have combustion reactions and reforming reactions. The purpose is to extract nitrogen from air for ammonia reaction. So, we need a reaction having burner in it for combustion. And for reforming reaction we need to have a packed bed reactor under it where the reforming reaction takes place. Packed bed reaction is required as it is a catalytic reaction and gaseous phase reaction and also it gives more conversion per weight of catalyst.

Weight of Catalyst:

The catalyst used in secondary reformer: RKS-2&; RKS-2-7H

Catalyst composition = NiO, Promotor/Carrier = MgO/ Al_2O_3

Design equation for the reactor design is,

$$\frac{W}{F_{AO}} = \int_0^{X_A} \frac{dX_A}{-r_a} \qquad \text{(ii)}$$

Rate equation $-r_a = K\, C_{Ch4}^{n}\, C_{H2O}^{m}$

Order of reaction is n = 1, m= 1 Conversion Xa = 90%

$F_{Ao} = C_{Ao}.V_o$, $C_{Ch4} = C_{Ao} = 0.98\ mol/m^3$

$M = C_{H2O}/\ C_{Ch4} = 8.70$

$K = 2.5 \times 10^{-2}\ mol^{-1}\ sec^{-1}$

Putting this value in equation (ii)

$$,\frac{W}{Ch4.Vo} = \int_0^{X_A} \frac{dX_A}{KCh4(1-Xa)(M-Xa)}$$

Solving the integral by area under the curve we get,

Weight of catalyst = 32,568

Weight of catalyst = 32,568 kg of catalyst

Bulk density of catalyst = 1070 kg/m^3 (Bulk density)

Volume of catalyst:

Volume of catalyst = Weight of catalyst/ Bulk density of catalyst

Volume of catalyst = 32568/1070

$V_c=30.43m^3$

Volume of reactor:

Voidage = 0.25

$E = (V_b - V_c)/V_b$

Volume of reactor bed = $V_b = 40.57\ m^3$

With 20% clearance, $V_r = 1.20*Vb$

$$\mathbf{Vr = 48.684\ m^3}$$

Height of reactor:

$H_r = Vr/\ \pi*D^2/4$, $D_r = 4m$

$$\mathbf{H_r = 3.87m}$$

Space Velocity:

Space Velocity = Volumetric Feed Flow rate/ Volume of reactor

Space Velocity= 457.5/48.684*3600

$$\mathbf{Space\ velocity = 2.6 \times 10^3 sec^{-1}}$$

Comparison:

Specifications	**100% Mari Gas**	**70% Mari Gas and 30% LNG**
Internal Dia of tube	3m	3m
Height of reactor	6.19 m	6.87m
Volume of Catalyst:	28.4m³	30.43m³
Volume of reactor	43.76m³	48.684 m³
Voidage	0.25	0.25
Weight of Catalyst:	29,580 kg	32,568 kg

Table 25 R-102 Specification comparison

Due to change in feed, outlet of primary reformer is changed which is inlet of secondary reformer, due to which in order to get desirable conversion volume of reactor and catalyst has also been increased.

7.1.3 High Temperature Shift R- 103:

Selection:

For high temperature shift reactor, it was chosen as packed bed reactor because it will give more conversion per weight of catalyst and high temperature shift conversion is a gaseous phase reaction. The reaction is exothermic and favoured at low temperature but due to high concentration of reactants converted into product, It will have low reaction rate, so for high rate of reaction it takes place at high temperature in presence of selective catalyst (iron based) which is being used. Stainless steel material of reactor is used as it is less corrosive and can bear reactor pressure and temperature [13].

This reactor involves the following reaction,

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

Conversion Xa = 77%

Catalyst used = SK-201-2

Catalyst composition = Fe_3O_4 ~80-90%

Promotor/Carrier = Cr_2O_3 ~8-13%/ CuO ~1-2%

Weight of Catalyst:

Design equation used for reactor design is,

$$\frac{W}{F_{AO}} = \int_0^{X_A} \frac{dX_A}{-r_a} \qquad \text{(iii)}$$

$-r_a = K\, C_{CO}^{\,n}\, C_{H2O}^{\,m}$ Order of reaction n = 1, m=0

$F_{Ao} = C_{Ao}.V_o$,

Vo = 276062 m^3/hr

After re arranging equation (iii) becomes,

$$\frac{W}{\mathrm{CCo.Vo}} = \int_0^{X_A} \frac{dX_A}{\mathrm{K\,CCO(1-Xa)}}$$

$$\frac{W}{\mathrm{Vo}} = \int_0^{X_A} \frac{dX_A}{\mathrm{K\,(1-Xa)}}$$

K = 6.031 sec^{-1}

Vo = 276062 m^3/hr

Putting these values we get weight of catalyst,

$$\frac{W}{276062} = \int_{0}^{0.77} \frac{dX_A}{6.031(1 - \mathrm{Xa})}$$

Weight of catalyst = 66829.79 kg

Volume of catalyst:

Bulk density of catalyst: 1050kg/m^3

Volume of catalyst = Weight of catalyst/ Bulk density of catalyst

Volume of catalyst = 66829.79/1050

V_c = 64.64 m^3

Volume of reactor:

Voidage = 0.044

E = $(V_b - V_c)/V_b$

V_b = 67.61 m_3

29% clearance gives the volume of reactor,

Vr = 1.29*Vb

V_r = 87.22m^3

Height of Reactor:

D_r = 4.3m

Hr = Vr/ π*D2/4

Height of reactor = 4.65m

Space velocity:

Space Velocity = Volumetric Feed Flow rate/ Volume of reactor

Space Velocity= 276057/87.22*3600

Space Velocity = 2.3 x10 -4 sec^{-1}

Comparison:

The comparison for both feed in HTS is being shown in table, the amount of carbon monoxide to convert into carbon dioxide is increased for new feed as compare to previous feed. Therefore amount of catalyst also increased to undergo the reaction at given conversion. In this way the volume of reactor also increased

Specifications	100% Mari Gas	70% Mari Gas and 30% LNG
Internal Dia of tube	4.3m	4.3m
Height of reactor	7 m	7.64 m
Volume of Catalyst:	82 m^3	86.12 m^3
Volume of reactor	102.5 m^3	111 m^3
Voidage	0.036	0.036
Weight of Catalyst:	86,920 kg	91,288 kg

.

Table 26 R-103 Specifications comparison

7.1.4 Low Temperature Shift R- 104:

Selection:

For low temperature shift reactor, it was chosen as packed bed reactor because it will give more conversion per weight of catalyst and CO conversion is a gaseous phase reaction. The reaction is exothermic and favoured at low temperature in presence of selective catalyst (iron based) which is being used. Here low temperature is maintained because amount of CO which needs to be converted is low hence low rate of reaction is not a problem in this case. Stainless steel material of reactor is used as it is less corrosive and can bear reactor pressure and temperature. Design calculations for the specification of low temperature shift are given below.

This reaction involves the following reaction, but at low temperature.

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

Conversion Xa = 92%

Catalyst used = Katalco 83-3

Catalyst composition = CuO 51%

Promotor/Carrier = $Al_2O_3$18%/ ZnO 31%

Weight of catalyst:

Design equation used for reactor design is,

$$\frac{W}{F_{AO}} = \int_0^{X_A} \frac{dX_A}{-r_a} \qquad \text{(iii)}$$

$-r_a = K C_{CO}^n C_{H2O}^m$ Order of reaction n = 1, m=1.4

$F_{Ao} = C_{Ao}.V_o$, A_0 = CO , B_O= H_2O

C_{Ao} = 8.44x10^{-4} mol/m^3

C_{Bo}= 0.0121 mol/m^3

$$\frac{W}{\text{CCo. Vo}} = \int_0^{X_A} \frac{dX_A}{\text{K CCO}^{2.4}(1 - xa)(M - 1.4xa)}$$

K = 1.23x10^4

By solving integral area under the curve we get,

$$\frac{W}{276050} = \frac{0.2024}{1.23\times10^4(8.44\times10^{-4})^{1.4}}$$

Weight of catalyst = 91288 kg

Volume of catalyst:

Bulk Density of Catalyst = 1060 kg/m^3

Volume of catalyst = Weight of catalyst/ Bulk density of catalyst

Volume of catalyst = 91288/1060

V_c = 86.12m^3

Volume of reactor:

E = (Vb – Vc)/Vb

E = 0.036

E = (Vb-Vc)/Vb

Vb = 89.33m^3

25% clearance of reactor with bed

Volume of reactor, Vr= 1.25*89.336

Vr = 111.67 m^3

Height of Reactor:

Dr = 4.3m

Hr = Vr/ π*D2/4

Height of reactor = 6.15m

Space velocity:

Space Velocity = Volumetric Feed Flow rate/ Volume of reactor

Space Velocity= 276057/111.67 *3600

Space Velocity = 0.685 sec^{-1}

Comparison:

Specifications	100% Mari Gas	70% Mari Gas and 30% LNG
Internal Dia of tube	4.3m	4.3m
Height of reactor	5.8 m	6.09 m
Volume of Catalyst:	61m^3	64.4m^3
Volume of reactor	84 m^3	87.22 m^3
Voidage	0.04	0.04
Weight of Catalyst:	64,050 kg	66,829 kg

Table 27 R-104 Specifications comparison

The comparison for both feed in HTS is being shown in table, as the amount of carbon monoxide to convert into carbon dioxide is increased as compare to previous feed. The amount of catalyst also increased to undergo the reaction at given conversion. In this way the volume of reactor also increased.

7.1.5 Methanator R-105:

Selection:

Methanator consist of two reactions in which carbon dioxide and carbon monoxide is being completely removed as it can cause coking in ammonia synthesis convertor. It is gaseous phase catalytic reaction, so we chose packed bed reactor. It can also give higher conversion per weight of catalyst. Stainless steel is used as material of reactor because it can bear reactor temperature and pressure. Design calculation for specification of reactor are given below.

These two reactions took place in methanator,

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

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

Weight of catalyst:

For both of reactions weight of catalyst is determined individually and then by combined weight of catalyst is being calculated.

Design equation used for reactor design is,

$$\frac{W}{F_{AO}} = \int_0^{X_A} \frac{dX_A}{-r_a} \quad \text{(iii)}$$

For reaction 1: -

$-r_a = K\, C_{CO2}^{n}\, C_{H2}^{m}$ Order of reaction n = 1, m=1

$F_{Ao} = C_{Ao}.V_o$, $A_0 = CO_2$, $B_O = H_2$

$C_{Ao} = 4.54\text{x}10^{-5}$ mol/m^3

$C_{Bo} = 0.03519$ mol/m^3

M = 775.26

$V_o = 16321.2 m^3/hr$

$k = koe^{-Ea/RT}$

K = 2.7 x 10^5 mol^{-1} sec^{-1}

$$\frac{W}{\text{Vo}} = \int_0^{X_A} \frac{dX_A}{\text{K}\, CCo2(1 - xa)(775.26 - xa)}$$

By solving the integral by area under the curve and putting the values we get,

$$\frac{W}{163213.1} = \frac{20.6}{2.7 \times 10^{5}(4.54\ \times 10^{-5})}$$

W_1 = 27426.84kgs of Catalyst

$\frac{W}{F_{AO}} = \int_0^{X_A} \frac{dX_A}{-r_a}$ (iii)

For reaction 2: -

$-r_a = K\, C_{CO}^{n}\, C_{H2}^{m}$ Order of reaction n = 1, m=1

$F_{Ao} = C_{Ao}.V_o$, A_0 = CO, $B_O = H_2$

$C_{Ao} = 1.11x10^{-4}$ mol/m^3

C_{Bo}= 0.03519 mol/m^3

M = 315.26

V_o=16321.2m^3/hr

k= ko$e^{-Ea/RT}$

K = 2.7 x 10^5 mol^{-1} sec^{-1}

$$\frac{W}{\text{Vo}} = \int_0^{X_A} \frac{dX_A}{\text{K}\, CCo(1 - xa)(315.26 - xa)}$$

By solving the integral by area under the curve and putting the values we get,

$$\frac{W}{163213.1} = \frac{83.31}{3.5 \times 10^{7}(1.11 \times 10^{-4})}$$

W_2 = 34999.36 kgs of Catalyst

$W_t = W_1 + W_2$

W_T= 62426.171 kg of catalyst

Volume of catalyst:

Bulk Density of Catalyst = 950 kg/m^3

Volume of catalyst = Weight of catalyst/ Bulk density of catalyst

Volume of catalyst = 62426.171/950

Vc = 65.71m^3

Bulk Density of Catalyst = 950 kg/m^3

Volume of catalyst = Weight of catalyst/ Bulk density of catalyst

Volume of catalyst = 62426.171/950

Vc = 65.71m^3

Volume of reactor:

E= 0.04

E = (Vb – Vc)/Vb

Vb= 68.44m^3

20% clearance to reactor with bed,

Vr= 1.20*(68.44)

Vr=82.128m^3

Height of Reactor:

Dr = 3.5m

Hr = Vr/ π*D^2/4

Height of reactor = 8.54 m

Space velocity:

Space Velocity = Volumetric Feed Flow rate/ Volume of reactor

Space Velocity= 16321.2/82.128*3600

Space Velocity = 5.52x10^{-5} sec^{-1}

Comparison:

Specifications	100% Mari Gas	70% Mari Gas and 30% LNG
Internal Dia of tube	3.5 m	3.5 m
Height of reactor	8.1 m	8.54 m
Volume of Catalyst	62 m^3	65.71 m^3
Volume of reactor	78 m^3	82.12 m^3
Voidage	0.04	0.04
Weight of Catalyst:	58,900 kg	62,426 kg

Table 28 R-105 Specifications comparison

As the amount of carbon dioxide and carbon monoxide is being increased as compare to previous feed. The amount of catalyst required to achieve conversion also increased. In this way the volume of catalyst also increased.

7.1.6 Ammonia reactor 106:

Selection:

For the selection of ammonia reactor, it is gaseous phase reaction in the presence of catalyst, so we choose packed bed reactor as it gives higher conversion per weight of catalyst. Stainless steel type 316 is being used as material of reactor as it can bear the reactor temperature and pressure. Calculation for the design of specification are given below [16].

The reaction takes place in ammonia reactor is

$N_2 + 3H_2 \rightarrow 2NH_3$

The conversion of this reaction is 29.3%

The catalyst used in this reaction is triple promoted iron oxide catalyst

(Fe_2O_3, FeO) promoted by **(K_2O, CaO, Al_2O_3).**

Weight of catalyst:

The design equation for the reactor is following:

$$\frac{W}{F_{AO}} = \int_0^{X_A} \frac{dX_A}{-r_a}$$

Where the rate equation is,

$$r_{NH_3} = 2k\left(K_a^2 a_{N_2}\left[\frac{a_{H_2}^3}{a_{NH_3}^2}\right]^{\alpha} - \left[\frac{a_{NH_3}^2}{a_{H_2}^3}\right]^{1-\alpha}\right)$$

Where

K = Equilibrium constant

k = rate constant

ai = activity coefficient of component i

α = constant which takes a value of 0.5-0.75. In this work it is assumed to be 0.5.

As,

ai=fi /fo

Fo is reference fugacity, taken as 1 atm.

Fugacity can be calculated from value of fugacity coefficient as: $fi = \phi i\ P$i

$$\emptyset_{H_2} = \left(exp \left(\begin{matrix} \exp(-3.802T^{0.125} + 0.541)\,P - exp(-0.1263T^{0.5} - 15.98)P^2 \\ + \left(300\left(\exp(-0.011901T - 5.941)\left(\exp\frac{P}{300}\right)\right)\right) \end{matrix} \right) \right)$$

$$\emptyset_{N_2} = \left(\begin{matrix} 0.93431737 + 0.2028538x10^{-3}T + 0.295896x10^{-3}P \\ -0.270727x10^{-6}T^2 + 0.4775207x10^{-6}P^2 \end{matrix} \right)$$

$$\emptyset_{NH_3} = \left(\begin{matrix} 0.1438996 + 0.2028538x10^{-2}T - 0.4487672x10^{-3}P \\ -0.1142945x10^{-5}T^2 + 0.2761216x10^{-6}P^2 \end{matrix} \right)$$

$\emptyset_{H2} = 1.002$ $\emptyset_{N2} = 1.080$ $\emptyset_{NH3} = 1.098$

Pi = xi*P for calculation of partial pressure.

For equilibrium constant:

$$\log_{10} K = -2.691122 \log_{10} T - 5.519265 \times 10^{-5} T + 1.848863 \times 10^{-7} T^2 + \frac{2001.6}{T} + 2.6899$$

Rate constant

K= Koexp(Ea/RT)

ko = Arrhenius coefficient = (8.849×10^{14})

E= Activation energy with temperature its mean value is 40765 Kcal/kmol

R=Universal Gas constant R (8.314 kJ/kmol.K)

Putting all the values and differential equations are solved using Euler's method with step size of 0.005. In the first bed reaction is carried out until, Xf=0.293. At this conversion, the required catalyst weight and volume is:

$$\frac{W}{4645} = \int_0^{X_A} \frac{dX_A}{-ra}$$

Weight of catalyst = 31536.47 kg

Volume of catalyst:

As we know the bulk density of catalyst= 1250kg/m^3

Volume of catalyst = Weight of catalyst/ Bulk density of catalyst

Volume of catalyst = 31536.47 /1250

Vc = 25.22918 m^3

Volume of Reactor:

E= 0.5

E = (Vb – Vc)/Vb

$V_b = 50.45\ m^3$

20% clearance to reactor with bed,

$V_r = 1.20*(50.45)$

$$\mathbf{V_r = 60.54 m^3}$$

Height of reactor:

$D_r = 4.3m$

$H_r = V_r/ \pi*D^2/4$

Height of reactor = 4.17m

Comparison:

As the amount of hydrogen and nitrogen is increased as compare to previous feed. The weight of catalyst also increased to achieve required conversion in a specific time. Volume of reactor also increases

Specifications	100% Mari Gas	70% Mari Gas and 30% LNG
Internal Dia of tube	3.5 m	3.5 m
Height of reactor	5.9 m	6.29 m
Volume of Catalyst:	23.8 m^3	25.2 m^3
Volume of reactor	57.24 m^3	60.54 m^3
Voidage	0.5	0.5
Weight of Catalyst:	30,225 kg	31,536 kg

Table 30 R-106 specifications comparison

7.2 Heat Exchangers Designing:

7.2.1 E-2(Shell and Tube Heat Exchanger)

Water (Shell side):	Process gas (Tube side):
$K = 0.6108$ W/m.°C	$K = 0.1287$ W/m.°C
$\mu = 8.9 \times 10^{-4}$ Ns/m^2	$\mu = 2.17 \times 10^{-5}$ Ns/m^2
$C_p = 4.312$ kJ/kg. °C	$C_p = 2.203$ kJ/kg. °C
$\rho = 1000$ kg/m^3	$\dot{m} = 5.25$ kg/s
$\dot{m} = 67.92$ kg/s	$\rho = 8.1$kg/m^3

$T_{hi} = 429$ °C ; $T_{ho} = 328$ °C

$T_{ci} = 30$ °C ; $T_{co} = 45$ °C

Heat duty:

$Q = \dot{m} \times C_p \times \Delta T$

$= 5.25 \times 2.203 \times 101$

$= 12777.7$ kW

Flow rate of Water:

$Q = \dot{m} \times C_p \times \Delta T \rightarrow \dot{m} = \frac{Q}{\mathrm{Cp} \times \Delta \mathrm{T}}$

$\dot{m} = \frac{12777.7}{4.312 \times 15}$

$\dot{m} = 67.92$ kg/s

Log Mean Temperature Difference (LMTD):

$\mathrm{LMTD} = \frac{\Delta T2 - \Delta T1}{\ln\left(\frac{\Delta T2}{\Delta T1}\right)}$ ΔT_2 = Temperature difference of hot fluids

ΔT_1 = Temperature difference of cold fluids

$\mathrm{LMTD} = \frac{101 - 15}{\ln(101/15)}$

LMTD = 325.19 ^{o}C

R = $\frac{T_1 - T_2}{t_2 - t_1}$

S = $\frac{t_2 - t_1}{t_1 - t_1}$

R = $\frac{429-328}{45-30}$ = 2.24

S = $\frac{45-30}{429-45}$ = 0.112

From graph, 1 shell and 2 tube passes

F_t = 0.98

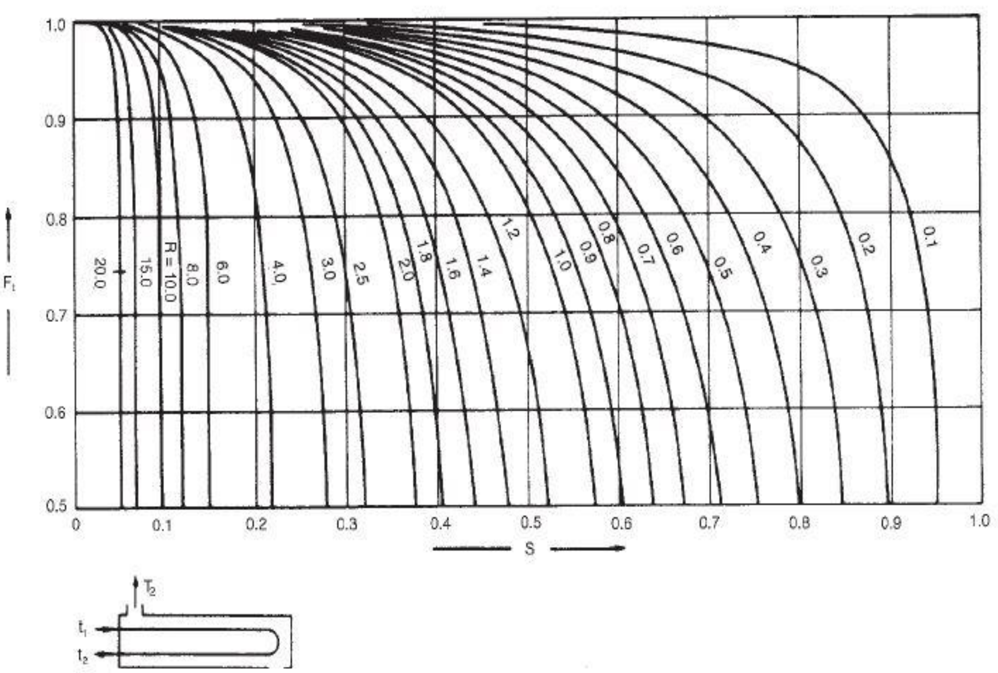
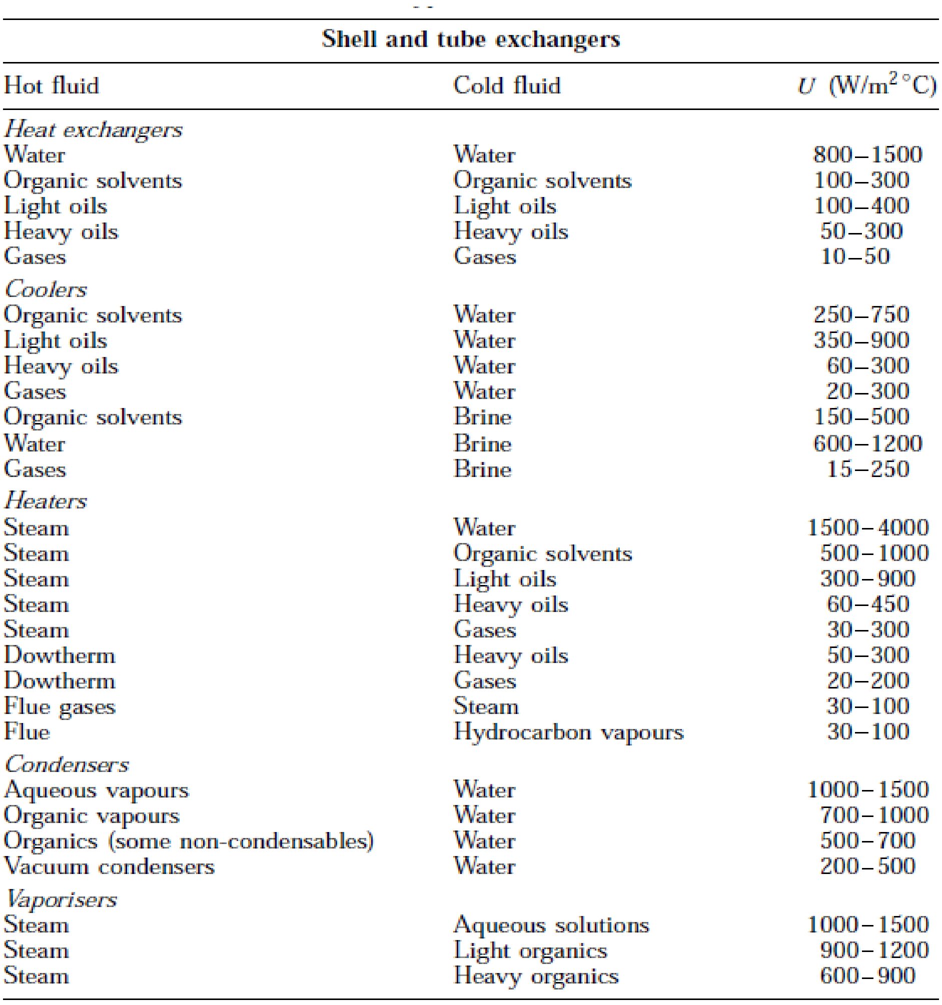


Figure 20 Correction Factor graph

So, corrected LMTD (Δt) = 0.987 x325.19 = 318.68 ^{o}C

From figure,



Shell and tube exchangers		
Hot fluid	Cold fluid	U (W/m^2 °C)
Heat exchangers		
Water	Water	800–1500
Organic solvents	Organic solvents	100–300
Light oils	Light oils	100–400
Heavy oils	Heavy oils	50–300
Gases	Gases	10–50
Coolers		
Organic solvents	Water	250–750
Light oils	Water	350–900
Heavy oils	Water	60–300
Gases	Water	20–300
Organic solvents	Brine	150–500
Water	Brine	600–1200
Gases	Brine	15–250
Heaters		
Steam	Water	1500–4000
Steam	Organic solvents	500–1000
Steam	Light oils	300–900
Steam	Heavy oils	60–450
Steam	Gases	30–300
Dowtherm	Heavy oils	50–300
Dowtherm	Gases	20–200
Flue gases	Steam	30–100
Flue	Hydrocarbon vapours	30–100
Condensers		
Aqueous vapours	Water	1000–1500
Organic vapours	Water	700–1000
Organics (some non-condensables)	Water	500–700
Vacuum condensers	Water	200–500
Vaporisers		
Steam	Aqueous solutions	1000–1500
Steam	Light organics	900–1200
Steam	Heavy organics	600–900

Figure 21 Shell and tube heat exchanger 'U' ranges

U = 300 W/m^2 ^{o}C from above table

$A = \frac{12777.7}{300 \text{ x } 318} = 133.65 m^2$

0.03m Outer diameter and 0.025m Internal diameter – BWG 17

Length of tube = L = 3.66

Area of tube = π x 3.66x 0.03 = 0.3447 m^2

Number of tubes required = $\frac{133.65}{0.334}$ = 387 tubes

Bundle diameter = $D_b = d.(\frac{N_s}{K})^{0.25}$ = 0.838 mm   From graph

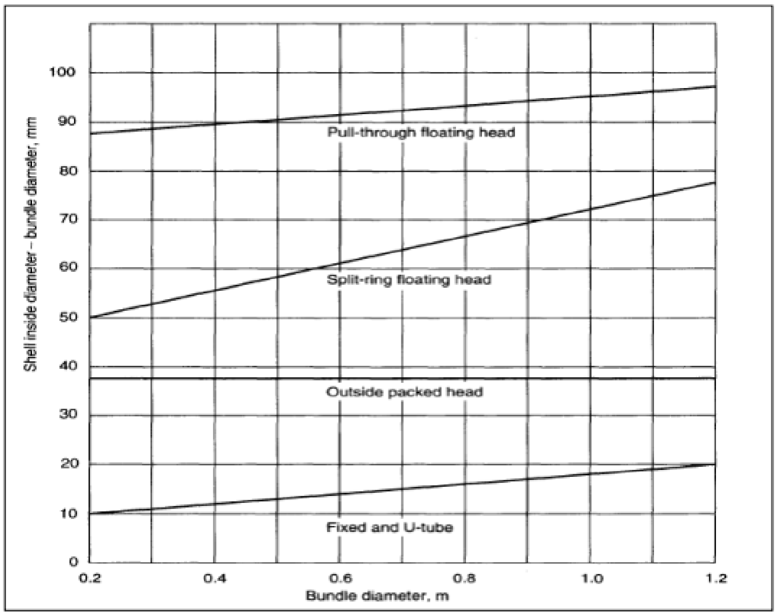
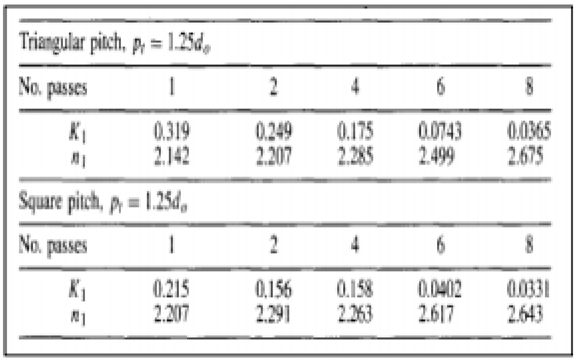


Figure 22 Bundle diameter graph

Clearance = 0.093



Triangular pitch, $p_t = 1.25d_o$

No. passes	1	2	4	6	8
K_1	0.319	0.249	0.175	0.0743	0.0365
n_1	2.142	2.207	2.285	2.499	2.675

Square pitch, $p_t = 1.25d_o$

No. passes	1	2	4	6	8
K_1	0.215	0.156	0.158	0.0402	0.0331
n_1	2.207	2.291	2.263	2.617	2.643

Figure 23 Pitch Factor table

Shell diameter = 0.838 +0.093 = 0.9134m

Shell side co-efficient:

Baffle spacing (l_B) = 0.5 x D_s = 0.4657 m

P_t = 1.25 x d_o = 0.0375m

$A_s = \frac{(P_t - d_0) \times D_s}{P_t}$ change

$A_s = 0.18628\ m^2$

Mass velocity (Ga) = $\frac{\omega}{A_s} = \frac{67.97}{A_s} = 346.61\ kg/m^2s$

$D_e = \frac{1.10}{d_0}(P_t^2 - 0.917 d_0^2)$

$D_e = 0.021$

$Re_b = 9571$

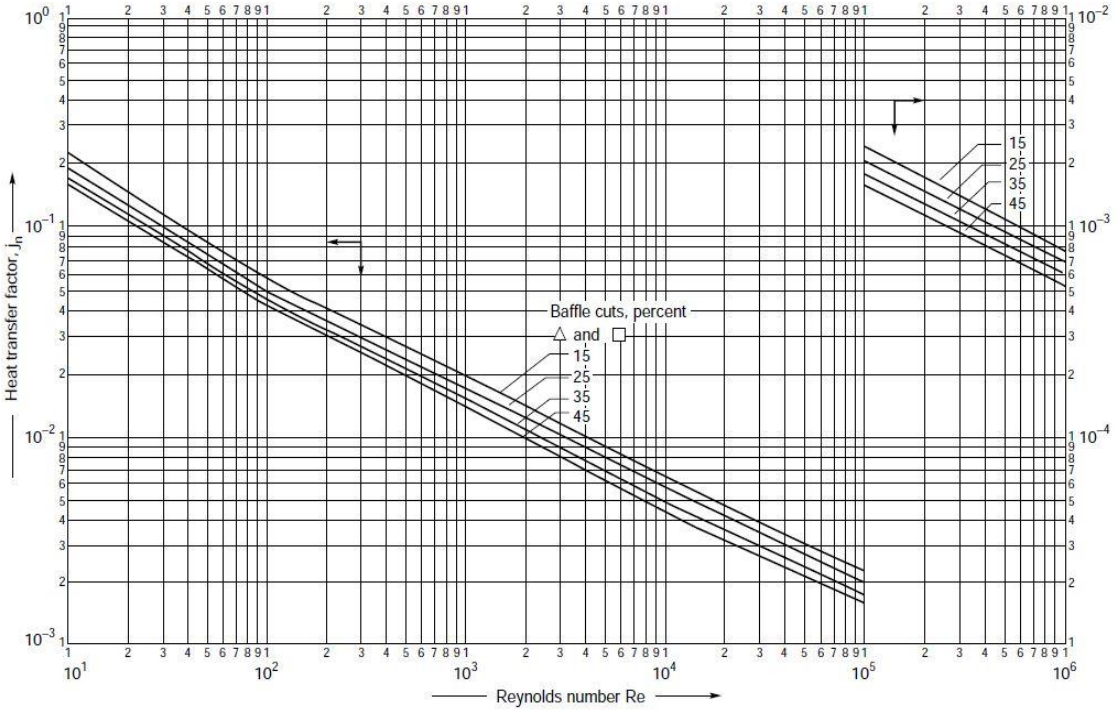


Figure 24 Reynolds number and J_h Graph

From figure, $J_h = 0.006$

$Pr = \frac{\mu \times Cp}{K} = 5.572$

$\frac{h_o}{\theta_s} = \frac{J_h \times Re \times Pr^{0.33} \times K_f}{De} = 28882.15\ W/m^2.\ ^oC$ $\theta_s = 1$

$h_o = 28882.15\ W/m^2.\ ^oC$

Tube side co-efficient:

Tube cross section area= $\frac{\pi}{4} d_i^2$

= 0.0005 m^2

Tube per pass = $\frac{387.72}{2}$ = 193.86 tubes

Total flow area = 193.86 x 0.0005 = 0.09693 m^2

Mass velocity (Ga) = $\frac{\omega}{A}$ = $\frac{52.22}{0.09693}$ = 538.73 kg/m^{2}s

u = $\frac{538.7}{8.81}$ = 61.21 m/s

Re= $\frac{\text{Di x Ga}}{\mu}$ = 620562.212

Pr = $\frac{\mu \text{ x Cp}}{K}$ = $\frac{2.17 \text{ x } 10^{-5} \text{ x } 2.23}{0.127}$ = 0.373

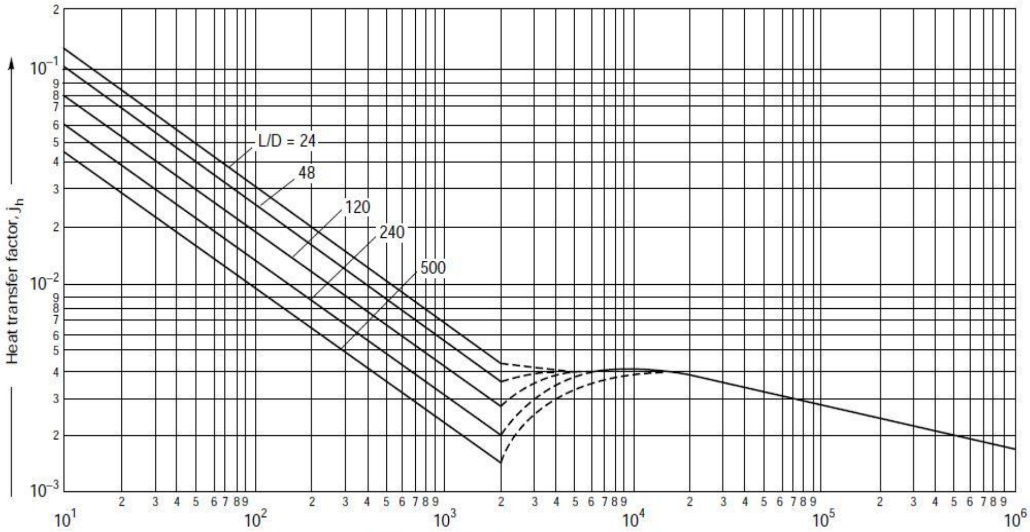


Figure 25 J_h factor graph for tube side

L/d = 146.4

From figure, heat transfer factor (J_h) = 0.003

$$\frac{h_i}{\theta_t} = \frac{J_h \text{ x Re x Pr}^{0.33} \text{ x K}_f}{\text{d}_\text{i}} = \frac{0.003 \text{ x } 620562.212 \text{ x } (0.337)^{0.33} \text{ x } 0.1287}{0.025}$$

$\frac{h_i}{\theta_t}$ = 6854 W/m^2.^{o}C

Mean Wall Temperature θ_t:

$(\frac{\mu}{\mu_w})^{0.14}$ = 1

h$_i$ = 6854 W/m^2.^{o}C

Overall coefficient:

$$h_{io} = h_i \text{ x } \frac{\text{ID}}{\text{OD}} = \text{ x } \frac{0.025}{0.03}$$

h_{io} =5711 W/m^2. ^{o}C

$$\text{U}_\text{c} = \frac{\text{hio x ho}}{\text{hio+ ho}} = 4760 \text{ W/m}^2.\ ^o\text{C}$$

For R$_\text{d}$ = 0.003

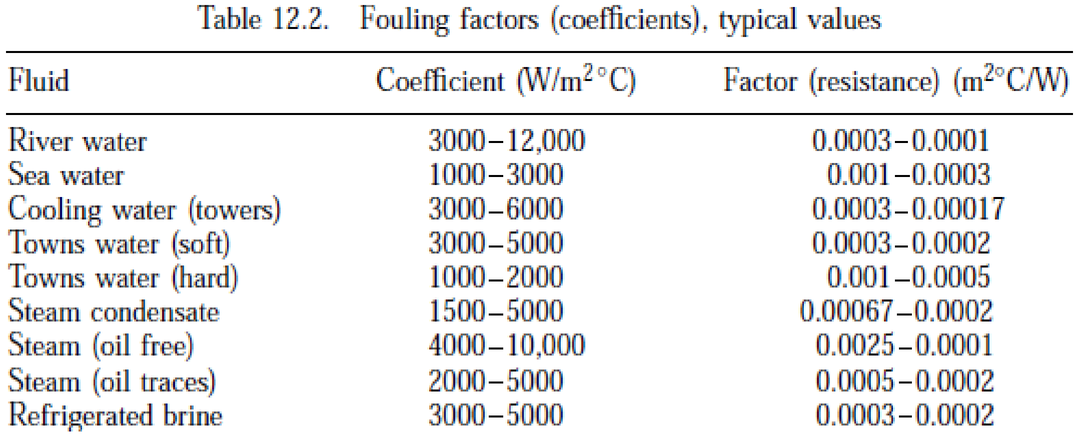


Table 12.2. Fouling factors (coefficients), typical values

Fluid	Coefficient (W/m^2°C)	Factor (resistance) (m^2°C/W)
River water	3000–12,000	0.0003–0.0001
Sea water	1000–3000	0.001–0.0003
Cooling water (towers)	3000–6000	0.0003–0.00017
Towns water (soft)	3000–5000	0.0003–0.0002
Towns water (hard)	1000–2000	0.001–0.0005
Steam condensate	1500–5000	0.00067–0.0002
Steam (oil free)	4000–10,000	0.0025–0.0001
Steam (oil traces)	2000–5000	0.0005–0.0002
Refrigerated brine	3000–5000	0.0003–0.0002

Table 31 Fouling factor table

$$\text{R}_\text{d} = \frac{\text{Uc}-\text{UD}}{\text{UD}*\text{Uc}}$$

U$_\text{D}$ = 302 W/m^2. ^{o}C

Since,

$$\frac{\text{U}_{\text{cal}} - \text{U}_{\text{ass}}}{\text{U}_{\text{ass}}} \text{ x } 100 = 7\ \% < 30\ \%$$

Hence the design is acceptable.

Pressure drop Calculation:

Shell side

$$\Delta P_s = 8 \text{ x } J_f \text{ x } \left(\frac{Ds}{d_e}\right)\left(\frac{L}{L_B}\right)\frac{\rho\mu^2}{2}\left(\frac{\mu}{\mu_w}\right)^{-0.14}$$

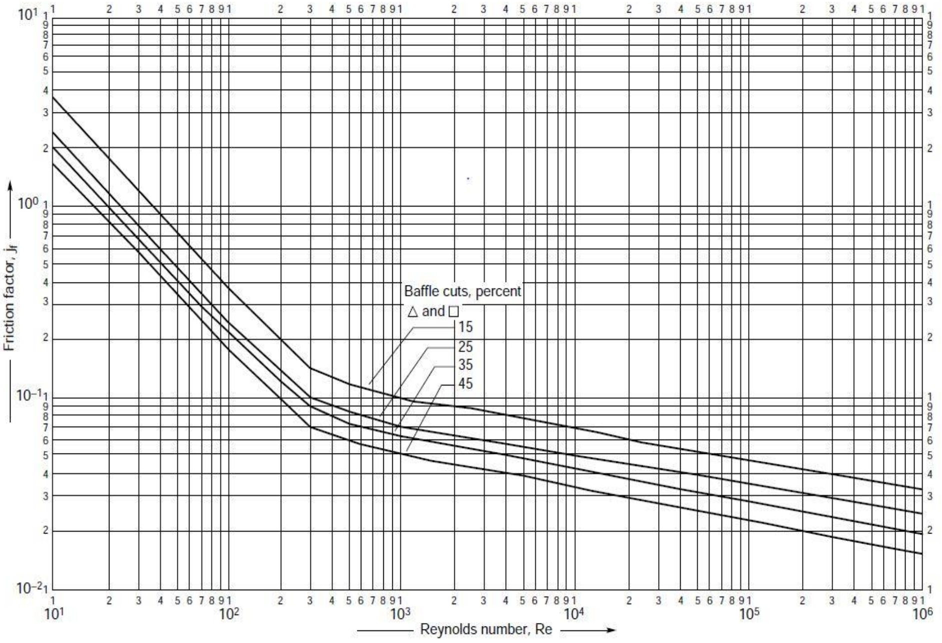


Figure 27 J_f factor graph

From figure, $J_f = 0.051$

$$= 8 \text{ x } 0.051 \text{ x } \left(\frac{0.9314}{0.021}\right)\left(\frac{3.66}{0.4657}\right)\left(\frac{997 \text{ x } 364.5}{997}\right)^2$$

ΔP_s = 2274 Pa = 0.32psi

Tube side

$$\Delta P_t = N_p \left[8 \text{ x } J_h \text{ x } \left(\frac{L}{d_i}\right)\left(\frac{\mu}{\mu_w}\right)^{-m} + 2.5\right] \text{ x } \frac{\rho {u_t}^2}{2}$$

$$= 2[8 \text{ x } 0.003 \text{ x } (387.72) + 2.5] \text{ x } \frac{8.8\ (61.21)^2}{2}$$

ΔPt = 38726.2 Pa = 5.6 psi

7.2.2 E-7 (Shell and Tube Heat Exchanger)

Water (Shell side): Process gas (Tube side):

$K = 0.6108$ W/m.°C

$\mu = 8.9 \times 10^{-4}$ Ns/m^2

$C_p = 4.312$ kJ/kg. °C

$\rho = 1000$ kg/m^3

$\dot{m} = 90.6$kg/s

$K = 0.1454$ W/m.°C

$\mu = 1.27 \times 10^{-5}$ Ns/m^2

$C_p = 3.967$ kJ/kg. °C

$\dot{m} = 14.94$ kg/s

$\rho = 12.99$kg/m^3

$T_{hi} = 115.4$°C ; $T_{ho} = 49$ °C

$T_{ci} = 30$ °C ; $T_{co} = 40$ °C

Heat duty:

$Q = \dot{m} \times C_p \times \Delta T$

$= 90.6 \times 4.312 \times 10$

$= 3805$ kW

Flow rate of Water:

$Q = \dot{m} \times C_p \times \Delta T \rightarrow \dot{m} = \frac{Q}{\mathrm{Cp} \times \Delta \mathrm{T}}$

$\dot{m} = \frac{3805}{4.312 \times 10}$

$\dot{m} = 90.6$kg/s

Log Mean Temperature Difference (LMTD):

$\mathrm{LMTD} = \frac{\Delta T2 - \Delta T1}{\ln\left(\frac{\Delta T2}{\Delta T1}\right)}$

ΔT_2 = Temperature difference of hot fluids

ΔT_1 = Temperature difference of cold fluids

$\mathrm{LMTD} = \frac{75.4 - 19}{\ln(75.4/19)}$

LMTD = 40.91 °C

$R = \frac{T_1 - T_2}{t_2 - t_1}$

$S = \frac{t_2 - t_1}{t_1 - t_1}$

$R = \frac{122-70}{30-20}$ = 6.64

$S = \frac{30-20}{122-20}$ = 0.13

From graph,1 shell and 4 tube passes

F_t = 0.98

So, corrected LMTD (Δt) = 0.98 x40.19 = 40.09 °C

U = 300 W/m^2 °C from above table

$A = \frac{3805\ 7}{300 \times 40.0}$ =316.3m^2

0.03m Outer diameter and 0.025m Internal diameter – BWG 17

Length of tube = L = 3.046

Area of tube = π x= 3.046x0.03 = 0.2869 m^2

Number of tubes required = $\frac{316.3}{0.2869}$ = 1102 tubes

Bundle diameter = D_b = d.($\frac{N_s}{K}$)$^{0.25}$ = 0.838 mm

From graph Clearance = 0.096

Shell diameter = 0.838 +0.096 = 0.96m

Shell side co-efficient

Baffle spacing (l_B) = 0.5 x D_s = 0.467 m

P_t = 1.25 x d_o = 0.0375m

$A_s = \frac{(P_t - d_0) \times D_s}{P_t}$ change

A_s = 0.18628 m^2

Mass velocity (Ga) = $\frac{\omega}{A_s} = \frac{90.97}{0.1868}$ = 485.01 kg/m^{2}s

$D_e = \frac{1.10}{d_0}(P_t^2 - 0.917 d_0^2)$

$D_e = 0.021$

$Re_b = 12731$

From figure, $J_h = 0.007$

$Pr = \frac{\mu \text{ x } Cp}{K} = 5.6$

$\frac{h_o}{\theta_s} = \frac{J_h \text{ x Re x } Pr^{0.33} \text{ x } K_f}{De}$ = 4720 W/m^2. ^{o}C $\theta_s = 1$ (6.26)

h_o = 4720 W/m^2. ^{o}C

Tube side co-efficient

Tube cross section area= $\frac{\pi}{4} d_i^2$

= 0.0005 m^2

Tube per pass = $\frac{1102}{4}$ = 275.586 tubes

Total flow area = 275.586 x 0.0005 = 0.1377 m^2

Mass velocity (Ga) = $\frac{\omega}{A} = \frac{14.94}{0.1377}$ = 538.73 kg/m^{2}s

u = $\frac{108.52}{12.99}$ = 8.52 m/s

Re= $\frac{Di \text{ x } Ga}{\mu}$ = 216923.84

$Pr = \frac{\mu \text{ x } Cp}{K} = \frac{1.22x10-5(\,3.967x10-3)}{0.1454}$ = 0.348

L/d = 121.84

From figure, heat transfer factor (J_h) = 0.0038

$\frac{h_i}{\theta_t} = \frac{J_h \text{ x Re x } Pr^{0.33} \text{ x } K_f}{d_i}$ = 3385 W/m^2.^{o}C

$\frac{h_i}{\theta_t}$ = 3385 W/m^2.^{o}C

Mean Wall Temperature θ_t:

$(\frac{\mu}{\mu_w})^{0.14} = 1$

h_i = 3385 W/m^2.^{o}C

Overall Co-efficient:

$\boldsymbol{h_{io}} = \boldsymbol{h_i} \times \frac{\mathbf{ID}}{\mathbf{OD}} = \times \frac{\mathbf{0.025}}{\mathbf{0.03}}$

h_{io} =2820 W/m^2. ^{o}C

$U_c = \frac{\text{hio x ho}}{\text{hio+ ho}}$ = 1765 W/m^2. ^{o}C

For R_d = 0.003

$R_d = \frac{Uc-UD}{UD*Uc}$

U_D = 280.38 W/m^2. ^{o}C

Since,

$\frac{U_{cal} - U_{ass}}{U_{ass}}$ x 100 = 7 % < 30 %

Acceptable design is achieved.

Pressure drop:

Shell side

$\Delta P_s = 8 \times J_f \times \left(\frac{Ds}{d_e}\right) \left(\frac{L}{L_B}\right) \frac{\rho\mu^2}{2} (\frac{\mu}{\mu_w})^{-0.14}$

From figure, J_f = 0.007

$= 8 \times 0.007 \times (\frac{0.9314}{0.021})(\frac{3.066}{0.467})(997)\ (\frac{485.5}{997})^2$

ΔP_s = 88.25 Pa

Tube side

$\Delta P_t = N_p\ [8 \times J_h \times \left(\frac{L}{d_i}\right) (\frac{\mu}{\mu_w})^{-m} + 2.5] \times \frac{\rho {u_t}^2}{2}$

$= 2[8 \times 0.0038 \times (121.84) + 2.5] \times \frac{12.7\ (8.531)^2}{2}$

$\Delta Pt = 5863$ Pa

7.2.3 E-5 (Shell and Tube Heat Exchanger)

Water (Shell side):	Process gas (Tube side):
$K = 0.6108$ W/m.°C	$K = 0.1408$ W/m.°C
$\mu = 8.9 \times 10^{-4}$ Ns/m^2	$\mu = 1.236 \times 10^{-5}$ Ns/m^2
$C_p = 4.312$ kJ/kg. °C	$C_p = 3.913$ kJ/kg. °C
$\rho = 1000$ kg/m^3	$\dot{m} = 20.61$ kg/s
$\dot{m} = 79.36$ kg/s	$\rho = 12.1$kg/m^3

$T_{hi} = 105$°C ; $T_{ho} = 36$ °C

$T_{ci} = 60$ °C ; $T_{co} = 70$ °C

Heat duty:

$Q = \dot{m} \times C_p \times \Delta T$

$= 20.61 \times 3.913 \times 69$

$= 3566$ kW

Flow rate of Water:

$Q = \dot{m} \times C_p \times \Delta T \rightarrow \dot{m} = \frac{Q}{Cp \times \Delta T}$

$\dot{m} = \frac{3566}{4.312 \times 10}$

$\dot{m} = 79.36$ kg/s

Log Mean Temperature Difference (LMTD):

$LMTD = \frac{\Delta T2 - \Delta T1}{\ln\left(\frac{\Delta T2}{\Delta T1}\right)}$

ΔT_2 = Temperature difference of hot fluids

ΔT_1 = Temperature difference of cold fluids

LMTD = $\frac{69-10}{\ln(69/10)}$

LMTD = 29.40 ^oC

R = $\frac{T_1 - T_2}{t_2 - t_1}$

S = $\frac{t_2 - t_1}{t_1 - t_1}$

R = $\frac{105-36}{70-60}$ = 6.9

S = $\frac{70-60}{105-60}$ = 0.22

From graph, 1 shell and 2 tube passes

F_t = 0.96

So, corrected LMTD (Δt) = 0.96 x29.40= 28.22 ^oC

From figure, U = 300 W/m^2 ^oC

A = $\frac{3566}{300 \times 28.22}$ = 271m^2

0.019m Outer diameter and 0.016m Internal diameter – BWG 17

Length of tube = L = 6.096

Area of tube = π x6.096x0.019 = 0.3637 m^2

Number of tubes required = $\frac{271}{0.3637}$ = 745 tubes

Bundle diameter = D_b = d.$(\frac{N_s}{K})^{0.25}$ = 0.714 mm

From graph Clearance = 0.092

Shell diameter = 0.714 +0.092 = 0.806m

Shell side co-efficient

Baffle spacing (l_B) = 0.5 x D_s = 0.403 m

P_t = 1.25 x d_o = 0.02375m

$A_s = \frac{(P_{t-}\ d_0)\ x\ D_s}{P_t}$ change

$A_s = 0.1612\ m^2$

Mass velocity (Ga) = $\frac{\omega}{A_s} = \frac{79.37}{0.1612} = 492.301\ kg/m^2s$

$D_e = \frac{1.10}{d_0}(P_t^2 - 0.917d_0^2)$

$D_e = 0.0134$

$Re_b = 8302$

From figure, $J_h = 0.0003$

$Pr = \frac{\mu\ x\ Cp}{K} = 1.49$

$\frac{h_o}{\theta_s} = \frac{J_h\ x\ Re\ x\ Pr^{0.33}\ x\ K_f}{De}$ = 1281 W/m². °C $\theta_s = 1$ (6.26)

h_o = 1281 W/m². °C

Tube side co-efficient

Tube cross section area= $\frac{\pi}{4}d_i^2$

= 0.000201 m²

Tube per pass = $\frac{745}{2}$ = 372.56 tubes

Total flow area = 372.57 x 0.000201 = 0.074574 m²

Mass velocity (Ga) = $\frac{\omega}{A} = \frac{20.61}{0.03744} = 276\ kg/m^2s$

$u = \frac{276}{12.1} = 22.802\ m/s$

$Re = \frac{Di\ x\ Ga}{\mu} = 357267.18$

$Pr = \frac{\mu\ x\ Cp}{K} = \frac{1.22x10-5(\ 3.967x10-3)}{0.1454} = 0.3418$

L/d = 381

From figure, heat transfer factor (J_h) = 0.0035

$\frac{h_i}{\theta_t} = \frac{J_h \text{ x Re x Pr}^{0.33} \text{ x K}_f}{d_i} = 7721 \text{ W/m}^2.^oC$

$\frac{h_i}{\theta_t} = 7721 \text{ W/m}^2.^oC$

Mean Wall Temperature θ_t:

$(\frac{\mu}{\mu_w})^{0.14} = 1$

$h_i = 7721 \text{ W/m}^2.^oC$

Overall Co-efficient:

$h_{io} = h_i \text{ x } \frac{\text{ID}}{\text{OD}} = 7721 \text{ x } \frac{0.016}{0.019}$

$h_{io} = 6501 \text{ W/m}^2.\ ^oC$

$U_c = \frac{\text{hio x ho}}{\text{hio+ ho}} = 1124 \text{ W/m}^2.\ ^oC$

For R_d = 0.003 from table

$R_d = \frac{Uc-UD}{UD*Uc}$

$U_D = 312 \text{ W/m}^2.\ ^oC$

Since,

$\frac{U_{cal} - U_{ass}}{U_{ass}} \text{ x } 100 = 14.3\ \% < 30\ \%$

Acceptable design

Pressure drop:

Shell side

$\Delta P_s = 8 \text{ x } J_f \text{ x } \left(\frac{Ds}{d_e}\right) \left(\frac{L}{L_B}\right) \frac{\rho\mu^2}{2} (\frac{\mu}{\mu_w})^{-0.14}$

From figure, J_f = 0.003

$= 8 \text{ x } 0.003 \text{ x } (\frac{0.806}{0.034})(\frac{6.096}{0.403})(\frac{(997) \text{ x } 492}{997})^2$

$\Delta P_s = 5301 \text{ Pa}$

Tube side

$\Delta P_t = N_p\ [8 \times J_h \times \left(\frac{L}{d_i}\right)(\frac{\mu}{\mu_w})^{-m} + 2.5] \times \frac{\rho {u_t}^2}{2}$

$= 2[8 \times 0.0035 \times (381) + 2.5] \times \frac{12.7\ (22)^2}{2}$

$\Delta Pt = 77117$ Pa

7.2.4 E-3 (Shell and Tube Heat Exchanger)

Water (Shell side):	Process gas (Tube side):
$K = 0.6108$ W/m.^{o}C	$K = 0.1408$ W/m.^{o}C
$\mu = 8.9 \times 10^{-4}$ Ns/m^2	$\mu = 1.236 \times 10^{-5}$ Ns/m^2
$C_p = 4.312$ kJ/kg. ^{o}C	$C_p = 3.913$ kJ/kg. ^{o}C
$\rho = 1000$ kg/m^3	$\dot{m} = 22.3$ kg/s
$\dot{m} = 67.86$ kg/s	$\rho = 12.1$kg/m^3

$T_{hi} = 328^{o}C$; $T_{ho} = 203\ ^{o}C$

$T_{ci} = 30\ ^{o}C$; $T_{co} = 77\ ^{o}C$

Heat duty:

$Q = \dot{m} \times C_p \times \Delta T$

$= 67.86 \times 3.913 \times 45$

$= 15000$ kW

Flow rate of Water

$Q = \dot{m} \times C_p \times \Delta T \rightarrow \dot{m} = \frac{Q}{Cp \times \Delta T}$

$\dot{m} = \frac{15000}{4.312 \times 45}$

$\dot{m} = 67.86$ kg/s

Log Mean Temperature Difference (LMTD):

LMTD = $\frac{\Delta T2 - \Delta T1}{\ln(\frac{\Delta T2}{\Delta T1})}$ ΔT_2 = Temperature difference of hot fluids

ΔT_1 = Temperature difference of cold fluids

LMTD = $\frac{125 - 45}{\ln(125/45)}$

LMTD = 78.30 °C

R = $\frac{T_1 - T_2}{t_2 - t_1}$

S = $\frac{t_2 - t_1}{t_1 - t_1}$

R = $\frac{328-203}{75-30}$ = 2.77

S = $\frac{75-30}{328-30}$ = 0.15

From graph, 1 shell and 2 tube passes

F_t = 0.98

So, corrected LMTD (Δt) = 0.96 x78.03= 77.281 °C

U = 300 W/m^2 °C from above table

A = $\frac{1500000}{300 \text{ x } 77.22}$ = 646.98m^2

0.003m Outer diameter and 0.025m Internal diameter – BWG 17

Length of tube = L = 3.66

Area of tube = π x3.66x0.03= 0.345 m^2

Number of tubes required = $\frac{646.98}{0..345}$ =1875 tubes

Bundle diameter = D_b = d.$(\frac{N_S}{K})^{0.25}$ = 1.72 m

From graph Clearance = 0.95

Shell diameter = 1.72 +0.95 = 2.76m

Shell side co-efficient:

Baffle spacing (l_B) = 0.5 x D_s = 1.335 m

P_t = 1.25 x d_o = 0.0375m

$A_s = \frac{(P_{t-}\ d_0) \text{ x } D_s}{P_t}$ change

A_s = 0.534 m^2

Mass velocity (Ga) = $\frac{\omega}{A_s} = \frac{22.5412}{0.534}$ = 127 kg/m^2s

$D_e = \frac{1.10}{d_0}$ (P_t^2 – 0.917d_0^2)

D_e = 0.02123

Re_b = 3039

From figure, J_h = 0.0007

$Pr = \frac{\mu \text{ x } Cp}{K}$ = 0.00621

$\frac{h_o}{\theta_s} = \frac{J_h \text{ x Re x } Pr^{0.33} \text{ x } K_f}{De}$ = 9155.38 W/m^2. °C $\theta_s = 1$

h_o =9155.38 W/m^2. °C

Tube side co-efficient:

Tube cross section area= $\frac{\pi}{4}d_i^2$

= 0.0005 m^2

Tube per pass = $\frac{1875}{2}$ =937.56 tubes

Total flow area = 937.56 x 0.0005 = 0.4687 m^2

Mass velocity (Ga) = $\frac{\omega}{A} = \frac{52.19}{0.467}$ = 111.35kg/m^2s

u = $\frac{52.196}{12.6}$ = 73.7 m/s

$Re= \frac{Di \text{ x } Ga}{\mu}$ = 1360000

$Pr = \frac{\mu \text{ x } Cp}{K}$ = 0.0556

L/d = 146

From figure, heat transfer factor (J_h) = 0.0035

$$\frac{h_i}{\theta_t} = \frac{J_h \text{ x Re x Pr}^{0.33} \text{ x } K_f}{d_i} = 3680 \text{ W/m}^2.^{o}\text{C}$$

$$\frac{h_i}{\theta_t} = 3680 \text{ W/m}^2.^{o}\text{C}$$

Mean Wall Temperature $\boldsymbol{\theta_t}$:

$$(\frac{\mu}{\mu_w})^{0.14} = 1$$

h_i = 3680 W/m^2.^{o}C

Overall coefficient:

$$h_{io} = h_i \text{ x } \frac{\text{ID}}{\text{OD}} = 3680 \text{ x } \frac{0.025}{0.03}$$

$$h_{io} = 3066 \text{ W/m}^2.\ ^{o}\text{C}$$

$$U_c = \frac{\text{hio x ho}}{\text{hio+ ho}} = 2297 \text{ W/m}^2.\ ^{o}\text{C}$$

For R_d = 0.003 from table

$$R_d = \frac{\text{Uc}-\text{UD}}{\text{UD}*\text{Uc}}$$

U_D = 291 W/m^2. ^{o}C

Since,

$$\frac{U_{cal} - U_{ass}}{U_{ass}} \text{ x } 100 = 3{,}06\ \% < 30\ \%$$

Acceptable design

Pressure drop:

Shell side

$$\Delta P_s = 8 \text{ x } J_f \text{ x } \left(\frac{Ds}{d_e}\right) \left(\frac{L}{L_B}\right) \frac{\rho\mu^2}{2} (\frac{\mu}{\mu_w})^{-0.14}$$

From figure, J_f = 0.005

$= 8 \times 0.005 \times (\frac{2.67}{0.0213})(\frac{3.66}{1.335})(\frac{1000) \times 492}{1000})^2$

$\Delta P_s = 27000$ Pa

Tube side

$\Delta P_t = N_p\ [8 \times J_h \times \left(\frac{L}{d_i}\right)(\frac{\mu}{\mu_w})^{-m} + 2.5] \times \frac{\rho {u_t}^2}{2}$

$= 2[8 \times 0.0017 \times (146) + 2.5] \times \frac{8.11\ (13.72)^2}{2}$

$\Delta Pt = 68.5k$ Pa

7.2.5 E-10 (Shell and Tube Heat Exchanger)

Process Gas (Shell side):	Process gas (Tube side):
K = 0.1452 W/m.°C	K = 0.1408 W/m.°C
$\mu = 1.204 \times 10^{-5}$ Ns/m^2	$\mu = 1.236 \times 10^{-5}$ Ns/m^2
C_p = 4.312 kJ/kg. °C	C_p = 3.913 kJ/kg. °C
ρ = 8.1 kg/m^3	ρ = 12.1kg/m^3

$T_{hi} = 346°C$; $T_{ho} = 223\ °C$

$T_{ci} = 44\ °C$; $T_{co} = 164\ °C$

Heat duty:

$Q = \dot{m} \times C_p \times \Delta T$

$= 31.9 \times 3.913 \times 123$

$= 11900$ kW

Log Mean Temperature Difference (LMTD):

$LMTD = \frac{\Delta T2 - \Delta T1}{\ln(\frac{\Delta T2}{\Delta T1})}$ ΔT_2 = Temperature difference of hot fluids

ΔT_1 = Temperature difference of cold fluids

LMTD = $\frac{123-120}{\ln(123/120)}$

LMTD = 121.49 °C

R = $\frac{T_1 - T_2}{t_2 - t_1}$

S = $\frac{t_2 - t_1}{t_1 - t_1}$

R = $\frac{346-223}{164-44}$ = 1.023

S = $\frac{164-44}{346-44}$ = 0.39

From graph, 1 shell and 2 tube passes

F_t = 0.96

So, corrected LMTD (Δt) = 0.96 x121.49= 116.63 °C

From figure, U = 300 W/m^2 °C from above table

A = $\frac{1190000}{300 \text{ x } 116.22}$ = 102.03m^2

0.025m Outer diameter and 0.022m Internal diameter – BWG 17

Length of tube = L = 4m

Area of tube = π x4x0.025 = 0.191 m^2

Number of tubes required = $\frac{102.03}{0.19}$ = 534 tubes

Bundle diameter = D_b = d.($\frac{N_s}{K}$)$^{0.25}$ = 0.8074 mm

From graph Clearance = 0.089

Shell diameter = 0.8074 +0.089 = 0.8969m

Shell side co-efficient:

Baffle spacing (l_B) = 0.5 x D_s = 0.445 m

$P_t = 1.25 \text{ x } d_o = 0.031255\text{m}$

$A_s = \frac{(P_t - d_0) \text{ x } D_s}{P_t}$ change

$A_s = 0.178 \text{ m}^2$

Mass velocity (Ga) = $\frac{\omega}{A_s} = \frac{85}{0.179} = 477\text{kg/m}^2\text{s}$

$D_e = \frac{1.10}{d_0}(P_t^2 - 0.917 d_0^2)$

$D_e = 0.0177$

$Re_b = 4036$

From figure, $J_h = 0.0021$

$Pr = \frac{\mu \text{ x } Cp}{K} = 0.0031$

$\frac{h_o}{\theta_s} = \frac{J_h \text{ x } Re \text{ x } Pr^{0.33} \text{ x } K_f}{De} = 1449 \text{ W/m}^2.\,^{o}\text{C}$ $\theta_s = 1$

$h_o = 1449 \text{ W/m}^2.\,^{o}\text{C}$

Tube side co-efficient:

Tube cross section area= $\frac{\pi}{4} d_i^2 = 0.00038 \text{ m}^2$

Tube per pass = $\frac{534}{2} = 267$ tubes

Total flow area = 267 x 0.00038 = 0.1014m^2

Mass velocity (Ga) = $\frac{\omega}{A} = \frac{130.35}{0.1014} = 128.5 \text{ kg/m}^2\text{s}$

$u = \frac{27128{,}56}{8.11} = 15.802 \text{ m/s}$

$Re = \frac{Di \text{ x } Ga}{\mu} = 13523$

$Pr = \frac{\mu \text{ x } Cp}{K} = 0.0035$

L/d = 181

From figure, heat transfer factor (J_h) = 0.0035

$\frac{h_i}{\theta_t} = \frac{J_h \text{ x Re x Pr}^{0.33} \text{ x K}_\text{f}}{\text{d}_\text{i}}$ = 6869 W/m².°C

$\frac{h_i}{\theta_t}$ = 6869 W/m².°C

Mean Wall Temperature $\boldsymbol{\theta_t}$:

$(\frac{\mu}{\mu_w})^{0.14} = 1$

h_i = 6869W/m².°C

Overall Co-efficient

$h_{io} = h_i \text{ x } \frac{\text{ID}}{\text{OD}} = 6869 \text{ x } \frac{0.022}{0.025}$

h_{io} = 6044 W/m². °C

$U_c = \frac{\text{hio x ho}}{\text{hio+ ho}}$ = 1168 W/m². °C

For R_d = 0.003 from table

$R_d = \frac{\text{Uc}-\text{UD}}{\text{UD}*\text{Uc}}$

U_D = 259 W/m². °C

Since,

$\frac{U_{cal} - U_{ass}}{U_{ass}}$ x 100 = 13 % < 30 %

Acceptable design

Pressure drop

Shell side

$\Delta P_s = 8 \text{ x J}_\text{f} \text{ x } \left(\frac{Ds}{d_e}\right) \left(\frac{L}{L_B}\right) \frac{\rho\mu^2}{2} (\frac{\mu}{\mu_w})^{-0.14}$

From figure, J_f = 0.0021

$= 8 \text{ x } 0.003 \text{ x } (\frac{0.025}{0.0177})(\frac{4}{0.45})(\frac{(9.1) \text{ x } 5.8}{2})^2$

ΔP_s = 29.01k Pa

Tube side

ΔP_t = N_p $[8 \text{ x } J_h \text{ x } \left(\frac{L}{d_i}\right)(\frac{\mu}{\mu_w})^{-m} + 2.5] \text{ x } \frac{\rho {u_t}^2}{2}$

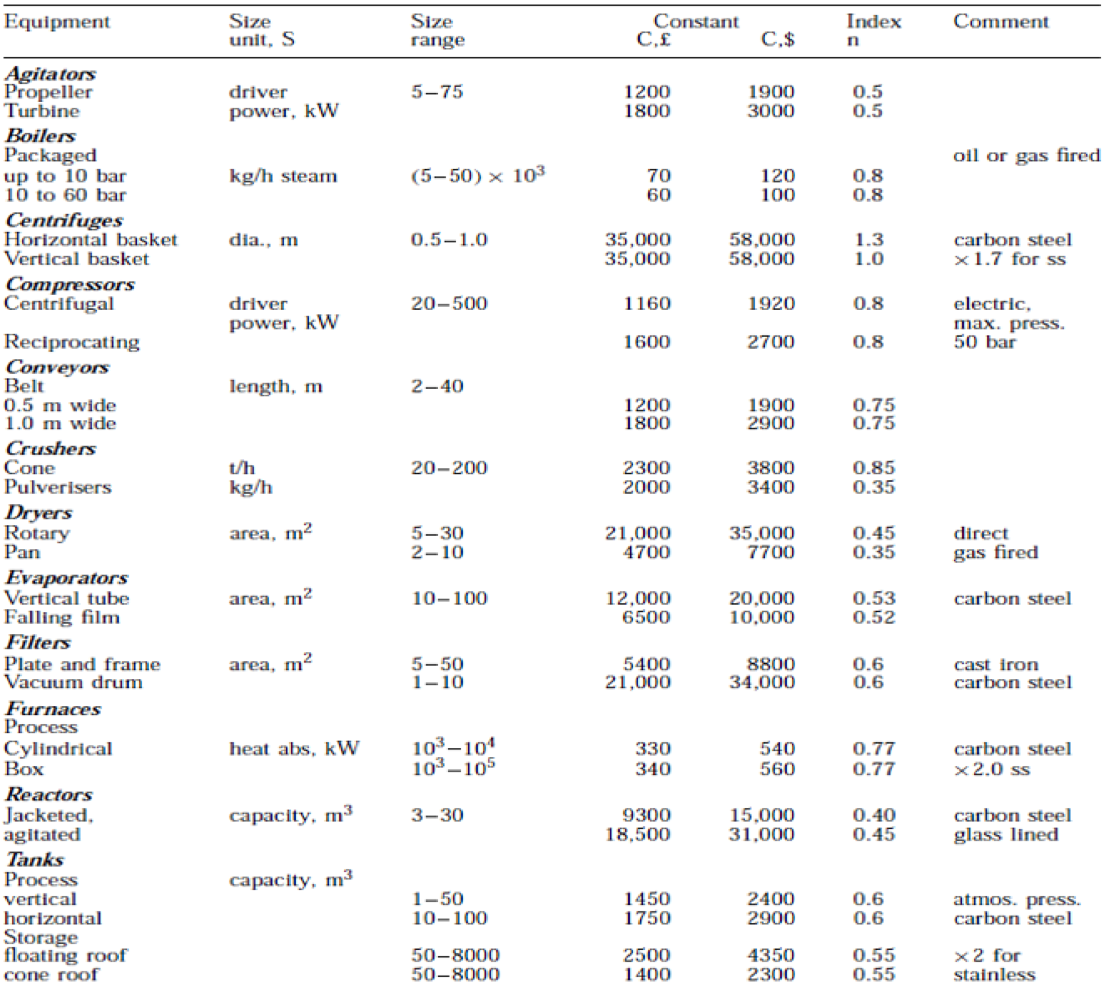
$= 2[8 \text{ x } 0.04 \text{ x } (181) + 2.5] \text{ x } \frac{8.1 \ (15.8)^2}{2}$

ΔPt =68.8k Pa

CHAPTER 8

COSTING AND HAZOP ANALYSIS

Costing is very important because after efficiency it is the economics that matter in designing a plant. The design Engineer must be able to make quick, fast costing estimates to decide between different processes and to choose between them. Chemical plants are made to build profits, so it is very important to do cost analysis before any plant can be placed. For calculating cost of reactors in plant following table was used:



Equipment	Size unit, S	Size range	Constant C,£	Constant C,$	Index n	Comment
Agitators						
Propeller	driver power, kW	5–75	1200	1900	0.5	
Turbine			1800	3000	0.5	
Boilers						
Packaged						oil or gas fired
up to 10 bar	kg/h steam	$(5-50)\times10^3$	70	120	0.8	
10 to 60 bar			60	100	0.8	
Centrifuges						
Horizontal basket	dia., m	0.5–1.0	35,000	58,000	1.3	carbon steel × 1.7 for ss
Vertical basket			35,000	58,000	1.0	
Compressors						
Centrifugal	driver power, kW	20–500	1160	1920	0.8	electric, max. press. 50 bar
Reciprocating			1600	2700	0.8	
Conveyors						
Belt	length, m	2–40				
0.5 m wide			1200	1900	0.75	
1.0 m wide			1800	2900	0.75	
Crushers						
Cone	t/h	20–200	2300	3800	0.85	
Pulverisers	kg/h		2000	3400	0.35	
Dryers						
Rotary	area, m^2	5–30	21,000	35,000	0.45	direct gas fired
Pan		2–10	4700	7700	0.35	
Evaporators						
Vertical tube	area, m^2	10–100	12,000	20,000	0.53	carbon steel
Falling film			6500	10,000	0.52	
Filters						
Plate and frame	area, m^2	5–50	5400	8800	0.6	cast iron
Vacuum drum		1–10	21,000	34,000	0.6	carbon steel
Furnaces						
Process						
Cylindrical	heat abs, kW	10^3-10^4	330	540	0.77	carbon steel × 2.0 ss
Box		10^3-10^5	340	560	0.77	
Reactors						
Jacketed, agitated	capacity, m^3	3–30	9300	15,000	0.40	carbon steel
			18,500	31,000	0.45	glass lined
Tanks						
Process	capacity, m^3					
vertical		1–50	1450	2400	0.6	atmos. press.
horizontal		10–100	1750	2900	0.6	carbon steel
Storage						
floating roof		50–8000	2500	4350	0.55	× 2 for stainless
cone roof		50–8000	1400	2300	0.55	

` *Table 32 Costing Index table*

8.1 Costing of Reactors:

8.1.1 Primary Reformer:

Q_{avg}= 65800lcal/hr

Q_{avg}= 7652241W/m^2

C = 540 n = 0.77 From Table

$Ce= CQ^n$

Ce = (540) $(765241)^{0.77}$

Ce= $ 3111128

Cost of Catalyst = $ 450/ft^2

Vc = 31.8m^3

Total Cost of catalyst = $6221731.9

Total Cost of primary reformer = $ 3111128 + $ 6221731.9

Total Cost of primary reformer = $ 3733302

8.1.2 Secondary Reformer:

Weight of catalyst = 32568 kg

Vc = 30.43 m^3

V_R = 48.64 m^3

Cost of catalyst = $ 373/ft^3

Total cost of catalyst = Cc x Vc

Total cost of catalyst = $ 493495

C = 540 n = 0.77 Q = 1.19x10^5 kW

$Ce = CQ^n$

Ce = $ 4370840

Total cost of Secondary Reformer = $ 4864335

8.1.3 High Temperature Shift:

Weight of catalyst = 66289.7 kg

Cost of catalyst = $ 21/kg

Total cost of catalyst = $ 1403425.59

$V_c = 87.22\ m^3$

$$C_e = CQ^n$$

Q = Vr = 87.22 m 3

C = 31000   n = 0.40 From Table

Total cost of HTS = $ 1588611

8.1.4 Low Temperature Shift:

Weight of catalyst = 91288 kg

Cost of catalyst = $ 32/kg

Total cost of catalyst = $ 2921216

$V_c = 87.22\ m^3$

$$C_e = CQ^n$$

Q = Vr = 111.67 m 3

C = 31000   n = 0.45 From Table

Cost of reactor = $ 258781.4

Total Cost of LTS = $ 31999.7

8.1.5 Methanator:

Weight of catalyst = 68426 kg

Cost of catalyst = $ 15/kg

Total cost of catalyst = $ 936392

$$C_e = CQ^n$$

Q = Vr = 48.68 m 3

C = 15000   n = 0.45 From Table

Cost of reactor = $ 86181

Total Cost of methanator = $ 1022574

8.1.6 Ammonia Reactor:

Weight of catalyst = 31535.47 kg

Cost of catalyst = $ 43/kg

Total cost of catalyst = $ 1356068

Vc = 25.2 m^3

$$C_e = CQ^n$$

Q = Vr = 60.54 m 3

C = 15000   n = 0.45 From Table

Cost of reactor = $ 5869420.9

Total Cost of Ammonia reactor = $ 6005488

8.2 Costing of Furnace:

Q = 2.77×10^5 kW

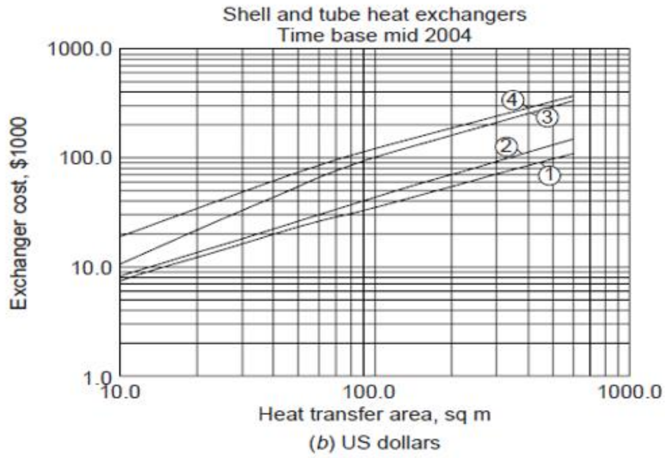
C = 560 n = 0.77

Ce = $560 \times (2.77 \times 10^6)^{0.77}$

Cost of furnace = $ 8687552

8.3 Costing of Heat Exchangers:

For costing of Heat Exchangers, the first step is to know the area of heat exchanger. By knowing the area, material of shell and tubes, pressure of heat exchanger and type of head of heat exchanger we can calculate the cost of heat exchanger in 2004



Materials	
Shell	**Tubes**
① Carbon steel	Carbon steel
② C.S.	Brass
③ C.S.	Stainless steel
④ S.S.	S.S.

Pressure factors	
1–10 bar	× 1.0
10–20	× 1.1
20–30	× 1.25
30–50	× 1.3
50–70	× 1.5

Type factors	
Floating head	× 1.0
Fixed tube sheet	× 0.8
U tube	× 0.85
Kettle	× 1.3

Figure 29 index graph (A) for costing of exchangers

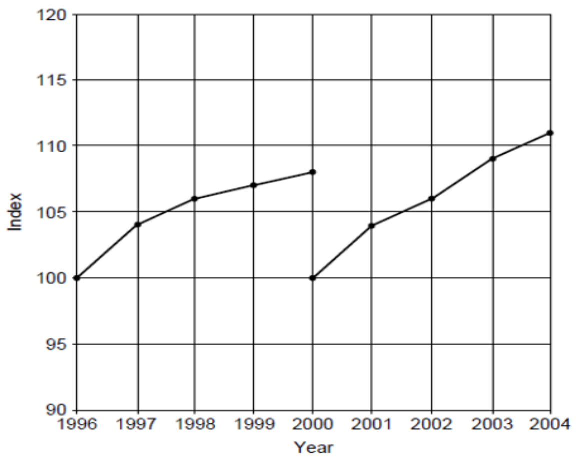


Figure 28 Cost indexing graph (B)

by using these graphs. We need calculate cost of heat exchanger for the 2020. For this purpose, graph (B) is used.

By using this formula, we can calculate the cost of heat exchanger in 2020.

$$\textbf{Cost in 2020} = \frac{\textbf{(Cost in 2004) (Index in 2020)}}{\textbf{(Index in 2004)}}$$

8.3.1 Exchanger E-1

Area = 336m^2

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Carbon steel shell and carbon steel tubes From Figure

Cost in 2004= (50,000) x (1.3) x (1) = $65000

Index in 2004 = 111 From Figure

Index in 2020 = 167

$$\text{Cost in 2020} = \frac{\text{(Cost in 2004) (Index in 2020)}}{\text{Index in 2004}}$$

Cost in 2020 = $97792

8.3.2 Exchanger E-2

Area = 133m^2

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Carbon steel shell and stainless-steel tubes From Figure

Cost in 2004= (60,000) x (1.25) x (0.85) = $63750

Index in 2004 = 111 From Figure

Index in 2020 = 167

$$\text{Cost in 2020} = \frac{\text{(Cost in 2004) (Index in 2020)}}{\text{Index in 2004}}$$

Cost in 2020 = $ 959125

8.3.3 Exchanger E-3

Area = 646m^2

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Carbon steel shell and carbon steel tubes From Figure

Cost in 2004= (105000) x (1.3) x (1) = $136500

Index in 2004 = 111 From Figure

Index in 2020 = 167

Cost in 2020 = (Cost in 2004) (Index in 2020)

Index in 2004

Cost in 2020 = $ 205346

8.3.4 Exchanger E-4

Area = 550 m^2

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Carbon steel shell and carbon steel tubes From Figure

Cost in 2004= (105000) x (1.3) x (1) = $130000

Index in 2004 = 111 From Figure

Index in 2020 = 167

Cost in 2020 = (Cost in 2004) (Index in 2020)

Index in 2004

Cost in 2020 = $ 197363

8.3.5 Exchanger E-5

Area = 271 m^2

Cost in 2004 = (Bare cost from figure) x (Pres. Factor) x (T.F)

Carbon steel shell and carbon steel tubes From Figure

Cost in 2004= (105000) x (1.3) x (1) = $104000

Index in 2004 = 111 From Figure

Index in 2020 = 167

Cost in 2020 = (Cost in 2004) (Index in 2020)

Index in 2004

Cost in 2020 = $ 156468

Similarly, the cost of remaining heat exchangers is calculated and given in the table below.

Exchangers	**Cost**
E-6	($)215144
E-7	($)195585
E-8	($)219055
E-9	($)234702
E-10	($)244481
E-11	($)260349
E-12	($)215144
E-13	($)191673

Table 33 Heat Exchangers Costs

8.4 Costing of Compressors:

For the costing of compressors, we need to know about the capacity of compressor. By using graph below with known capacity, we can calculate the purchased cost in 1998.

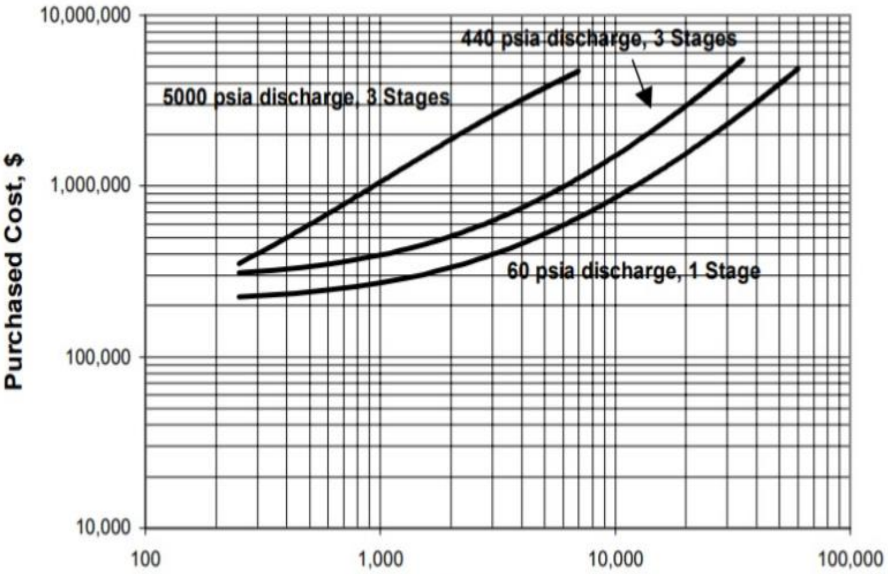


Figure 30 Compressor Costing graph

After calculating the purchase cost, we can find the cost index in 1998 and 2020.

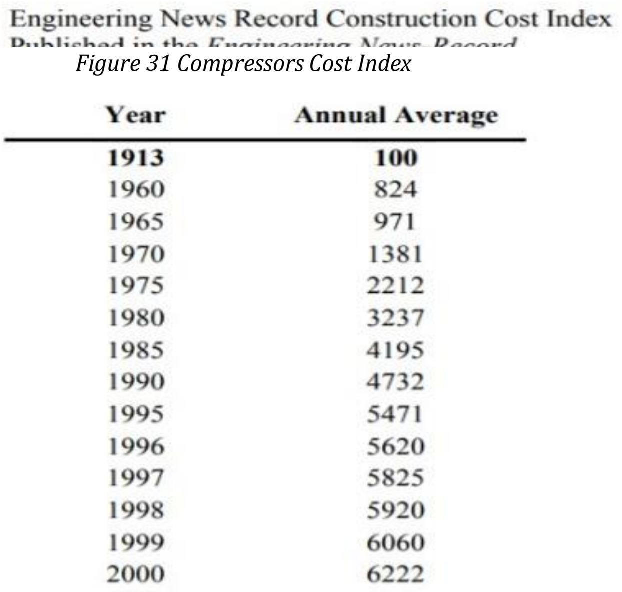
Using this formula, we the find the cost of compressor in 2020.

$$\text{Cost in 2020} = \frac{(\text{Cost in 1998})\ (\text{Index in 2020})}{\text{Index in 1998}}$$

8.4.1 Compressor C1:

Power = 1.36 MW

Capacity = 28439 ft^3/min



Engineering News Record Construction Cost Index
Published in the *Engineering News-Record*

Year	Annual Average
1913	**100**
1960	824
1965	971
1970	1381
1975	2212
1980	3237
1985	4195
1990	4732
1995	5471
1996	5620
1997	5825
1998	5920
1999	6060
2000	6222

Figure 31 Compressors Cost Index

Cost of compressor in 1998 = $1500000 From Figure

$$\text{Cost in 2020} = \frac{(\text{Cost in 1998})\ (\text{Index in 2020})}{\text{Index in 1998}}$$

Cost index in 1998 = 5920 From Figure

Cost index today = 6288

Cost in 2020 = 1500000 x 6288/5920

Cost in 2020 = $1593243

8.4.2 Compressor C2:

Power = 8.48 MW

Capacity =177331 ft^3/min

Cost of compressor in 1998 = $8000000 From Figure

Cost in 2020 = (Cost in 1998) (Index in 2020)

Index in 1998

Cost index in 1998 = 5920 From Figure

Cost index today = 6288

Cost in 2020 = 8000000 x 6288/5920

Cost in 2020 = $ 8497297

8.4.3 Compressor C:3

Power = 4.60 MW

Capacity = 98285 ft^3/min

Cost of compressor in 1998 = $ 500000 From Figure

Cost in 2020 = (Cost in 1998) (Index in 2020)

Index in 1998

Cost index in 1998 = 5920 From Figure

Cost index today = 6288

Cost in 2020 = 500000 x 6288/5920

Cost in 2020 = $ 5310810

8.4.4 Compressor C4:

Power = 9.9 MW

Capacity = 207026 ft^3/min

Cost of compressor in 1998 = $ 11000000 From Figure

Cost in 2020 = (Cost in 1998) (Index in 2020)

Index in 1998

Cost index in 1998 = 5920 From Figure

Cost index today = 6288

Cost in 2020 = 11000000 x 6288/5920

Cost in 2020 = $ 1163783.78

8.4.5 Compressor C5:

Power = 7.03 MW

Capacity = 147218 ft^3/min

Cost of compressor in 1998 = $ 7000000 From Figure

Cost in 2020 = (Cost in 1998) (Index in 2020)

Index in 1998

Cost index in 1998 = 5920 From Figure

Cost index today = 6288

Cost in 2020 = 7000000 x 6288/5920

Cost in 2020 = $ 7435135

8.6 Total Cost:

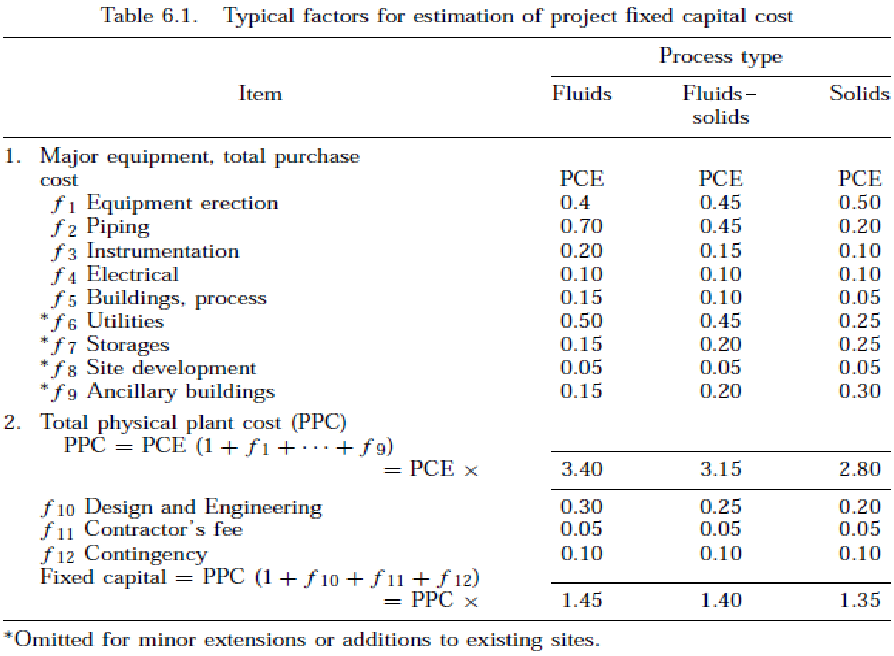


Table 6.1. Typical factors for estimation of project fixed capital cost

Item	Process type Fluids	 Fluids–solids	 Solids
1. Major equipment, total purchase cost	PCE	PCE	PCE
f_1 Equipment erection	0.4	0.45	0.50
f_2 Piping	0.70	0.45	0.20
f_3 Instrumentation	0.20	0.15	0.10
f_4 Electrical	0.10	0.10	0.10
f_5 Buildings, process	0.15	0.10	0.05
*f_6 Utilities	0.50	0.45	0.25
*f_7 Storages	0.15	0.20	0.25
*f_8 Site development	0.05	0.05	0.05
*f_9 Ancillary buildings	0.15	0.20	0.30
2. Total physical plant cost (PPC) PPC = PCE $(1 + f_1 + \cdots + f_9)$ = PCE ×	3.40	3.15	2.80
f_{10} Design and Engineering	0.30	0.25	0.20
f_{11} Contractor's fee	0.05	0.05	0.05
f_{12} Contingency	0.10	0.10	0.10
Fixed capital = PPC $(1 + f_{10} + f_{11} + f_{12})$ = PPC ×	1.45	1.40	1.35

*Omitted for minor extensions or additions to existing sites.

Figure 32 Capital cost estimation factors list

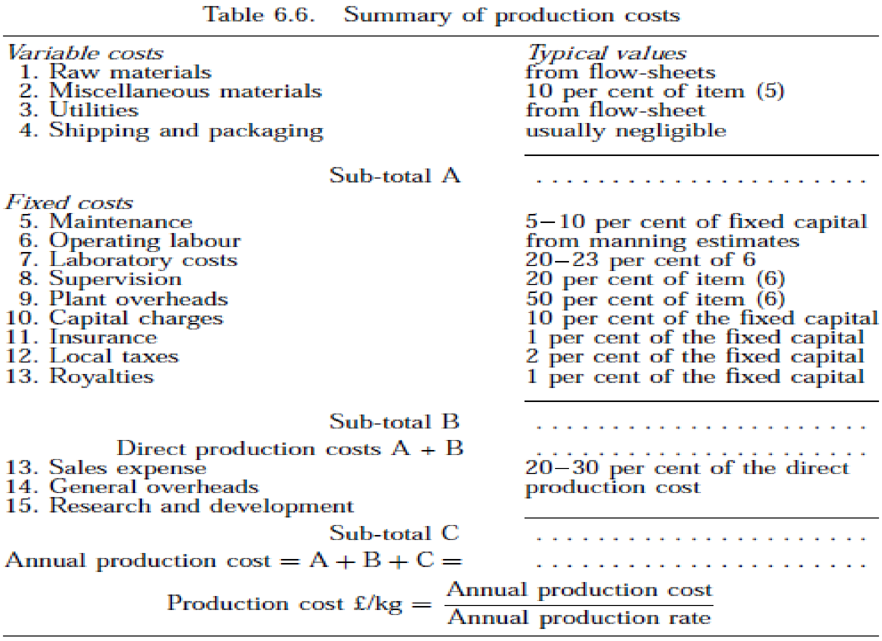


Table 6.6. Summary of production costs

Variable costs	*Typical values*
1. Raw materials	from flow-sheets
2. Miscellaneous materials	10 per cent of item (5)
3. Utilities	from flow-sheet
4. Shipping and packaging	usually negligible
Sub-total A	. . . .
Fixed costs	
5. Maintenance	5–10 per cent of fixed capital
6. Operating labour	from manning estimates
7. Laboratory costs	20–23 per cent of 6
8. Supervision	20 per cent of item (6)
9. Plant overheads	50 per cent of item (6)
10. Capital charges	10 per cent of the fixed capital
11. Insurance	1 per cent of the fixed capital
12. Local taxes	2 per cent of the fixed capital
13. Royalties	1 per cent of the fixed capital
Sub-total B	. . . .
Direct production costs A + B	. . . .
13. Sales expense 14. General overheads 15. Research and development	20–30 per cent of the direct production cost
Sub-total C	. . . .
Annual production cost = A + B + C =	. . . .

$$\text{Production cost £/kg} = \frac{\text{Annual production cost}}{\text{Annual production rate}}$$

Figure 33 Summary of production cost

Table 34 Physical Cost of Plant

PHYSICAL PLANT COST	
Items	**Cost ($ Million)**
Equipment Erection	0.4
Piping	0.70
Instrumentation	0.20
Electrical	0.10
Process	0.15
Storages	0.50
Purchase Cost Equipment (PCE)	87
Physical Plant Cost (PPC)	295

Table 33 Physical Cost of Plant

TOTAL CAPITAL COST	
Working Capital (WC)	29.09 $ Million
Total Capital	458 $ Million

Table 34 Total capital cost of the plant

FIXED CAPITAL	
Items	Cost ($ Million)
Design and Engineering	0.30
Contingencies	0.05
Contractor's fees	0.10
Fixed Capital (FC)	428

Table 35 Fixed capital Cost of the plant

FIXED OPERATING COST	
Items	**Cost ($ Million)**
Maintenance	32.9
Capital Charges	45.8
Plant Overhead	9.2
Local Taxes	4.5
Royalties	4.5
Fixed Operating Cost (FOC)	96.5

Table 36 Fixed Operating Cost of the Plant

VARIABLE OPERATING COST	
Items	**Cost ($ Million)**
Raw Material	.8
Shipping and packaging	Negligible
Variable Operating Cost (VOC)	.8

Table 37 Variable Operating Cost of the plant

TOTAL OPERATING COST	
Items	**Cost ($ Million)**
Direct Production Cost	97.29
Sales Expense	19.5
General Overheads	
Research & Development	
Total Operating Cost	116.3

Table 38 Total operating cost of the plant

REVENUE & NET PROFIT	
Cost of Urea	$ 27/kg
Per annum Urea Production	724 e+06 kg
Revenue	$145 Million
Net Profit	$ 32 Million

Table 39 Revenue and profits

8.7 HAZOP Analysis:

HAZOP analysis stands for hazard and operability analysis. This safety analysis finds its application in identifying potential risks and hazards surrounding an industrial process. HAZOP analysis deals to minimize accidents at plants that can lead to undesirable products. Following are the goals achieved by HAZOP hazard analysis:

1. Identification of hazard.
2. Extent of the damage.
3. Guidelines to solve the problem.

8.7.1 HAZOP Guide Words:

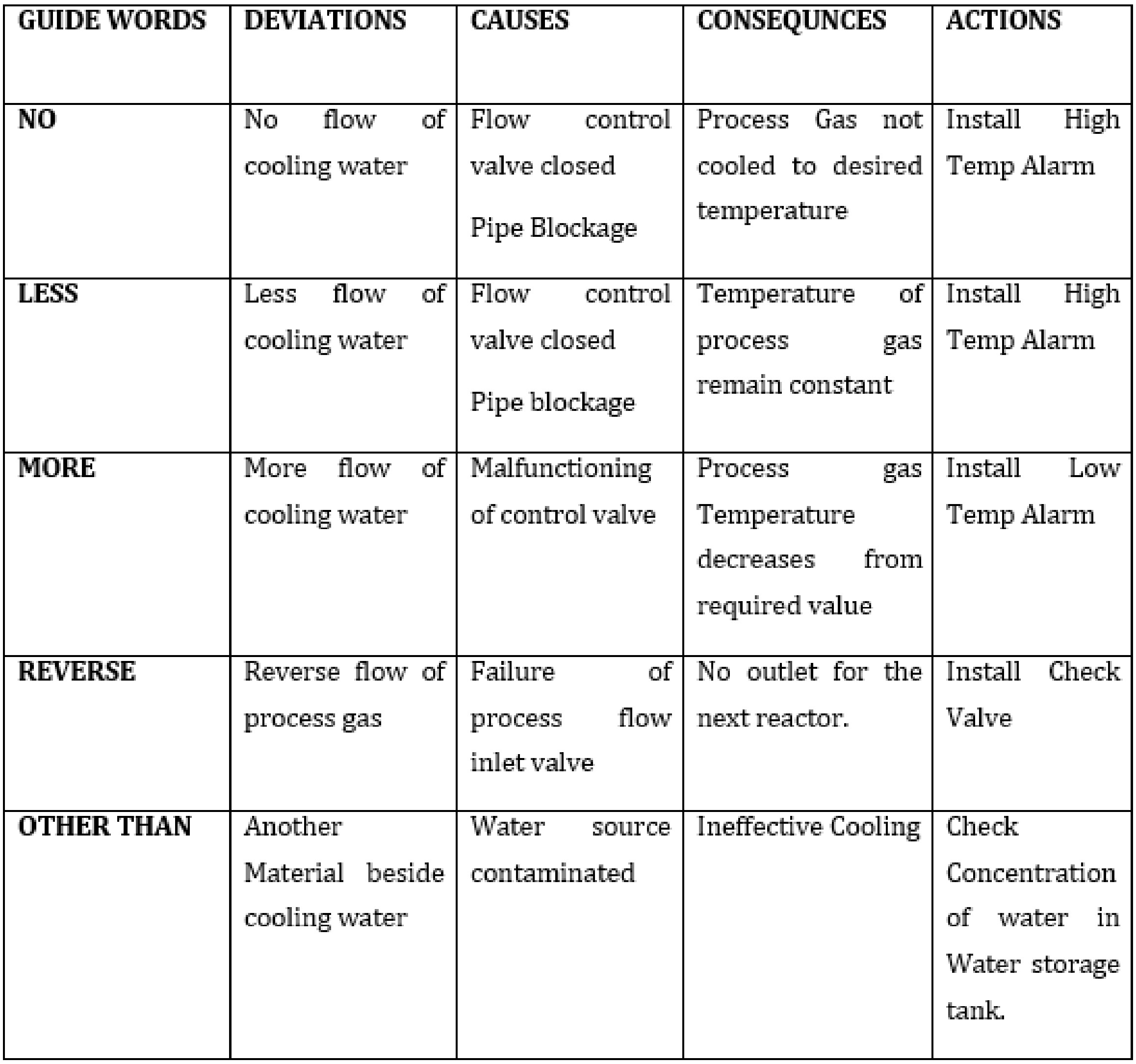
HAZOP utilizes a set of Guide words that identify probable deviations that can occur during a chemical process. These guide words prove their worth by allowing a systematic and consistent procedure through brainstorming. Some of these guide words which are selected by the risk management teams and are used for hazard management are as follows:

Sr. No.	Guide Word	Meaning
1	NO or NOT	Complete contradiction of original design
2	LESS	Decrease quantitatively
3	MORE	Increase quantitatively
4	AS WELL AS	Qualitative enhancement

5	REVERSE	Opposite to the original design
6	OTHER THAN/INSTEAD	Entire design
7	PART OF	Qualitative modification/decrease

Table 40 HAZOP guide words

8.7.2 HAZOP Analysis of heat exchanger:



GUIDE WORDS	**DEVIATIONS**	**CAUSES**	**CONSEQUNCES**	**ACTIONS**
NO	No flow of cooling water	Flow control valve closed Pipe Blockage	Process Gas not cooled to desired temperature	Install High Temp Alarm
LESS	Less flow of cooling water	Flow control valve closed Pipe blockage	Temperature of process gas remain constant	Install High Temp Alarm
MORE	More flow of cooling water	Malfunctioning of control valve	Process gas Temperature decreases from required value	Install Low Temp Alarm
REVERSE	Reverse flow of process gas	Failure of process flow inlet valve	No outlet for the next reactor.	Install Check Valve
OTHER THAN	Another Material beside cooling water	Water source contaminated	Ineffective Cooling	Check Concentration of water in Water storage tank.

Table 41 HAZOP analysis of heat exchanger

8.7.3 HAZOP analysis of secondary reformer:

GUIDE WORDS	DEVIATIONS	CAUSES	CONSEQUNCES	ACTIONS
NO	No Flow of reactant	Feed Line valve is closed	No reaction takes place	Open Valve To Allow Reactant into the reactor
		Feed Line is blocked	No reactant enters the reactor	Clear Blockage in feedline
Less	Less Flow of Reactant	Feed Line is loose.	Reactants falls out from field line before entering reactor	Join/Repair and seal Feed line.
	Less Temperature	Fuel Line is loose.	Incomplete Combustion will take place.	Repair and seal fuel line
	Less Pressure	Opening of PCV.	Required conversion is not achieved.	Install pressure Control Loop.
	Less Residence Time	Valve Opening Malfunction	Less optimum yield	Periodic maintenance or check up on the valve
More	More Pressure	Compressor Control Failure	Reactor operate at high Temperature than required.	Trigger automatic Shutdown.

	More Temperature	Very high amount of fuel and air flow inside the reactor.	Damage catalyst and equipment inside the reactor	Equip reactor system with fail open or fail close valve system.
As well as	Another Material beside Fuel.	Contamination of fuel storage tank.	Inadequate Combustion in reactor.	Check Composition of fuel in storage tank

Table 42 HAZOP analysis of secondary reformer

CHAPTER 9

INTRUMENTATION AND PROCESS CONTROL

Instrumentation is termed as the use of measuring instruments to look and control a process variable such as level, pressure, flow, temperature, speed. The process control is an emerging field of study in engineering which deals with designing of algorithms and devices for maintaining the output of a specific process with a required range. In process control, control loops and controllers are used to regulate a chemical process and improve effectiveness of the process [17]. Controller is a tool that collects data from a measurement device, contrasts that statistical information to a predetermined set point, and, if needed, suggests a control component to take remedial act. Following are the type of process variables:

1. Measured/controlled Variable-is the situation of the process fluid that must be kept at the specified set point.
2. Manipulated Variable- The factor that is changed to keep the process variable at set point.

Set point is the value for a process variable which is to be retained at a constant value.

9.1 Control loop and its components:

A control loop is the base on which the architect of process control systems that are applied in the industry. Physically a control loop consists of all the elements that are necessary to control and adjust the set point of the controlled variable.

Following are the components of the control loop:

1. Sensors- Device that can access the changes in condition and it is the first thing to measure the variable process in control loop.
2. Transducer- Device that translates a mechanical signal into an electrical signal.
3. Converters or relay- The conversion of one signal to another signal from a device
4. Transmitters- The conversion of reading from sensor to a signal.

5. Signals which are pneumatic, analog or digital.
6. Indicators-It is a device which is readable by humans and can show information like pressure gauge.
7. Records- To record the output from a device.

9.2 Control loop Applied on heat exchanger:

Temperature control loop is applied on the heat exchangers in order to maintain the temperature of the process gas outlet. The control loop applied is based on feedback mechanism in which the action of the system is based on the output of the process.

Manipulated Variable: Cooling water Flow rate

Controlled Variable: Temperature of process gas

TIT: Temperature Indicator Transmitter

TRC: Temperature Recorder Controller

TY: Temperature Relay

TCV: Temperature Control Valve

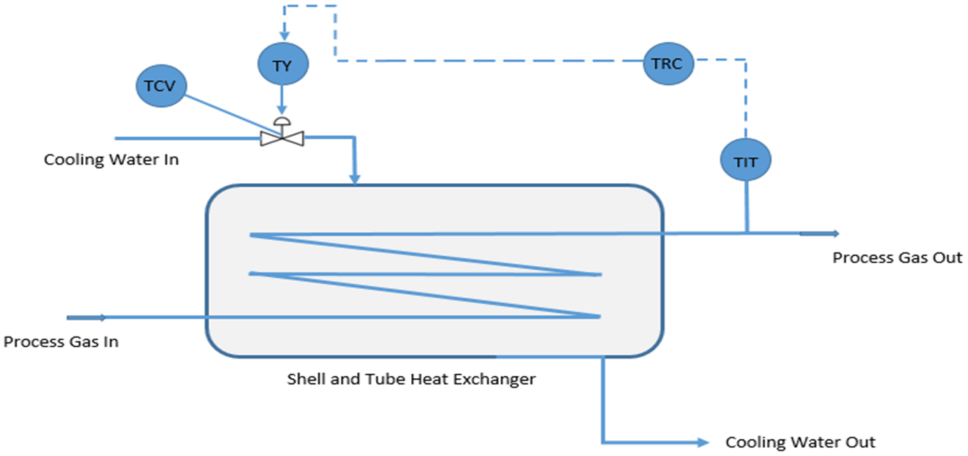


Figure 34 PID of Exchanger

9.3 Control Loops applied on Secondary Reformer:

Following control loops were applied on the secondary reformer in order to maintain the temperature of outlet of the process stream and to maintain the pressure of the reactor.

Manipulated Variable	Controlled Variable
Flue Gases Flow rate	Reactor Pressure
Flow Rate of Fuel	Process Stream Outlet Temperature
Flow Rate of Air	Flow Rate Fuel In

Table 43 Variable table

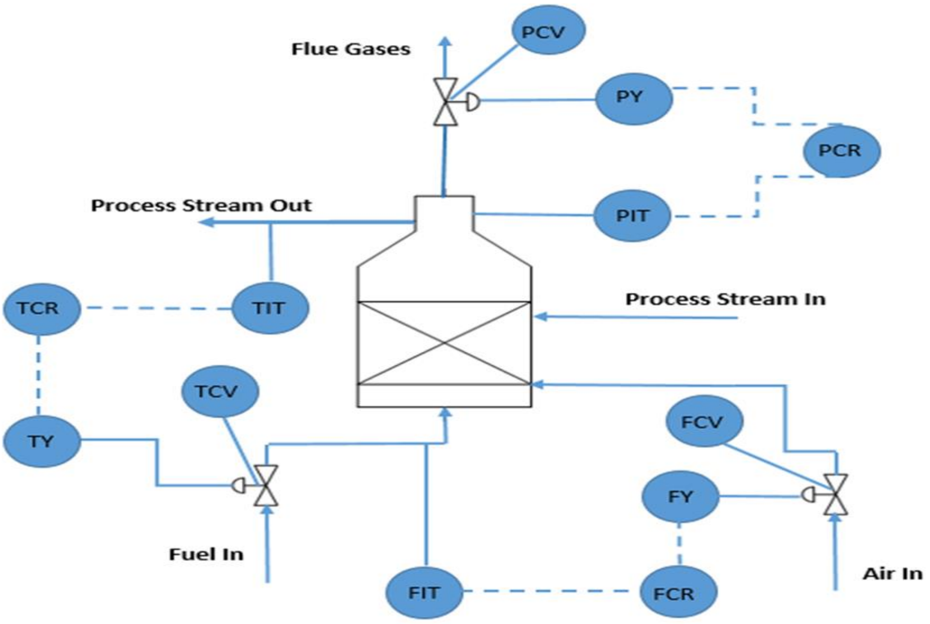


Figure 35 PID of reformer

PIT: Pressure Indicator Transmitter

PCR: Pressure Controller Recorder

PY: Pressure Relay

PCV: Pressure Control Valve

FIT: Flow Indicator Transmitter

FCR: Flow Controller Recorder

FV: Flow Relay

FCV: Flow Control Valve

Transfer functions were calculated according to the give calculations and the PID of the control loops was simulated using Simulink. The transfer calculations are given as follows with the Simulink simulation and its results.

9.4 Temperature Control Loop on Secondary Reformer

9.4.1 Transfer Function Derivation:

Apply energy balance on the Secondary Reformer:

(Rate of flow of energy into tank) − (Rate of flow of energy out of tank) +
(Rate of flow of energy on from heat source) =
(Rate of energyaccumulation in tank)

Connecting this energy balance in symbols results in:

$$wC\left(T_i - T_{ref}\right) - wC\left(T - T_{ref}\right) + q = \rho VC\frac{dT}{dt} \quad - (1)$$

Where T_{ref} is the reference of temperature and C is the capacity of heat of the fluid. At steady state, dT/dt is zero, and (1) can be written as:

$$wC(T_{is} - T_s) + q_s = 0 \quad - (2)$$

Where subscript s has been used to indicate steady state. Subtracting Eq. (2) from Eq. (1) gives:

$$wC(T_i - T_{is}) - wC(T - T_s) + q - q_s = \rho VC\frac{d(T - T_s)}{dt} \quad - (3)$$

If we assume that T_i is constant (and so T_i = T_{is}) and introduce the deviation variables,

$$T' = T - T_s$$

$$Q = q - q_s$$

Eq. (3) becomes:

$$-wCT' + Q = \rho VC\frac{dT'}{dt} \quad - (4)$$

Taking Laplace transforms of Eq. (4) gives:

$$-wCT'(s) + Q(s) = \rho VCsT'(s) \quad - (5)$$

Rearranging Eq. (5) produces the following first-order transfer function relating T’(s) and Q(s):

$$\frac{T'(s)}{Q(s)}=\frac{1/wC}{\left(\frac{\rho V}{w}\right)s+1}=\frac{K}{\tau s+1}\quad -(6)$$

9.4.2 Transfer Function for thermocouple:

In order to derive it following assumptions are made:

1.Only film around bulb provide resistance to heat transfer.

2.Only mercury has thermal capacity and can attain uniform temperature instantly.

3. No expansion or contraction of glass wall containing mercury will take place

By applying unsteady-state energy balance,

$$(Input\ rate)-(Output\ rate)=(Rate\ of\ accumulation)$$

We will get:

$$hA(x-y)-0=mC\frac{dy}{dt}\qquad -(7)$$

In the equation above,

A = surface area of the bulb

C = heat capacity of mercury,

m = Mass in pounds of Mercury.

t = time in unit hour.

h = heat transfer coefficient, Btu/(ft3*h*°F)

This equation states when the heat flows through any film of resistance surrounding a bulb internal energy (mercury) will increase. This increment in value of internal energy can also tell us about temperature changes as well as mercury’s expansion depicts thermometer reading.

Value of h always depends upon the flow rate and fluid characteristics in which it is present and also the dimensions of involving bulb, h is always constant for thermometer

For steady state process:

$$hA(x_s - y_s) = 0 \qquad -(8)$$

Difference of Equation 7 and 8 will give us

$$hA\{(x - x_s) - (y - y_s)\} = mC\frac{d(y - y_s)}{dt} \qquad -(9)$$

X=$(x - x_s)$

Y=$(y - y_s)$

Then Eq 9 will now become

$$hA(X - Y) = mC\frac{dY}{dt}$$

Now, if

$\tau = \frac{mC}{hA}$

Then

$$(X - Y) = \tau\frac{dY}{dt} \qquad -(10)$$

Taking Laplace of Eq 10

$$X(s) - Y(s) = \tau s\, Y(s)$$

By Rearranging,

$$\frac{Y(s)}{X(s)} = \frac{1}{\tau s + 1}$$

9.5 PID of Temperature control loop

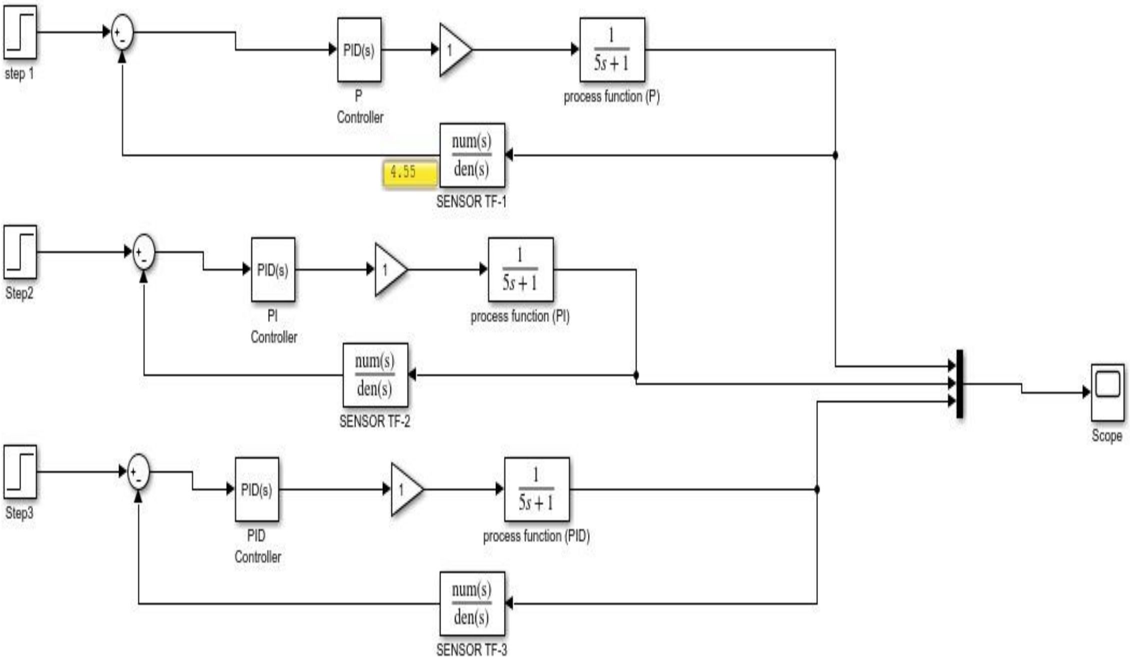


Figure 36 temperature Control PID

Data

P=10

I=3

D=3

For sensor, τ =0.5

For system, τ =0.5

Simulink Plot of Temperature control Loop

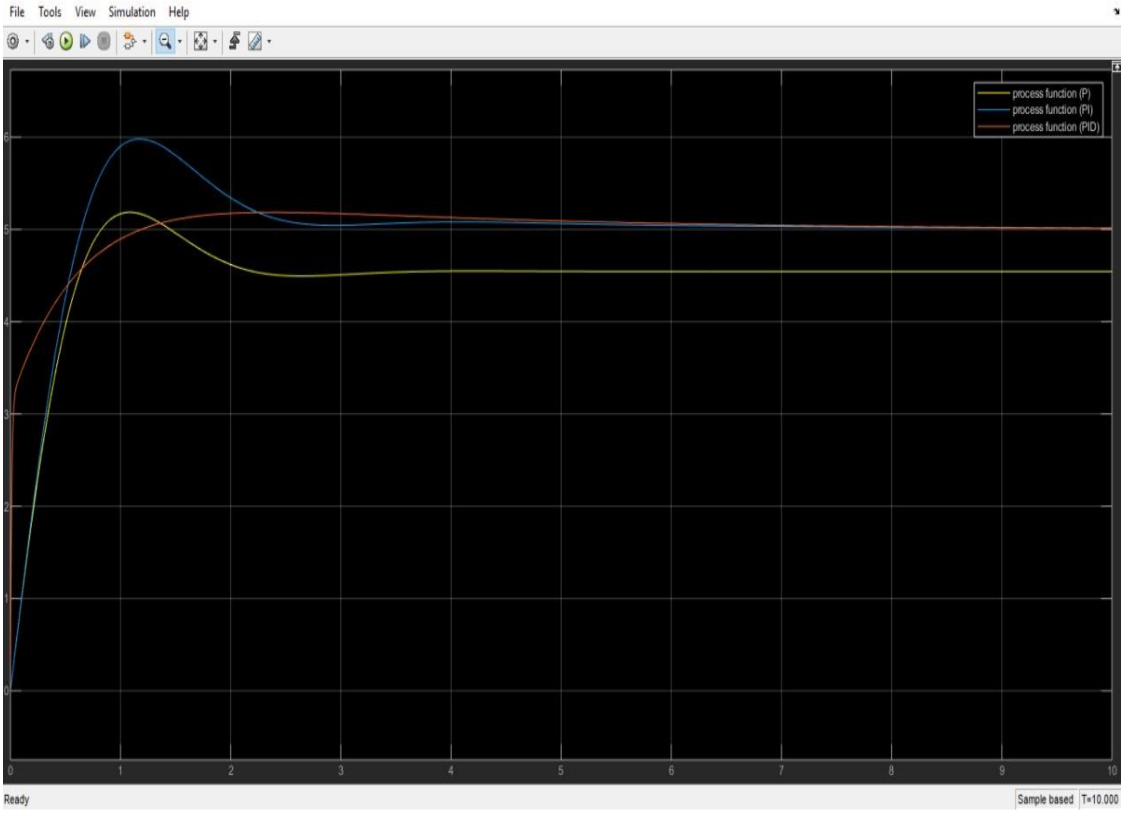


Figure 37 Simulink Plot

CONCLUSION

The feed comparison is as follows:

Mari Gas Feed	
Components	**Moles**
Nitrogen	337
Carbon dioxide	149.73
Methane	1494.5
Ethane	3.75
Propane	0
Butane	0
Total Moles	**1984**

Table 45 Mari feed composition

Mari Gas (70%) and LNG (30%)	
Components	**Moles**
Nitrogen	235.85
Carbon dioxide	104.52
Methane	1604.68
Ethane	29.617
Propane	6.504
Butane	2.471
Total Moles	**1984**

Table 44 Mixed feed composition

From the feed comparison it is clear that number of moles of methane and ethane has increased on the other hand moles of nitrogen has decreased.

Through the moles comparison and cost analysis it is found that using a mixture of RLNG (30%) and Mari gas (70%) as the feed stream in the pre-existing ammonia plant is feasible for profitable urea production. To conclude, the cost of urea produced by using the suggested methodology increases but as suggested by the material balance the amount of moles of ammonia and subsequently urea are also increasing due to increasing H:C ratio as proved by material balance. This increased production of Urea can increase the revenue and the company can still earn profits in the manageable range. Moreover, the RLNG which is to be imported from Qatar is subsidized by the Government of Pakistan to give incentives to the fertilizer companies. The subsidy on RLNG also helps improve the profits for FFC. The thermal demand of the process will also be augmented but within a bearable range as evidenced by energy balance. This thermal energy demand can easily be managed as it is in a considerable rage. The increased moles of feed require the reactors to have more catalyst for good conversion. The usefulness of this study is in the fact that FFC will not require to change its whole operation rather few alterations to the plant can help them produce urea without big investments. The only adverse impact using a mixture of RLNG and Mari gas is the potential of coke formation due to the presence of propane and butane in the feed. Other alternative technologies like coal or biomass gasification will require FFC huge investments for new equipment and to setup a new plant. Our proposed solution will not only help FFC to continue produce urea at a steady pace but allows urea production to increase with minor changes to pre-existing plant equipment.

Product Stream using Mari Gas	
Components	**Moles**
Nitrogen	6.03
Methane	17.18
Water	0
Hydrogen	14.77
Ammonia	2567
Argon	2.99
Total Moles	**2607.97**

Table 47 Product composition with Mari feed

Product Stream Using Mixture	
Components	**Moles**
Nitrogen	1.17
Methane	13.57
Water	0
Hydrogen	41.93
Ammonia	2712
Argon	4.27
Total Moles	**2774.44**

Table 46 Product composition using mixed feed

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