Manufacture of Ethylene: Process and Plant Design

Shubham Bobde & Shibashish Devidutta Jaydev

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ABSTRACT

Ethylene, scientifically known as Ethene, is one of the industrial relevant petrochemicals and a acts a feed stock for a large number of commercial chemicals being produced. The ability to be easily polymerized renders ethylene the quality to be easily converted to polyethylene, which in turn is an economically important product. Ethylene is so widely used in the chemical industry that the estimated usage of Ethylene in the year of 2016 was marked at around 160 million tonnes. Taking into due considerations the importance and the increasing demand of ethylene, this plant design project is an investigation into the nuances of the industrial ethylene production. It will explore the various methods available for manufacturing Ethylene, decide on the most relevant method and delve into the design of a suitable chemical plant. The project will make an effort to understand the bottlenecks encountered during the production and then suggest suitable, plausible design variation to overcome the bottlenecks.

ACKNOWLEDGEMENTS

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Needless to mention, this project could not have been completed without the crucial help of Mr. Venkatesh Sowdhill, our final year project coordinator. He has been greatly kind in giving us ample amount of time to clear several doubts. We would also like to thank the non-teaching staff of the department who have been instrumental in providing us with field data regarding various chemicals.

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CHAPTER 1 INTRODUCTION

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1.1 MOTIVATION FOR SELECTING ETHYLENE

The motive behind the selection of Ethylene lies in the huge economic value of the chemical. Apart the economical value only, the multitude of ways in which it can be employed as a feed stock is also a major reason for the selection. In spite of having a very simple structure, what makes Ethylene an interesting subject of study is the large scale requirement of the same. The easy chemical transformation of Ethylene into ethanol and polyethylene makes it even more demanding and valuable. The global growth rate of demand is around 3.5%, forecast over the next 5 years. The current capacity of world scale plants is around 1 million tons per year up from the 300 thousand tons per year world scale plants of the late 70s to early 80s. The clearly indicated growing demands calls for a thorough study of the production processes.

1.2 STRUCTURE AND PHYSICAL PROPERTIES

Ethylene is the simplest unsaturated organic chemical, with the chemical form $CH_2 = CH_2$. The IUPAC name of the chemical compound is Ethene.

1.2.1 Structure

This hydrocarbon has four hydrogen atoms bound to a pair of carbon atoms that are connected by a double bond. All six atoms that comprise ethylene are coplanar. The molecule is also relatively rigid: rotation about the C-C bond is a high energy process that requires breaking the -bond. Ethylene is a colorless gas with a sweet odor and taste. It is lighter than air. The -bond in the ethylene molecule is responsible for its useful reactivity. The double bond is a region of high electron density, thus it is susceptible to attack by electrophiles.

1.2.2 Properties

Few relevant properties of Ethylene are as follows:

- Molar Mass: 28.05 g/mol

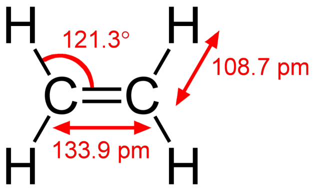


Figure 1.1: 2-D structure of Ethylene showing the bond degrees and the angles.

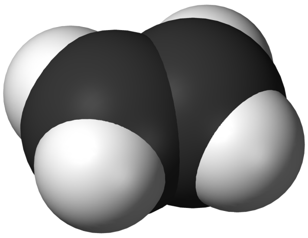


Figure 1.2: 3-D structure of Ethylene.

- Appearance: Colorless
- Density 1.178 kg/m^3 at 15 °
- Solubility in water 2.9 mg/L
- Melting Point 104 K
- Boiling Point 169.5 K
- Solubility in Ethanol 4.22 mg/L
- Critical Temperature 282.65 K
- Critical Pressure 50.5 atm
- Flash Point 137 K
- Autoignition temperature 815.9 K
- Standard molar enthalpy for formation +52.47 kJ/mol
- Standard molar entropy 219.32 $J/K/mol$

1.3 MAJOR USES

- In the United States and Europe, approximately 90% of Ethylene is used to produce Ethylene oxide, Ethylene dichloride, Ethyl benzene and Polyethylene.

- The largest outlet, accounting for 60% of ethylene demand globally, is polyethylene. Low density polyethylene (LDPE) and linear low density polyethylene (LLDPE) mainly go into film applications such as food and non-food packaging, shrink and stretch film, and non-packaging uses. High density polyethylene (HDPE) is used primarily in blow moulding and injection moulding applications such as containers, drums, household goods, caps and pallets. HDPE can also be extruded into pipes for water, gas and irrigation, and film for refuse sacks, carrier bags and industrial lining.

- The next largest consumer of ethylene is ethylene oxide (EO) which is primarily used to make ethylene glycol. Most monoethylene glycol (MEG) is used to make polyester fibres for textile applications, PET resins for bottles and polyester film. MEG is also used in antifreeze applications. Other EO derivatives include ethyoxylates (for use in shampoo, kitchen cleaners, etc), glycol ethers (solvents, fuels, etc) and ethanolamines (surfactants, personal care products, etc).

- Other ethylene derivatives include alpha olefins which are used in LLDPE production, detergent alcohols and plasticizer alcohols; vinyl acetate monomer (VAM) which is used in adhesives, paints, paper coatings and barrier resins; and industrial ethanol which is used as a solvent or in the manufacture of chemical intermediates such as ethyl acetate and ethyl acrylate.

- Ethylene also serves as a hormone in plants. It acts at trace levels throughout the life of the plant by stimulating or regulating the ripening of fruit, the opening of flowers, and the abscission (or shedding) of leaves. Commercial ripening rooms use ”catalytic generators” to make ethylene gas from a liquid supply of ethanol.

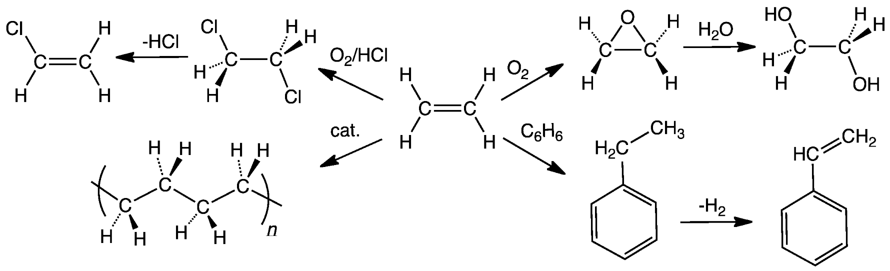


Figure 1.3: Clockwise from the upper right: its conversions to ethylene oxide, precursor to ethylene glycol; to ethyl benzene, precursor to styrene; to various kinds of polyethylene; to ethylene dichloride, precursor to vinyl chloride.

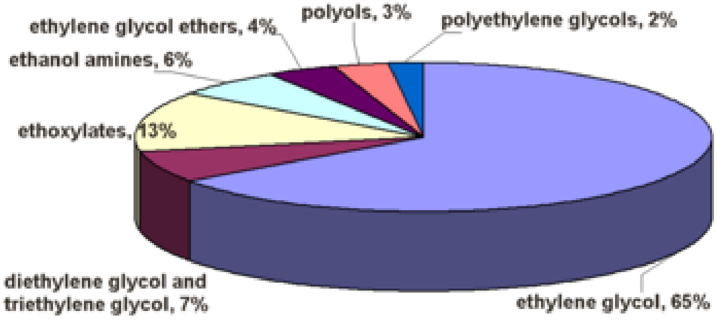


Figure 1.4: Pie chart showing the comparison of the produce of various chemicals from Ethylene.

1.4 PRODUCTION PROCESSES

The various commercial production processes that can be employed to produce Ethylene are as follows:

1.4.1 Methanol to Olefin Process

The Methanol to Hydrocarbons process was discovered at Mobil Oil in 1977. This process is used to convert methanol to products such as olefins and gasoline. The methanol can first be obtained from coal or natural gas. In the Methanol to Olefins (MTO) process, the methanol is then converted to olefins such as ethylene and propylene. The olefins can be reacted to produce polyolefins, which are used to make many plastic materials.

Critical to the successful application of the MTO process are acidic zeolite catalysts, which are discussed in detail on the zeolites page. Without these catalysts, the chemical reactions involved in the MTO process would be too slow for the process to be economically feasible.

The conversion of methanol to olefins on acidic zeolites takes place through a complex network of chemical reactions. The distribution of products and thus the selectivity depends on the temperature, among other factors. Selectivity is a measure of the amount of one product produced relative to others when the possibility to form multiple products exists. Selectivity depends on temperature through the Arrhenius law for the different rate constants.

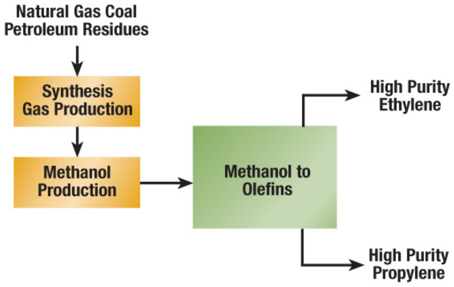


Figure 1.5: Block diagram for the Methanol-to-Olefin process.

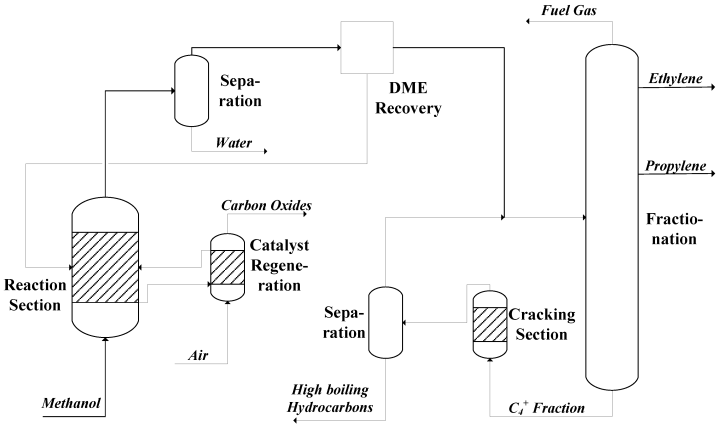


Figure 1.6: Process flow diagram for the Methanol-to-Olefin process.

1.4.2 Ethylene from renewable sources

Ethanol extracted from biological sources can be used to used to produce ethylene by use of dehydration process. This is normally employed in highly developed countries where the hydrocarbon resources are minimal. [1]

$$C_2H_5OH \longleftrightarrow C_2H_4 + H_2O \tag{1.1}$$

Due to the high purity of ethylene product required (99.96%), nearly all of the byproducts must be removed from the final product stream. These two factors form the basis of this plant design: a high temperature reactor and an intricately designed separations train with product purity as the main goal.

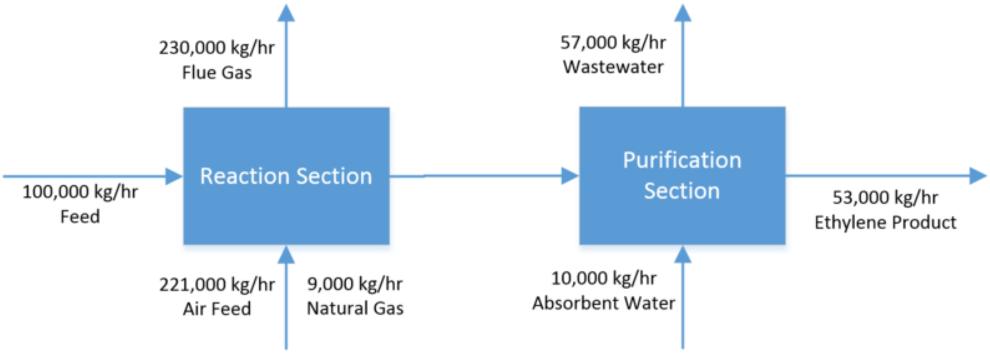


Figure 1.7: Block flow diagram for the production of Ethylene from Ethanol

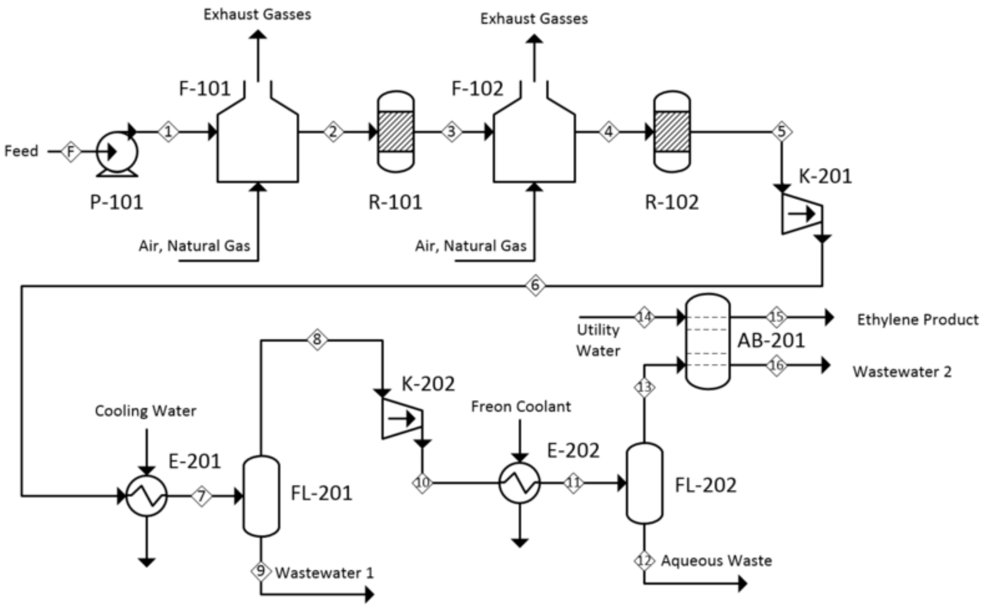


Figure 1.8: Process flow diagram for the production of Ethylene from Ethanol

1.4.3 Steam Cracking

This is the most commonly used, industrially viable process used for the purpose of production of Ethylene from either light or heavy feed stock. Light feed stock normally consists of Natural Gas liquids whereas the heavy feed stock normally refers to Naphtha and Gas oils. Modern ethylene plants incorporate following major process steps : cracking compression and separation of the cracked gas by low temperature fractionation. The nature of the feed stock and the level of pyrolysis severity largely determine the operating conditions in the cracking and quenching section.

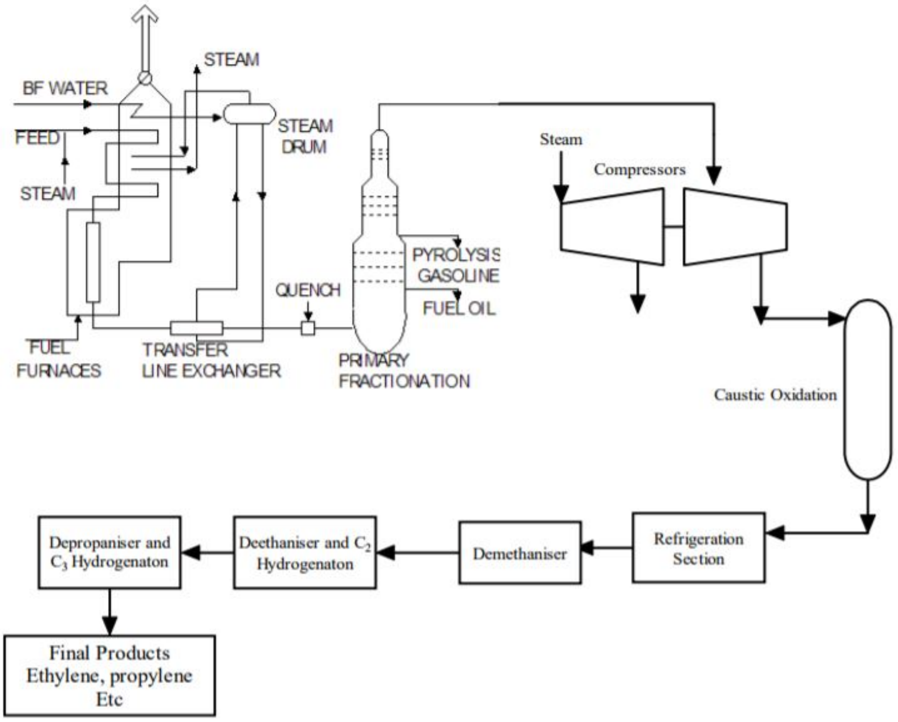


Figure 1.9: Process flow diagram for the production of Ethylene by steam cracking

1.5 PROCESS SELECTION

Of all the processes that have been described above, the production of ethylene from renewable sources is something which is not economically feasible in the context of India. Besides, this process is only suitable for production of considerably low quantities of Ethylene. As the requirements of the petrochemical industry in India is much higher than the produce that can be obtained from renewable source, the process is not an apt one for a developing country such as India.

As for the production of Ethylene from using Methanol, there is continuation formation of the coke on the zeolites catalyst being used. This is a cost intensive process and could be a bottleneck for cost reduction. The process was developed only in 1990s by UOP and is still under its nascent stages as far the technology available in concerned. Also there is globally only one MTO plant in Norway, which is indication of the fact that it is not economically suited to the requirements of a developing nation such as India.

The only reason that looks viable in this scenario is the process of steam cracking. Some reasons for selecting the steam cracking process would be the following:

- The process is simple in principle, without the requirement of any catalyst or indicator making it easy to design and also cost friendly
- Coke formed can be removed easily compared to the MTO process
- Steam is a very widely available utility.
- It is an effective way to convert heavy petrochemicals (undesired) to light and valuable petrochemicals (desired).
- As Naphtha is a readily available feed stock unlike the ethanol form bio-sources, it is easy to start the process.
- It is a well established and well-described process with readily available technical backup to facilitate the smooth design of the plant.

1.6 STEAM CRACKING OF NAPHTHA

1.6.1 Major sections

The main sections that are usually present in the steam cracking process are as follows [2]:

- **Hot Section**: It consists of convection zone and radiant zone. In the convection zone, hydrocarbon feed stock is preheated and mixed with steam and heated to high temperature. In the convection zone the rapid rise in temperature takes place and pyrolysis reaction takes place. The addition of dilution steam enhances ethylene yield and reduces the coking tendency in the furnace coils. The production of the pyrolysis reaction consists of a wide range of saturated and unsaturated hydrocarbons.

- **Quenching Section**: To avoid subsequent reaction the effluent are fixed in their kinetic development by sudden quench first by indirect quench by water to 400-450 °C in transfer line exchanger or quench boiler. This is a large heat exchanger that is a bundle of metal tubes through which the gases pass and around which is circulated water under pressure. The hot water produced is used to generate steam for use in the plant. In the next step the quench is done by heavy product of pyrolysis

- **Cold Section**: After compression, caustic scrubbing and drying the light effluents enter the cold section of the unit which performs the separation of hydrogen to various concentration (ii) ethylene containing 99.4% (iii) 95% propylene (iv) A C4 cut containing 25-50% butadiene (v) pyrolysis gasoline which is rich in aromatic hydrocarbons. The complexity of the separation section of a cracker increases markedly as the feed changes from ethane.

1.6.2 Reactions involved in steam cracking

The reactions involved in thermal cracking of hydrocarbons are quite complex and involve many radical steps. The thermal cracking reaction proceeds via a free radical mechanism. Two types of reactions are involved in the thermal cracking (i) primary cracking where the initial formation of paraffin and olefin takes place (ii) secondary cracking reaction where light products rich in olefins are formed. The total cracking reactions can be grouped as follows:

- Initiation Reaction
- Propagation Reaction
- Addition Reaction
- Isomerization Reaction
- Termination Reaction
- Molecular crystallisation reaction.

1.6.3 Generalized composition of the feed stock

Naphtha are mixture of alkane, cycloalkanes, and aromatic hydrocarbons depending on the type of oil from which the naphtha was derived. The group properties of these components greatly influence the yield pattern of the pyrolysis products. A full range naphtha boiling range approximately 20 to 200 °C would contain compound, with from 4-12 carbon atms. Short naphtha boiling point range from 100-140 °C and long chain naphtha boiling point lies around 200-220 °C. The steam cracking of the naphtha yields wide variety of products, ranging from hydrogen to highly aromatic heavy liquid fractions. The thermal stability of hydrocarbons increases in the following order: parafins, naphthenes, aromatics. Yield of ethylene as well as that of propylene is higher if the naphtha feed stock is rich in paraffins.

1.6.4 Constraints

- Material properties: tube material properties limit the temperature that can be maintained in the furnace and thereby the feedstock conversion.
- Controllability: It is difficult to control the process as it has large number of unit operations and processes taking place.
- Scale: As the requirement for the ethylene is very high and steam cracking is the cheapest process, it is a constrain to keep up with the huge demand. The turbines and compressors being used need to developed of large sizes to meet the requirements. [3]

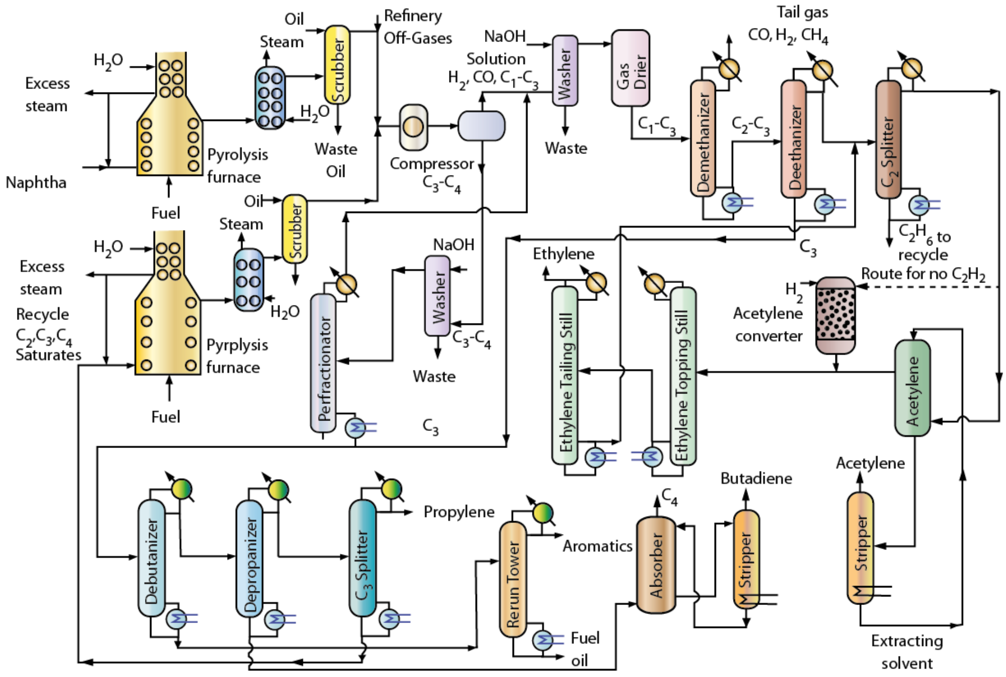


Figure 1.10: Process flow diagram for the production of Ethylene from steam cracking of Naphtha with balanced flow streams

CHAPTER 2 MASS BALANCE

To evaluate the mass balance across each process equipment, first we need to decide on the final process flow diagram of the whole process. The final process flow diagram of the process flow diagram is as shown below. According to the data available from the MOL group, Hungary, we are able to determine the composition of the products being obtained from the steam cracking on naphtha in a fired tubular furnace. The typical composition of the product stream emerging from the furnace, assuming no mass of naphtha is accumulated in the furnace is shown in the table below [4].

Table 2.1: Composition of outlet stream of furnace

Hydrocarbon	Percetage(%)
H_2S, CO_2	1.03
CH_4	15.35
C_2H_2	0.69
C_2H_4	31.02
C_2H_6	3.42
C_3H_6, C_3H_4	16.21
C_3H_8	0.38
$C4$	9.54
$C5-C12$	22.36

Certain assumptions that have been used for mass balance which has been carried out hereafter. The assumptions have been obtained through suitable reference from the industrial data. [5] The assumptions are as follows:

- The percentage recovery from distillation unit has been taken to be 99% and the percentage purity of the top stream has been assumed to be 99.5%.
- The percentage conversion of the acetylene in the hydrogenator has been assumed to 100% as only front-end hydrogenation has been considered for the naphtha.
- The overall recovery of ethylene has been taken as 97%. This is the amount of ethylene that can be recovered with respect to the ethylene coming out of the furnace.

2.1 DAILY PRODUCTION OUTPUT OF ETHYLENE

The daily production output of ethylene has been fixed to be **500** metric tons/day. As we assume 97% recovery of the amount present with respect to the ethylene that is coming of the furnace, we have to calculate the amount of the naphtha that will be required for the purpose.

- Amount of ethylene finally needed: **500 metric tons/day**
- Amount of ethylene coming out of the furnace: **500/0.97=515 metric tons/day**
- Amount of naphtha required to produce this amount of ethylene **515/0.3102=1660 metric tons/day**

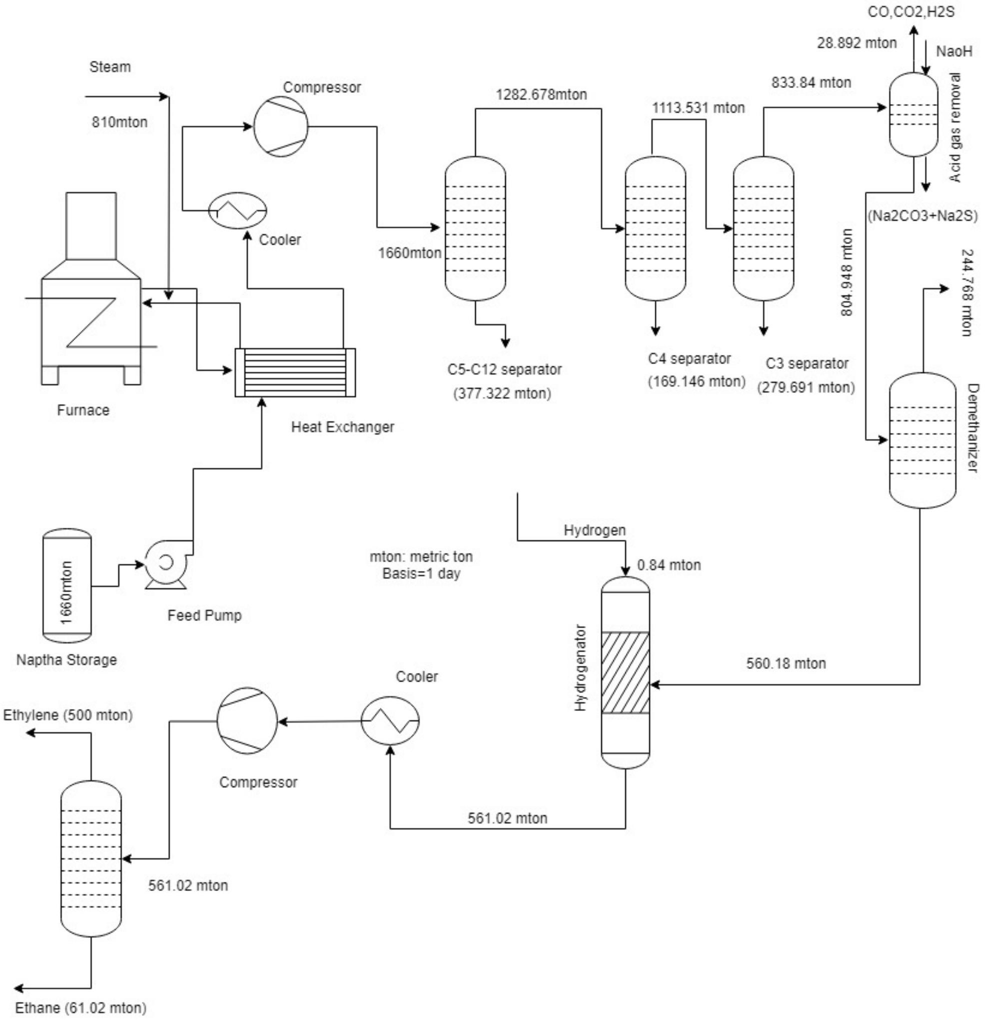


Figure 2.1: Simple process flow diagram for the production of Ethylene from steam cracking of Naphtha with the flow rates of every stream

2.2 MASS BALANCE ACROSS THE FURNACE

Naphtha going in is 1660 metric tons/day.

Product outflow has been shown in the table below, assuming no naphtha is lost in the process of coking.

Table 2.2: Amount of each component in outlet stream of furnace

Hydrocarbon	Amount(metric-tons/day)
H_2S, CO_2	17.098
CH_4	254.81
C_2H_2	11.454
C_2H_4	514.932
C_2H_6	56.772
C_3H_6, C_3H_4	269.086
C_3H_8	6.308
$C4$	158.364
$C5 - C12$	371.176
$Total$	1660

The operating conditions which gives us the output product that we get in the table 3.1 has been highlighted below.

- Temperature= 800-850 degree celsius
- Steam Required= 0.5 kg/kg of naphtha
- Residence time 0.1-0.5 seconds

The addition of the steam for the purpose of dilution ensures that we are able to achieve a near 100 % conversion of naphtha since there is no coking taking place in the furnace.

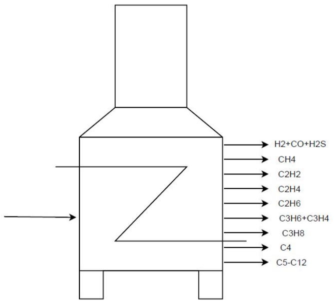


Figure 2.2: Streams coming out of the furnace

2.3 MASS BALANCE ACROSS THE DEC5-DEC12 UNIT

The mass balance over the distillation column has been carried out using the basic mass conservation equation used in the distillation column, which are as follows.

$$F = D + R \tag{2.2}$$

$$Fx_f = Dy_D + R.x_R \tag{2.3}$$

- Feed (F)= 1660 metric tons/day
- x_f= (C1-C4 present in the feed)/total feed= (1.03+15.35+0.69+31.02+3.42+16.21+0.38)/100*16 0.681
- y_D= the purity of the top stream has been assumed to be 99.5 %
- Dy_D= the recovery has been assumed to 99 %, hence the value of Dy_D is valued to be 99/100*0.681*1660=1276.264
- D (distillates stream containing C1-C4)= 1282.678
- R = 377.322
- x_R= 0.0341

Above shown are the model calculation which have been used for all the remaining distillation units namely the debutaniser, depropaniser and demethaniser.The distillate stream of one column acts as the feed for the next column.

2.4 MASS BALANCE ACROSS DEBUTANISER

Table 2.3: Various parameters and their values across the debutaniser unit

Parameter	Value
F	1282.678
x_f	0.8725
Dx_D	1107.9638
x_D	99.5 %
D	1113.531
R	169.146
x_R	0.066

2.5 MASS BALANCE ACROSS DEPROPANISER

Table 2.4: Various parameters and their values across the depropaniser unit

Parameter	Value
F	1113.531
x_f	0.7526
Dx_D	829.67
x_D	99.5 %
D	833.84
R	279.6921
x_R	0.03

2.6 MASS BALANCE ACROSS THE CAUSTIC WASH UNIT

The chemical reactions taking place in the caustic wash unit are as follows.

$$\mathrm{H_2S(aq) + NaOH(aq) \longrightarrow NaHS(aq) + H_2O} \tag{2.4}$$

$$\mathrm{NaHS(aq) + NaOH(aq) \longrightarrow Na_2S(aq) + H_2O} \tag{2.5}$$

$$\mathrm{CO_2(aq) + NaOH(aq) \longrightarrow NaHCO_3(aq) + H_2O} \tag{2.6}$$

$$\mathrm{NaHCO_3(aq) + NaOH(aq) \longrightarrow Na_2CO_3(aq) + H_2O} \tag{2.7}$$

- Amount of $H_2S + CO_2$ which is left over at the end of three distillation can be evaluated as to be $1.03 * 1660 * (0.99)^3 = 16.6$ metric tons/day.
- Lets say that the conversion would requires X kgs of NaOH. All the X kg of NaOH will be flushed out in the outlet stream as the efficiency of the scrubber is assumed to be 100%.

Table 2.5: Inlet parameters in the caustic wash unit

Parameter	Value
F	833.84 metric tons/day
$x_f(C1 - C2)$	0.975
$H_2S + CO_2$	16.6
$NaOH$	X metric tons/day
Net inlet $(F + NaOH)$	833.84 + X

Table 2.6: Outlet parameters in the caustic wash unit

Parameter	Value
$Na_2S + 4H_2O + Na_2CO_3$ + unrecovered hydrocarbons	X+12.26 metric tons/day
Non-distillate Outlet	28.892 metric tons/day
Distillate outlet (0.975*0.99*833.84/0.995)	804.948 metric tons/day

To calculate the net outlet we have assumed that the stream purity is 99.5 % and the recovery of the required C1-C2 fractions is also 99 %.

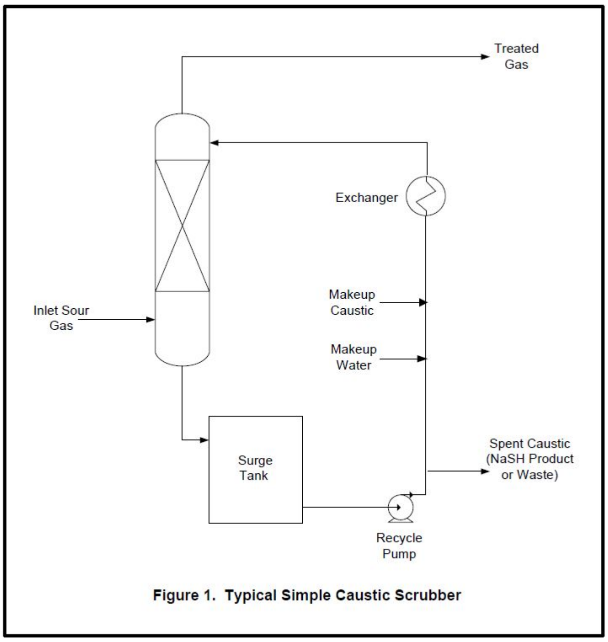


Figure 1. Typical Simple Caustic Scrubber

Figure 2.3: A typical caustic scrubber unit

2.7 MASS BALANCE ACROSS DEMETHANISER

The demethaniser is the unit which uses pressurised inputs within the distillation columns to keep the C1-C2 units in the form of liquid to facilitate the separation.

Table 2.7: Various parameters and their values across the demethaniser unit

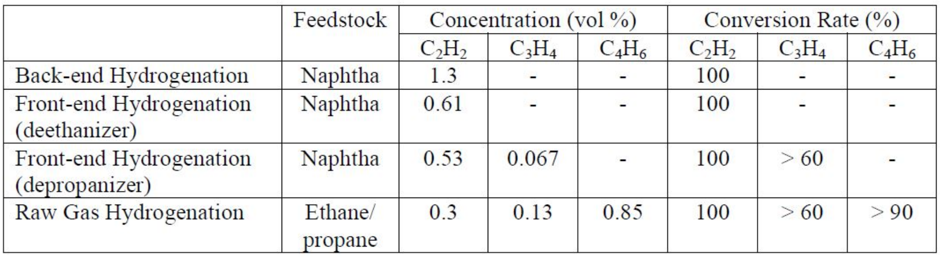
Parameter	Value (mton/day)
F	804.948
x_f	0.31
R	560.18
D	244.768

2.8 MASS BALANCE ACROSS THE HYDROGENATOR

The reaction which is involved in the hydrogenator is as follows.

$$C_2H_2 + H_2 \longrightarrow C_2H_4 \tag{2.8}$$

With the reference from the table presented below, it can be seen that when a front-end hydrogenator is operated on the C_2H_2 coming out of naphtha, the efficiency of the hydrogenator is assumed to be 100 %.



	Feedstock	Concentration (vol %)			Conversion Rate (%)		
		C_2H_2	C_3H_4	C_4H_6	C_2H_2	C_3H_4	C_4H_6
Back-end Hydrogenation	Naphtha	1.3	-	-	100	-	-
Front-end Hydrogenation (deethanizer)	Naphtha	0.61	-	-	100	-	-
Front-end Hydrogenation (depropanizer)	Naphtha	0.53	0.067	-	100	> 60	-
Raw Gas Hydrogenation	Ethane/ propane	0.3	0.13	0.85	100	> 60	> 90

Figure 2.4: Table showing the conversion for all types of hydrogenation [5]

- Amount of C_2H_2 that is to be converted is $0.69 * 0.99^5 * 1660 = 10.89$ metric ton/day.
- Number of moles of acetylene to be removed is 10.89/26 = 0.42 metric ton moles.
- Amount of hydrogen required is same as the amount of acetylene to be converted which is 0.42 metric ton moles/day which translates to 0.84 metric tons/day.
- Amount of ethylene that is produced due to the reaction is 0.42*28= 11.76 metric tons/day.

Here the outlet exceeds the inlet only by the weight of hydrogen added as this converts the acetylene 100 % into ethylene.

2.9 MASS BALANCE ACROSS DEETHANISER

The distillate recovered from the ethaniser unit is the final product which is removed at the rate of 500 metric tons/day and has purity of 99.5 % i.e 0.5 % of it contains some remnant ethane which can’t be separated.

Table 2.8: Amount of each component in outlet stream of furnace

Parameter	Value (metric tons/day)
Inlet	560.18
Amount of acetylene removed	10.89
Amount of Hydrogen supplied	0.84
Amount of ethylene produced	11.73
Outlet	561.02

Table 2.9: Various parameters over the de-ethaniser unit

Parameter	Value
F	561.89
x_f	0.894
Dx_D	497.5
x_D	99.5 %
D	500

2.10 OVERALL MASS BALANCE

As the net is being obtained as 0.00, it is seen that net mass balance over the whole unit has been successfully obtained.

Table 2.10: Mass balance across all streams

Parameter	Amount (metric tons/day)
Feed into furnace	(+) 1660
Amount of Hydrogen added	(+) 0.84
Residue of DeC5-DeC12	(-) 377.322
Residue of Debutaniser	(-) 169.146
Residue of Depropaniser	(-) 279.692
Outlet of the caustic remover	28.892
Residue of Demethaniser	(-) 244.768
2Residue of Deethaniser	(-) 61.02
Ethylene obtained finally	(-) 500
Net	0.00

CHAPTER 3 ENERGY BALANCE

The energy balance, in general, has to take care of the equilibrium between the net energy input and the net energy output of the system. In ideal cases, where there are no losses to the atmosphere, the energy output should be exactly same as the energy input. However, this is not possible in the real scenario. The major assumptions are made with respect to the heat capacity of the various components. The following are the major assumptions which have been made during the energy balance:my

- The heat capacity of the hydrocarbon streams is assumed to be constant across all temperatures. This is attributed to the very low variation in the heat capacity with change in temperature.
- The distillation columns being used as cryogenic distillation columns which implies that the heat is to be removed before the streams are allowed into the distillation column.
- The reference temperature to calculate the enthalpy has been taken to be 273 K
- Heat transfer over the caustic wash units has been treated as negligible.
- In each distillate stream of every distillation column, the enthalpy due to the
- Changes in the heat capacity of the component due to their presence in a mixture has been neglected.

3.1 METHODOLOGY

The inlet and the outlet enthalpies of each stream has been evaluated. The net enthalpy change happening over a unit has can be evaluated by the following equations:

$$H_i = mC_p(T_i - T_*) \tag{3.9}$$

$$H_o = mC_p(T_o - T_*) \tag{3.10}$$

$$\Delta H = H_o - H_i \tag{3.11}$$

The difference in the enthalpy is supplied by the usage of any utility. **The nature of the utility and the specifications are discussed in the design part of the project**

3.2 ENERGY BALANCE OVER THE FURNACE

Key points taken into consideration in this energy balance are:

- Each component of the hydrocarbon stream is formed at a temperature of 700 degree Celsius, which is nearly the temperature at which the reaction is carried out.
- The Naphtha stream is assumed to enter into the 180 degree Celsius, after passing through the transfer line heat exchanger.
- The fuel used in natural gas which is found to have a calorific value of 12500 kcal/kg.

Table 3.1: Components and their enthalpies in the furnace

Hydrocarbon	Amount(metric-tons/day)	CP (kcal/kg/K)	$\Delta H X 10^3$
H_2S, CO_2	17.098	0.242857143	2906.66
CH_4	254.81	0.528571429	94402.54
C_2H_2	11.454	0.402380952	3226.21
C_2H_4	514.932	0.364285714	131307.66
C_2H_6	56.772	0.416666667	16558.5
C_3H_6, C_3H_4	269.086	0.357142857	67271.5
C_3H_8	6.308	0.397619048	1755.726667
$C4$	158.364	0.397619048	44077.98
$C5-C12$	371.176	0.49	127199.492
Naphtha	1660	0.495	-147906
Fuel	105218.4		-1315230
Net			0

As the net enthalpy over the furnace is seen to be zero, we can say that the energy quantities have been balanced over the furnace.

3.3 ENERGY BALANCE OVER THE TRANSFER LINE HEAT EXCHANGER

Key points in energy balance over this unit are:

- The inlet temperature of Naphtha is assumed to be 25 degree Celsius
- The outlet temperature of Naphtha is taken to be 180 degree Celsius
- The Naphtha is heated using the process to process heat transfer
- The inlet enthalpy of Naphtha at 25 degree celsius is 1660*1000*(25-0)*0.495 = 20542500 kcal
- The outlet enthalpy of Naphtha at 180 degree is 1660*1000*(180-0)*0.495= 147946000 kcal
- The net enthalpy difference of Naphtha passing through the heat exchanger= 127363500 kcal
- We are considering that the outlet stream of the furnace is at 700 degree Celsius.
- The temperature of the outgoing hydrocarbon stream 700-127363500/1064508.384= 581 degree Celsius.

Table 3.2: Components and their parameter while passing through the transfer line heat exchanger

Hydrocarbon	Amount(metric-tons/day)	CP (kcal/kg/K)	mC_pX10^3
H_2S, CO_2	17.098	0.242857143	4.152371429
CH_4	254.81	0.528571429	134.8607714
C_2H_2	11.454	0.402380952	4.608871429
C_2H_4	514.932	0.364285714	187.5823714
C_2H_6	56.772	0.416666667	23.655
C_3H_6, C_3H_4	269.086	0.357142857	96.10214286
C_3H_8	6.308	0.397619048	2.508180952
$C4$	158.364	0.397619048	62.96854286
$C5-C12$	371.176	0.49	181.71356
Steam	810	0.452	366.42
Total	2470		1064.580384

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3.4 ENERGY BALANCE OVER THE QUENCHING UNIT

- The inlet temperature of the stream is 581 degree celsius
- The outlet temperature of the stream is 250 degree Celsius
- Referring to the table 3.12 1064.58*1000*(581-250)= 352376107 kcal is the total amount of heat that is to be removed by the utility
- The specifications of the utility across the cooling unit will be discussed in the design part.

3.5 ENERGY BALANCE OVER THE COOLING UNIT

The key consideration in balancing the heat over the cooling unit are as follows:

- Heat of vaporisation of the steam is taken to be 539 kcal/kg
- Heat of vaporisation of 83 kcal/kg
- The inlet temperature of the stream is taken to be 581 degree Celsius and the outlet temperature is taken to be 30 degree 35 degree Celsius.
- Subcooling is considered only for the process steam.
- For the cracked hydrocarbon, only the stream consisting of C5-C12 is assumed to be condensed; however sub-cooling is not considered for the same.

3.6 ENERGY BALANCE OVER THE DE-BUTANISER UNIT

Keys points to be considered for the heat balance across the de-butaniser unit are as follows:

- The inlet temperature of the distillation unit is taken to 30 degree Celsius.
- The operating temperature of the de-debutaniser is -10 degree Celsius in order to keep the butane in the liquid state.
- The lower temperature of the de-butaniser is maintained with the help of a chilling train which a refrigerant in the form of a utility.

Table 3.3: Components and their parameter while passing through the transfer line heat exchanger

Stream	Heat absorbed X 10^3
H_2S, CO_2	892.7598571
CH_4	28995.06586
C_2H_2	990.9073571
C_2H_4	40330.20986
C_2H_6	5085.825
C_3H_6, C_3H_4	20661.96071
C_3H_8	539.2589048
$C4$	13538.23671
$C5-C12$	69848.4674
Steam	544204.2857
Total	725086.9774
Heat removed by utility	-725086.9774
Net heat	0

- Heat balance over the de-butaniser unit and other distillation units are taken as per the following equation:

$$H(feed)+H(utility)=H(distillate)+H(residue) \tag{3.12}$$

- In the energy balance shown, it can be noted that the enthalpy of the distillate due to presence of small quantity has been neglected as this is a negligible quantity.
- Similarly, the contribution of the small quantity of the top product in the enthalpy of the residue is also taken as negligible.

Table 3.4: Components in the feed and their parameter while passing through the debutaniser unit

Hydrocarbon	Amount(metric-tons/day)	CP (kcal/kg/K)	$H(feed)\ X10^3$
H_2S, CO_2	16.92702	0.242857143	123.3254314
CH_4	252.59058	0.528571429	4005.364911
C_2H_2	11.33946	0.402380952	136.8834814
C_2H_4	509.78268	0.364285714	5571.196431
C_2H_6	56.20428	0.416666667	702.5535
C_3H_6, C_3H_4	266.39514	0.3571	2854.233643
C_3H_8	6.24492	0.397	74.49297429
$C4$	156.78036	0.397619048	1870.165723
Total Enthalpy			15338.2161

Table 3.6: Components in the streams and their parameter while passing through the de-butaniser unit

Hydrocarbon	$H\ X10^3$
Feed	15338.2161
Distillate	-4444.456623
Residue	-672.5567143
Heat removed by refrigerant utility	-20455.22943

Table 3.5: Components in the distillate stream and their parameter while passing through the de-butaniser unit

Hydrocarbon	Amount(metric-tons/day)	$H(distillate)\ X10^3$
H_2S, CO_2	16.7577498	-40.69739237
CH_4	250.0646742	-1321.770421
C_2H_2	11.2260654	-45.17154887
C_2H_4	504.6848532	-1838.494822
C_2H_6	55.6422372	-231.842655
C_3H_6, C_3H_4	263.7311886	-941.8971021
C_3H_8	6.1824708	-24.58268151
Total Enthalpy		-4444.456623

3.7 ENERGY BALANCE ACROSS THE DE-PROPANISER UNIT

Key points to be considered for the energy balance across the de-propaniser unit are as follows:

- The entering stream which is coming from the debutaniser is at a temperature of -10 degree Celsius.
- The exiting stream which is coming out of de-propaniser is at -44 degree celsius, which also happens to be operating temperature of the depropaniser unit.

Table 3.7: Components in the feed stream and their parameter while passing through the de-propaniser unit

Hydrocarbon	Amount(metric-tons/day)	$H(feed)\ X10^3$
H_2S, CO_2	16.7577498	-40.69739237
CH_4	250.0646742	-1321.770421
C_2H_2	11.2260654	-45.17154887
C_2H_4	504.6848532	-18338.494822
C_2H_6	55.6422372	-231.842655
C_3H_6, C_3H_4	263.7311886	-941.8971021
C_3H_8	6.1824708	-24.58268151
Total Enthalpy		-4444.456623

Table 3.8: Components in the distillate stream and their enthalpies while passing through the de-propaniser unit

Hydrocarbon	$H(distillate)\ X10^3$
H_2S, CO_2	-177.2778412
CH_4	-5757.631953
C_2H_2	-196.7672669
C_2H_4	-8008.483446
C_2H_6	-1009.906605

Table 3.9: Components in various streams and their enthalpies in the de-propaniser unit

Hydrocarbon	H $X10^3$
Feed	-4444.456623
Distillate	-15150.1
Residue	-4430.2896
Heat removed by refrigerant utility	-15135.90009

3.8 ENERGY BALANCE OVER THE DE-METHANISER UNIT

The key points taken into consideration while balancing the energy over the de-methaniser unit are as follows:

- As the methane is a liquid below -145 degree Celsius and C_2 products are liquid below -89 degree Celsius, we set the operating temperature of the de-methaniser unit somewhere in between.
- The operating temperature of the de-methaniser column, operating with the help of a chilling train and a refrigerant is set to -100 degree celsius.

Table 3.10: Components in the feed stream and their parameter while passing through the de-methanisr unit

Hydrocarbon	Amount(metric-tons/day)	$H(feed)\ X10^3$
CH_4	245.0883872	-5700.055633
C_2H_2	11.0026667	-194.7995942
C_2H_4	494.6416246	-7928.398612
C_2H_6	54.53495668	-999.8075391
Total Enthalpy		-14823.06138

Table 3.11: Components in the distillate stream and their enthalpies while passing through the de-methaniser unit

Hydrocarbon	$H(distillate)\ X10^3$
CH_4	-12825.12518

Table 3.12: Components in various streams and their enthalpies in the de-methaniser unit

Hydrocarbon	$H\ X 10^3$
Feed	-14823.06138
Distillate	-12825.12518
Residue	-20166.48
Heat removed by refrigerant utility	-18168.5438

CHAPTER 4   DESIGN OF EQUIPMENTS

4.1 FURNACE DESIGN

The following are some of the design consideration for designing a furnace for the purpose of steam cracking

- Heaters shall be designed for uniform heat distribution
- Multi-pass heaters shall be designed for hydraulic and thermal symmetry of all passes. The number of passes shall be minimized. Each pass shall be a single circuit
- Average heat flux density in the radiant section is normally based on single row of tubes with two nominal tube diameter spacing.
- The maximum allowable inside film temperature for any process service shall not be exceeded in the radiant, shield, or convection sections.
- minimum radiation loss 2.5% the total heat input
- Natural draft needs 25% excess air when oil is the primary fuel and 20% excess air when fuel gas is the primary fuel. In case of forced draft operation, 20% Excess air for fuel oil and 15% Excess air for fuel gas
- Heaters shall be designed such that a negative pressure of at least 0.10 inches of water (0.025 kilopascals) is maintained in the radiant and convection sections at maximum heat release with design excess air.
- The flue gas dew point can be predicted, and the minimum tube-metal temperature can be kept high enough to prevent condensation, if the fuel's sulfur content has been correctly stated. (For estimated flue gas dew points with respect to sulfur content in fuel oil and gas
- In a well-design heater, the radiant-section heat duty should represent more than 60% to 70% of the total heat duty
- The bridge wall temperature should range between 800C to 1,000C.

- Higher radiant flux means less heat transfer surface area for a given heat duty; hence, a smaller furnace.

- The higher the film temperature, the greater is the tendency of the fluid (particularly a hydrocarbon) to crack and deposit a layer of coke.

- Heat-transfer fluids tend to degrade quickly at high film temperatures.

- The coke layer acts as an insulator, retarding heat transfer, which could cause tube overheating and lead to tube failure.

- Also, a heavy coke deposit can restrict the flow through the coil, lowering the inside heat transfer coefficient and further increasing the tube wall temperature.

- The smallest firebox for a certain duty will obviously produce the cheapest design.

- The flame impingement and consequent tube failure that could result can be avoided by specifying a minimum safe distance between burners and tubes, based on experience

- Provision for thermal expansion shall take into consideration all specified operating conditions, including short term conditions such as steam-air decoking.

- The convection section tube layout shall include space for future installation of sootblowers. or steam lancing doors.

- The convection section shall incorporate space for future addition of two rows of tubes.

- When the heater is designed for fuel oil firing, soot-blowers shall be provided for convection section cleaning.

- Vertical cylindrical heaters shall be designed with maximum height to diameter ratio of 2.75, where the height is the radiant section height and the tube circle diameter.

- Shield sections shall have at least three rows of bare tubes.

- Convection sections shall be designed to minimize flue gas bypass. Baffles may be employed.

- The minimum clearance from grade to burner plenum or register shall be 6 feet 6 inches (2.0 meters) for floor fired heaters.

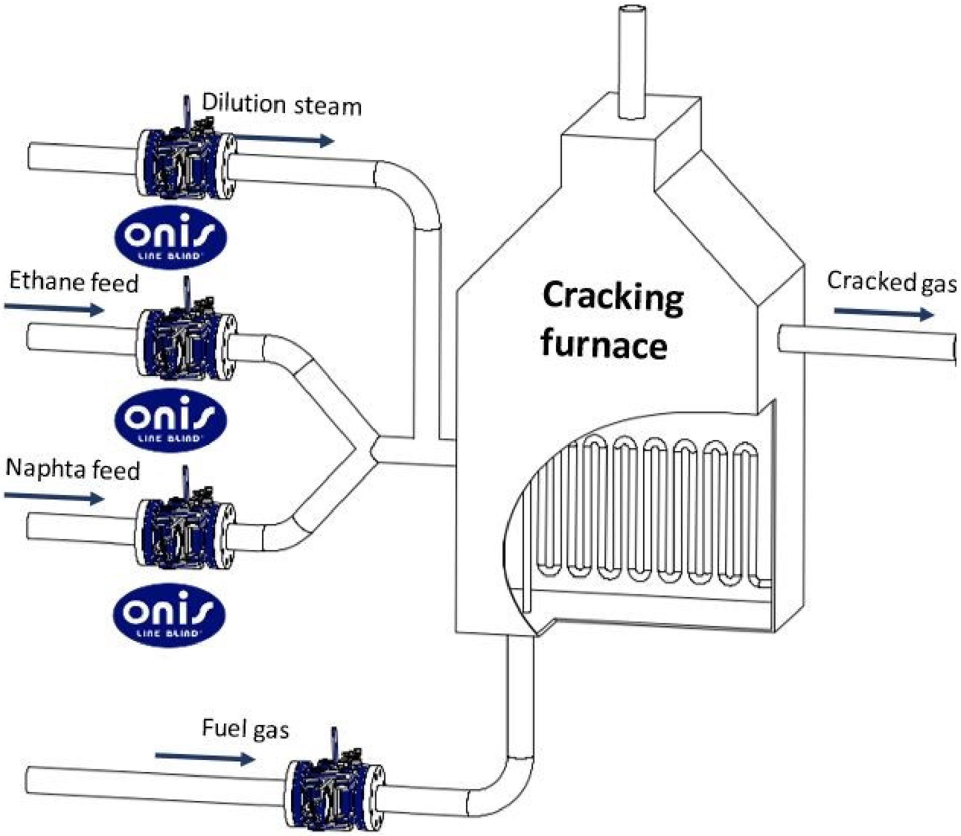

Figure 4.1: Rough 3-D sketch of a Furnace Design

4.1.1 Design Steps

- The length of the furnace tubes is taken to be 20 m and the diameter(OD) is taken to be 219 mm.
- The centre to centre distance of the tubes placed inside the furnace is taken as 394 mm.

- The total heat that is liberated by the fuel is, Q_f= 1315230000 kcal.

- The radiant section area of a single can be calculated from the dimension of a single tube. Area =$2\pi rL$=2*3.14*219/2/1000*20= 13.75 m^2.

- The radiant section heat flux is 920661000/1.09 = 844643199 $kcal/m_2/day$, as the basis for all the calculations has been taken to be a day.

- The temperature of the gases above the bridge-wall has been approximately taken to be 900 degree Celsius.

- The temperature of the gas above the bridge-wall is around 150 degree more than the bridge-wall temperature, and can be taken around 1900 degree Fahrenheit

- The temperature of the tube surface to ensure maximum heat flux is taken to 200 degree Fahrenheit.

- By use of the tube surface temperature (Ts) and the temperature of the gas above the bridge-wall, we can evaluate the heat flux through the radiant section, q , by using the following equations.

$$\frac{\Sigma Q}{\alpha_{cp} A_{cp} f} = 2 * q \tag{4.13}$$

- Where, $\alpha_{cp} A_{cp}$ is the net cold surface area and f is the overall exchange factor.

- The value of $\frac{\Sigma Q}{\alpha_{cp} A_{cp} f}$ arrived at by use of the chart show in figure 4.2, is 70,000 $btu/hr/ft^2$.

- Taking this amount on a daily basis and in the units of $kcal/day/m^2$, we get 70000*24*2.71= 4552800

- Thus the value of q is calculated to be 4552800/2=2276400 $kcal/m^2/day$

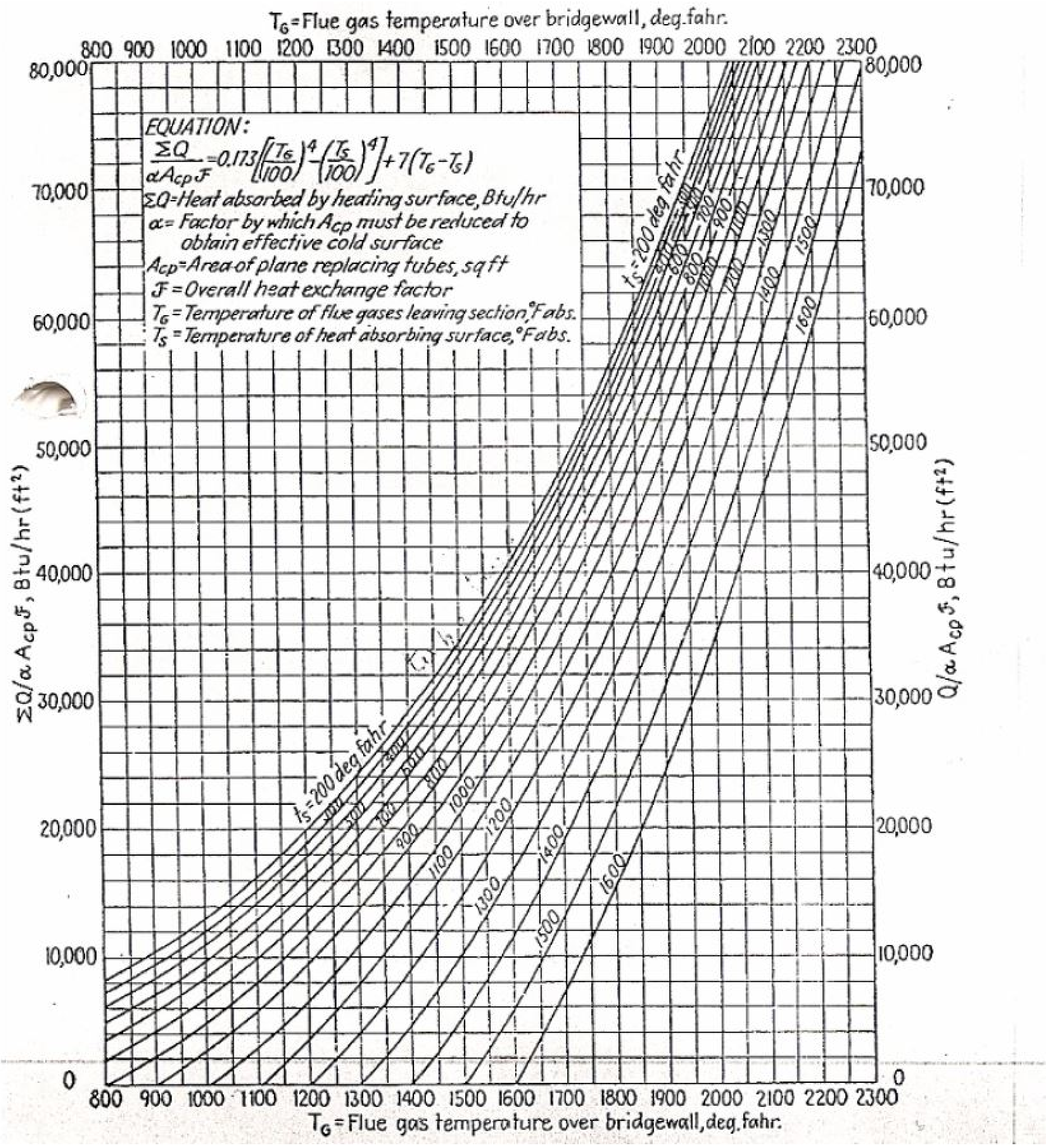

Figure 4.2: The graph showing the relationship between Tg, Ts and $\frac{\Sigma Q}{\alpha_{cp}A_{cp}f}$

- Th fuel being used is natural gas which has a calorific value of 12500 kcal/kg
- The amount fuel required is, m_{fuel} 1315230000/12500= 106 tons.
- For natural gas, the air to fuel ratio maintained is taken to be, G'= 10:1.
- Amount of air needed is 1060 tons.

- Assuming 25% of excess air is used, we need 1.25*1060 tons= 1325 tons.

- We also require the usage of atomizing steam to be able to ensure easy mixing of the fuel with the air and burn quickly. However in this case, we will neglect the usage of atomising steam for the sake of convenience.

- Amount of heat associated with the inlet air, Q_{air}= 1325*1000*0.25*(105-25)=26304600 kcal.

- The amount of heat absorbed by the furnace wall is, Q_w= 2%*Q_f = 0.02*1315230000= 26304600 kcal.

- The total heat absorbed by the tubes is, $\Sigma Q = Q_f + Q_{air} - Q_w$ = 1315230000+26304600-26304600= 1315230000 kcal/hr

$$N_{tubes} = \frac{\Sigma Q}{2\pi rLq} = \frac{1315230000}{2*3.14*20*0.1095*2267400*} = 42 \tag{4.14}$$

$$A_{cp} = ctc*L*N = 42*0.394*42 = 695m^2 \tag{4.15}$$

- Where N=no. of tubes, L=length of tubes, ctc=centre to centre distance of the tubes.

- Placing the tubes in a horizontal fashion within the furnace, we get height of the furnace required to just place the tubes would be, ctc*N= 0.394*42= 16.548 m.

- Taking the clearance between the tubes and the furnace wall to be 0.22 meters, the total width of the furnace can be taken as L+2*clearance= 20.44m

Table 4.1: Various parameters and their values relating to the design of the furnace

Parameters	Values
length of a furnace tube(m)	20
diameter of the furnace tube (mm)	219
centre-to-centre distance of the tubes (mm)	394
heat exchange area of single tube (m^2)	13.75
tube surface temperature ($^\circ F$)	200
gas temperature over bridgewall ($^\circ F$)	1900
air to fuel ratio	10:1
Calorific value of the fuel (kcal/kg)	12500
number of tubes	42
cold plane area (m^2)	695
Clearance b/w tubes and furnace wall(m)	0.22
Total width of the furnace(m)	20.44

4.2 DESIGN OF DISTILLATION COLUMN

The considerations for design of the distillation column are as follows:

- A de-methaniser is being designed.
- The operating pressure for the de-methaniser is 15 atm and the operating temperature is -100 °Celsius.

4.2.1 Determination of number of trays

- The calculation of the relative volatility is done using Raoult's law.

$$P_t = X_f.P_A + (1 - X_f).P_B \tag{4.16}$$

- Here the P_t is the total pressure which is equal to 15*760=11400 mm Hg
- X_f, as evaluated in the mass balance section has been taken as 0.31
- As the more volatile component that is to be removed is Methane, the vapour pressure of methane, P_A if found out using the following equation and graph.

$$log_{10}P_{mmHg} = 6.61184 - \frac{389.93}{266 + T_{°C}} \tag{4.17}$$

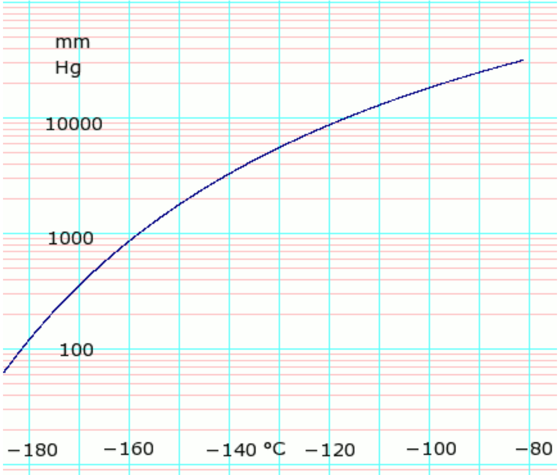

Figure 4.3: Vapor pressure data for methane

- The vapor pressure of methane at the temperature of -100 °C is found to me 18317.4 mm Hg.

- Using equation 16, P_B can be evaluated to be 8292.17 mm Hg.

$$\alpha = \frac{P_A}{P_B} = \frac{18317.4}{8289.17} = 2.3 \tag{4.18}$$

- With the reflux ratio fixed at 4, the following curve is being obtained by usage of the McCabe-Thiele Method.

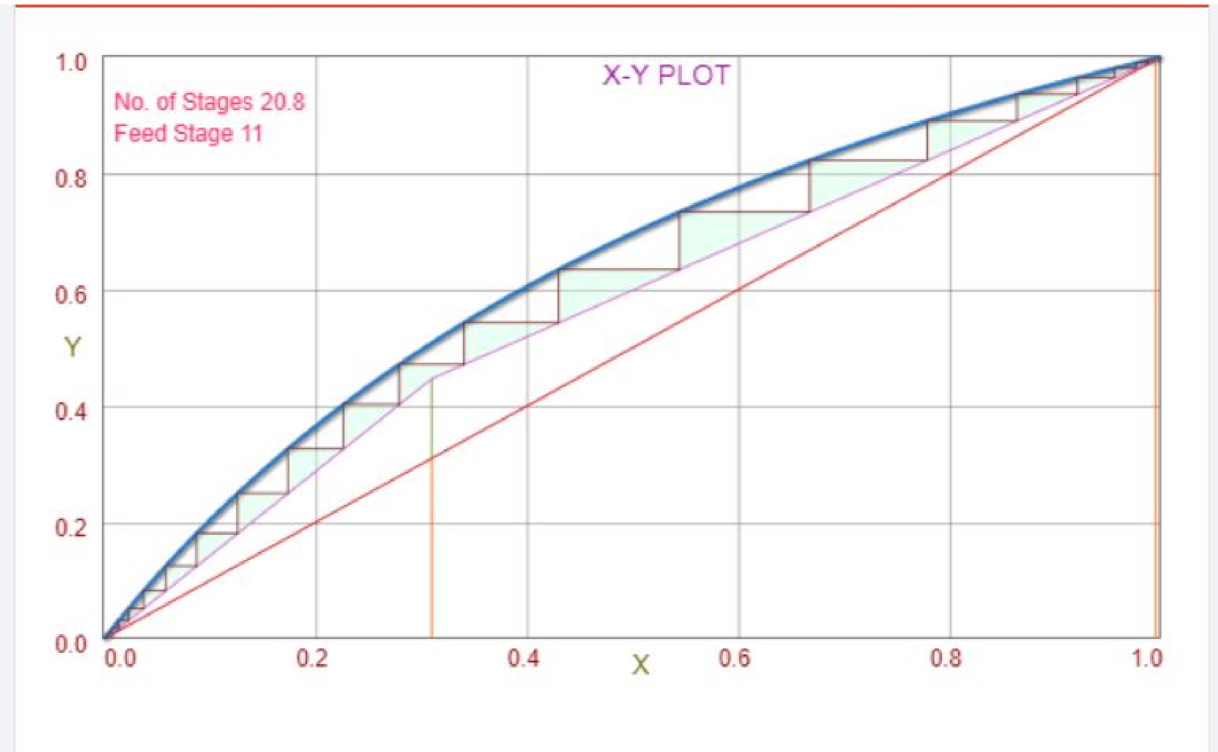

Figure 4.4: McCabe-Thiele solution for the de-methaniser unit.

- The number of stages is found to be 20.8 and the feed tray is found to be the 11th tray.

4.2.2 Determination of column diameter

The various pre-requisite parameters which have to be taken into consideration for the evaluation of the column diameter are presented in table 4.2

Table 4.2: Various parameters and their values required for the calculation of column diameter

Parameters	Values
Number of stages	20
stage efficiency	80%
Real number of stages	25
Mol. wt of the feed	22.875
Slope of the bottom line (McCabe-thiele curve)	1.5
Slope of the top line (McCabe-thiele curve)	0.65
Feed flow rate (mton/day)	804.948
Feed flow rate(kmol/day)	$35.19X10^3$
Top product rate (D)(mton/day)	244.768
Top product rate (kmol/day)	$15.298X10^3$
Gas flow rate at the top, V= D(R+1) (kmol/day)	$76.49X10^3$
Residue flow rate at the bottom (F-D)	$19.892X10^3$
Tray spacing (mm)	600

The important formulae used in the process of evaluating the column diameter are:

$$F_{LV} = \frac{L}{V}(\rho_V/\rho_L)^{0.5} \tag{4.19}$$

$$u_f = K.\frac{\rho_L - \rho_V}{\rho_V} \tag{4.20}$$

Assuming the actual velocity to be 85% of the actual velocity to keep a check on flooding

$$u_v = 0.85u_f \tag{4.21}$$

$$Max-volumetric-flow-rate, V_m = \frac{mass-flow-rate}{\rho} \tag{4.22}$$

$$A = \frac{V_m}{u_v} \tag{4.23}$$

$$diameter, d = (\frac{4A}{\pi})^{0.5} \tag{4.24}$$

The evaluated parameters of the top section of the distillation column have been shown in the table 4.3 below:

Table 4.3: Parameters of the top half of the column

Parameters	Values
$\rho_V(kg/m^3)$	0.656
$\rho_L(kg/m^3)$	422.6
F_{LV}	0.026
$u_f(m/s)$	3.04
$u_v(m/s)$	2.6
$Vm(m^3/s)$	21.59
$A(m^2)$	8.3
$d(m)$	3.25

The evaluated parameters of the bottom section of the distillation column have been shown in the table 4.4 below:

Table 4.4: Parameters of the bottom half of the column

Parameters	Values
$\rho_V(kg/m^3)$	1.18
$\rho_L(kg/m^3)$	570
F_{LV}	0.0682
$u_f(m/s)$	2.63
$u_v(m/s)$	2.24
$Vm(m^3/s)$	12.96
$A(m^2)$	4.9
$d(m)$	2.48

- The chosen diameter of distillation column in 3.25 m. To standardize the size, we take 12 ft size= 3.66 m.
- The cross sectional area area of the distillation column is 10.52 m^2.
- The downcomer area is 12% of the total area, 12% of 10.52 = 1.26 m^2

4.3 DESIGN OF THE HEAT EXCHANGER

The transfer line heat exchanger is being designed. The assumptions used in the heat exchanger are as follows:

- Steady state heat flow
- Convective Heat Transfer Coefficient is Constant throughout process
- The tube side fluid is taken to be naphtha.
- Data unavailable is assumed from reliable sources.

The design details of the transfer line heat exchanger are as follows:

- Fluid on shell side: naptha
- Fluid on tube side: Product recovery
- Heat capacity on tube side : Cp =1.8828 $kcal/kg/°C$
- For Shell side: Inlet temperature = 25 °C, Outlet temperature = 180 °C
- For Tube Side: Inlet temperature = 700 °C Outlet temperature = 581 °C
- Shell side flow rate = 1660 metric ton/day
- Tube Side flow rate = 2470 metric ton/day
- Heat Load= 1064548.384*(700-581)=126681257.7 Kcal/day = 6134.656112 KW
- U=550 W/m^2 °C
- ΔT_{LMTD}=36/.0669=537.8
- Provisional Area= 6134656.112/550*537.8=21.567 m_2
- Outside Diameter of tube = 25mm
- Inside Diameter of tube = 19mm
- Tube Length= 6m

Tube side coefficient and ΔP_t

Tube side Coefficient :

- Mean Temperature = 640.5 °C
- Tube Cross sectional area =π/4*192=283.53 mm^2
- Area of one Tube = πd_oL= 0.471 m^2
- No. of tubes = $\frac{Provisional area}{Area of one tube}$= 21.567/.471 = 44.034=46.
- Tubes per pass = 45 (No. of Passes is taken as 1)
- Total flow area per pass = 46*283.53*10^{-6} m_2 =12.98* $10^{-3} m^2$
- Density=700 kg/m^3
- Mass Velocity=3.134 m/s
- μ =0.73 CP
- Re=57098.8, k=.01
- Pr=uCp/k=.00073*1880/.15=577.26
- L/D=6000/19=315.79
- J_h=.0025(from graph)
- J_f= 0.0042 (from graph)

$$\frac{h_i * d}{k} = J_h * Re * pr^{0.33} (u/u_w)^{0.14} \tag{4.25}$$

- Using the above equation and putting the necessary values, we get, h_i=980 W/m^2 °C
- Pressure Drop:

$$\Delta P = N_p [8 J_f * (L/d) + 2.5] * \rho * \frac{u_t^2}{2} = 34648.28 N/m^2 = 5 psi \tag{4.26}$$

- The tube side pressure drop is within the allowable pressure drop on tube side of 10 psi.

Shell side coefficient and ΔP

- We used pull through floating head and therefore, the clearance for the calculated bundle diameter of 254.6 mm, is found to be 88 mm.
- D_s = Clearance *2+ = 88*2+ 254.6= 430.6 mm
- Baffle spacing is taken t 25% of the shell diameter l_b= $D_s/4$ =86.12 mm
- Pitch = outer diameter* 1.25 = 31.25
- Assuming Square Pitch for effective cleaning of the tubes, the effective diameter can be computed by the formula below:

$$d_e = \frac{1.27}{d_o}(P_t^2 - 0.785d_o^2) \tag{4.27}$$

- The effective diameter is evaluated to be 24.68 mm.
- The cross flow area is computed by the following equation:

$$CFA = \frac{P_t - d_o}{P_t} * D_s * l_b = 0.00741m^2 \tag{4.28}$$

- Velocity = 3.38 m/s
- Reynold's Number is:

$$Re = \frac{\rho * v * d_e}{\mu} = 88590.753 \tag{4.29}$$

$$Pr = \frac{\mu C_p}{k} = 10.17 \tag{4.30}$$

- J_h(from graph)= 2.2*10^{-3}
- We are neglecting the correction in heat transfer coefficient due to variation in viscosity
- The heat transfer coefficient on the shell side can be find using the equation below:
- We take the baffle cut to be 25%

$$\frac{h_s * d_e}{k_f} = J_h.Re.Pr^{1/3}(\mu/\mu_w)^{0.14} \tag{4.31}$$

- The h_s is computed to be 2514.07 $W/m^2/°C$
- The tube side fouling factor on the tube side is 8000 $W/m^2/°C$ and on the shell side is 9000 $W/m^2/°C$.
- Here the value of thermal conductivity of naphtha is, k_f, is taken to be 0.15 $W/m/°C$
- Here the value of thermal conductivity of naphtha is, k_w, is taken to be 43 $W/m/°C$
- The overall heat transfer coefficient of the shell and tube side is found out by the relation is as below:

$$\frac{1}{U_o} = \frac{1}{h_s} + \frac{1}{h_t} + \frac{d_o.ln(d_o/d_i)}{2k_w} + \frac{d_o}{d_i}\frac{1}{9000} + \frac{d_o}{d_i}\frac{1}{8000} \tag{4.32}$$

- The value of U_o is calculated to be 480.076 W/m^2.
- The pressure on the shell side can be found out to be:

$$\Delta P_s = 8J_f \frac{D_s}{d_e}\frac{L}{l_b}\frac{\rho.u_s^2}{2}(\frac{\mu}{\mu_w})^{-0.14} \tag{4.33}$$

- The shell side pressure drop on substitution of the values is found to be 20899.368 N/m^2

4.4 DESIGN OF REACTOR

The consideration for design of reactor design are as follows:

- The residence time of the reaction is taken to be 5 minutes, τ = 300 seconds.
- ρ=1.027 kg/m^3
- The volumetric flow rate, v_o is

$$\frac{V}{v_o} = \tau \tag{4.34}$$

- From the above equation, the volume of the reactor is calculated to be, 39.65 m^3.
- The ratio of $\frac{H}{D}$=4

$$\frac{\pi * d^2 * H}{4} = \pi * d^3 = 39.65 \tag{4.35}$$

- Solving the above equations, d= 2.328 m and H= 9.313 m.

- Assuming clearance above and below the bed to be 1/8th of H, the total height = (2*1/8+1)H= 5/4*H=11.64 m.

CHAPTER 5 PLANT AND SITE LAYOUT

The economic construction and efficient operation of a process unit will depend on how well the plant and equipment specified on the process flow-sheet is laid out. A detailed account of plant layout techniques cannot be given in this short section. The principal factors to be considered are:

1. Economic considerations: construction and operating costs
2. The process requirements
3. Convenience of operation
4. Convenience of maintenance
5. Safety
6. Future expansion
7. Modular construction

The main areas which are to be taken into consideration when coming up the plant layout are as follows:

- Plant area, which includes
    1. Area for furnace, which needs to be in one column because of its highly hazardous nature.
    2. Area for raw materials storage- naphtha and fuel. Storage area to be kept at sufficient distance from the furnace to avoid explosion in case of any mishap in the furnace
    3. Area for distillation columns to be kept at the center as this is the main unit in the
    4. Reactor area
    5. Area to setup the caustic wash unit
    6. A dedicated strip area to place the inter-stage compressors so as to keep them adjacent to the distillation columns.

7. Area for expansion in future
8. Area for utility generation, such as generation of steam and supply of other heat transfer utility

- Administrative area which constitute of the following area:
    1. Administrative office
    2. Training center
    3. Parking Area
- Area for fire safety
- Storage tank area.
- Waste treatment plant
- Control and maintenance area which has the following units
    1. Warehouse
    2. Maintenance
    3. Workshop
    4. Laboratory
    5. Control building

A plant layout taking into account the aforementioned units/areas has been designed in the attached appendix.

CHAPTER 6 SAFETY ANALYSIS

6.1 EXPOSURE POTENTIAL

Ethylene exists naturally in the environment, where it is produced by vegetation and other natural sources. It is also a combustion product from vehicle exhaust, forest fires, and cigarette smoke.

6.1.1 Workplace Exposure

Exposure can occur either in facilities that recover or produce ethylene, during transport, or in facilities that use ethylene. Workplace exposure e to elevated levels of ethylene is limited due to concerns for flammability. Ethylene is produced, distributed, stored, and consumed in closed systems. Those working with ethylene in manufacturing operations could be expose d during maintenance, sampling, testing, or other procedures. The American Conference of Governmental Industrial Hygienists (ACGIH) has adopted a Threshold Limit Value (TLV) of 200 ppm for ethylene as an 8-hour time-weighted average (TWA). Workers should consult with the appropriate regulatory agencies for exposure guidelines. Each manufacturing facility should have a thorough training program for employees and appropriate work processes, ventilation, and safety equipment in place to limit exposure. See Health Information.

6.1.2 Consumer exposure to products containing ethylene

No consumer uses of ethylene are known, so consumer exposure to commercially produced ethylene is unlikely. However, it is used to make plastics and other materials used in consumer products. For ex ample, plastic milk jugs and plastic bags a re made from high-density polyethylene, a polymer made from ethylene. Ethylene is naturally present in the environment, and the highest environmental concentrations are found in urban areas.

6.1.3 Environmental releases

Ethylene emissions from industrial facilities are subject to governmental requirements. As a result of these regulations and production facility operating conditions, the typical ambient air levels of ethylene will be significantly less than 1 ppm. Ethylene has a high vapor pressure and if released will tend to volatilize from water and soil and accumulate in the atmosphere. In the presence of oxygen, ethylene biodegrades with an estimated half-life of 1.9 days. It degrades more quickly in sunlight. This material is considered slightly toxic to aquatic organisms on an acute basis.

6.1.4 Large release

Industrial spills or releases are infrequent and generally contained. If a large spill does occur, the area should be isolated until the released gas has dispersed. Respiratory protection may be necessary for clean up. The primary hazard from a large release of ethylene is fire. Eliminate all sources of ignition immediately. Use only explosion-proof equipment; ground and bond all containers and handling equipment. Prevent entry into soil, ditches, sewers, waterways and/or groundwater.

6.1.5 In case of fire

Deny any unnecessary entry into the area. Do not attempt to extinguish the fire. Stop flow of material if possible and allow the fire to burn out. Extinguish small fires with water spray or fog, carbon-dioxide or dry-chemical extinguishers, or foam. If possible, fight the fire from a protected area or safe location. Firefighters should wear positive-pressure, self-contained breathing apparatus (SCBA) and protective firefighting clothing. The public should be warned of any downwind vapor explosion hazards. Vapors may travel a long distance and ignition or vapor flash-back may occur. Immediately withdraw all personnel from the area in case of rising sounds from venting safety device or discolorations of the container. Follow emergency procedures carefully.

6.1.6 Health information

Eye contact

Contact with liquefied ethylene can cause frostbite.

Skin contact

Contact with liquefied ethylene can cause frostbite. No adverse effects are expected from absorption through the skin.

Inhalation

Because ethylene is a gas at normal temperatures and pressures, inhalation is the primary route of exposure. Ethylene has a low level of toxicity. No risk to human health has been identified from occupational exposure or exposure of the general public to atmospheric levels of ethylene. However, excessive exposure by inhalation may cause headache, dizziness, anesthesia, drowsiness, unconsciousness, or other central nervous system effects. In confined or poorly ventilated areas, the gas can accumulate and result in unconsciousness due to displacement of oxygen. The American Conference of Governmental Industrial Hygienists (ACGIH) has established a Threshold Limit Value (TLV) for occupational exposure to ethylene of 200 parts per million (ppm) as a time-weighted average (TWA). The odor threshold for ethylene is reported to be from 270 to 600 ppm, so odor is not an adequate warning property to prevent excessive exposure to ethylene. Consult governmental regulations for exposure guidelines for the geographic region of interest.

Ingestion

Ingestion is unlikely because ethylene is a gas at normal temperatures. Ingestion of liquefied gas can cause frostbite of the lips, mouth, and throat.

6.1.7 Other health related information

Screening and long-term studies in laboratory animals suggest that exposure to ethylene does not affect fetal development or reproduction and does not cause cancer. Both in vivo and in vitro mutagenicity studies were negative for genetic toxicity. However, metabolic studies in animals and humans have revealed that ethylene is metabolized to ethylene oxide, which is known to have carcinogenic and mutagenic effects.

6.2 ENVIRONMENTAL INFORMATION

If released, ethylene tends to accumulate in the atmopsphere because of its high vapor pressure. In the presence of oxygen in the atmosphere, ethylene biodegrades with an estimated half-life of 1.9 days. It degrades even more quickly in sunlight. Ethylene is slightly toxic to aquatic organisms in the most sensitive species tested. The minute amounts of ethylene measured in water represent little, if any, environmental hazard to aquatic animals. Ethylene has a low bioconcentration potential. Because ethylene has low toxicity and low potential for exposure, it is not likely to have adverse effects on terrestrial wildlife. Ethylene exists naturally in the environment and is produced by plants as part of their life cycle and the ripening process. Ethylene also acts as a plant hormone in regulation of plant growth and development. Ethylene effects on plants appear to be complex, depend upon environmental conditions and stage of growth, and vary by species and cultivar.

6.3 PHYSICAL HAZARD INFORMATION

Ethylene is an extremely flammable gas or pressurized liquid and should be used only in well-ventilated areas. It should be kept away from heat and sources of ignition.

Ethylene is stable at recommended storage conditions of less than 52C (126F). Decomposition with rapid pressure build-up can occur at storage conditions above 180C (356F) and 1136 kPa (150 psig). Avoid contact with oxidizing materials (such as chlorine and oxygen), mineral acids, metals, and metal chlorides. Hazardous polymerization can also occur at high temperatures or in the presence of free-radical initiators or activated materials, such as activated carbon or molecular sieves.

6.4 MATERIAL SAFETY DATA SHEET

A sample material safety data sheet published by the Linde Group has been attached herewith for the reference of the readers on the material safety aspects of Ethylene.

THE LINDE GROUP

Linde

Safety data sheet
Ethylene

Creation date : 28.01.2005 Version : 6.0 DE / E SDS No. : 055a
Revision date : 04.01.2011 page 1 / 3

1 IDENTIFICATION OF THE SUBSTANCE/PREPARATION AND OF THE COMPANY

Product name
Ethylene
EC No (from EINECS): 200-815-3
CAS No: 74-85-1
Index-Nr. 601-010-00-3
Chemical formula C2H4
REACH Registration number:
Not available.
Known uses
Not known.
Company identification
Linde AG, Linde Gas Division, Seitnerstraße 70, D-82049 Pullach
E-Mail Address Info@de.linde-gas.com
Emergency phone numbers (24h): 089-7446-0

2 HAZARDS IDENTIFICATION

Classification of the substance or mixture

Classification acc. to Regulation (EC) No 1272/2008/EC (CLP/GHS)
Press. Gas (Liquefied gas) - Contains gas under pressure; may explode if heated.
Flam. Gas 1 - Extremely flammable gas.
STOT SE 3 - May cause drowsiness or dizziness.

Classification acc. to Directive 67/548/EEC & 1999/45/EC
F+; R12 | R67
Extremely flammable.
Vapours may cause drowsiness and dizziness.
Risk advice to man and the environment
Compressed gas.
Label Elements

- Labelling Pictograms

- Signal word
Danger

- Hazard Statements

H280	Contains gas under pressure; may explode if heated.
H220	Extremely flammable gas.
H336	May cause drowsiness or dizziness.

- Precautionary Statements

Precautionary Statement Prevention

P210	Keep away from heat/sparks/open flames/hot surfaces. - No smoking.
P260	Do not breathe gas, vapours.

Precautionary Statement Reaction

P304+P340+P315	IF INHALED: Remove victim to fresh air and keep at rest in a position comfortable for breathing. Get immediate medical advise/attention.
P377	Leaking gas fire: Do not extinguish, unless leakcan be stopped safely.
P381	Eliminate all ignition sources if safe to do so.

Precautionary Statement Storage

P403	Store in a well-ventilated place.

Precautionary Statement Disposal
None.

3 COMPOSITION/INFORMATION ON INGREDIENTS

Substance/Preparation: Substance.
Components/Impurities
Ethylene
CAS No: 74-85-1
Index-Nr.: 601-010-00-3
EC No (from EINECS): 200-815-3
REACH Registration number:
Not available.
Contains no other components or impurities which will influence the classification of the product.

4 FIRST AID MEASURES

Inhalation
In high concentrations may cause asphyxiation. Symptoms may include loss of mobility/consciousness. Victim may not be aware of asphyxiation. In low concentrations may cause narcotic effects. Symptoms may include dizziness, headache, nausea and loss of co-ordination. Remove victim to uncontaminated area wearing self contained breathing apparatus. Keep victim warm and rested. Call a doctor. Apply artificial respiration if breathing stopped.
Ingestion
Ingestion is not considered a potential route of exposure.

5 FIRE FIGHTING MEASURES

Specific hazards
Exposure to fire may cause containers to rupture/explode.
Hazardous combustion products
Incomplete combustion may form carbon monoxide.
Suitable extinguishing media
All known extinguishants can be used.
Specific methods
If possible, stop flow of product. Move container away or cool with water from a protected position. Do not extinguish a leaking gas flame unless absolutely necessary. Spontaneous/explosive re-ignition may occur. Extinguish any other fire.
Special protective equipment for fire fighters
In confined space use self-contained breathing apparatus.

6 ACCIDENTAL RELEASE MEASURES

Personal precautions
Wear self-contained breathing apparatus when entering area unless atmosphere is proved to be safe. Evacuate area. Ensure adequate air ventilation. Eliminate ignition sources.
Environmental precautions
Try to stop release. Prevent from entering sewers, basements and workpits, or any place where its accumulation can be dangerous.
Clean up methods
Ventilate area.

055a / EDV / 29.12.2010

Figure 6.1: Material Safety Data Sheet- Page 1

THE LINDE GROUP

Linde

Safety data sheet
Ethylene

Creation date :	28.01.2005	Version : 6.0	DE / E	SDS No. : 055a
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Label Elements

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Incomplete combustion may form carbon monoxide.
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All known extinguishants can be used.
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Clean up methods
Ventilate area.

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Figure 6.2: Material Safety Data Sheet- Page 2

CHAPTER 7 PROJECT ECONOMICS

7.1 GENERAL CONSIDERATIONS

- Basis for calculation of cost: 1 year
- Working capital- 15% of Capital Investment
- General expenses= 3% of manufacturing cost.
- Safety and maintenance cost = 7% of manufacturing cost
- Total capital= working capital+fixed capital.
- The following items will be considered for calculating the initial investment:
    1. Land cost
    2. Cost of all equipment
    3. Cost of consultant for plant design
    4. Cost of construction of buildings
    5. Cost of setting up utilities
    6. Cost of ensuring electrical connections and power supply.
- The following items will be considered in the working capital category:
    1. Salary of employees
    2. Salary of casual laborers
    3. Cash Reserve
    4. Maintenance and repair
    5. Product and in process inventory

The total cost production is calculated to be the actual cost involved in the manufacturing process summed with the general expenses incurred.
The various types of manufacturing costs are as follows:

1. Raw Materials
2. Utilities
3. Labor and Supervision
4. Maintenance and repair
5. Operating Supplies
6. Taxes and Insurance
7. Plant Overhead
8. Depreciation

The various types of general expenses to be added to the manufacturing cost are as follows:

1. Administrative Expenses
2. Distribution and Selling Cost
3. Interest on capital Investment

The factors to be considered for the economic analysis:

1. Gross Annual Income
2. Annual Cost of Production
3. Gross Profit
4. Income Taxes
5. Net Profit

Th table below shows the delivered equipment cost of equipments that would be used in the plant:

Table 7.1: Approximate installation and maintenance cost of the equipment being used

Equipment	Installation cost(Rs)	Annual Operational cost (Rs)
Furnace	234000000	0
Compressors (4)	520000000	260000000
Distillation columns (5)	1625000000	672750000
Refrigeration setup	520000000	65000000
Quench section	4550000	24050000
Total	2424500000	1021800000

Table 7.2: Approximate Cost of Raw materials

Product	Rate(Rs per kg)	Flow rate (kg/yr)	Total cost($)
Heavy Naphtha	20	605900000	1211800000
Steam	0.5	295650000	14782500
Natural Gas	36	38690000	1,392,840,000
Total			13525622500

Table 7.3: Approximate revenue generated from the major products

Product	Rate(Rs per kg)	Flow rate (kg/yr)	Total revenue ($)
Ethylene	71.55	182500000	13057875000
Methane	35	89425000	3129875000
Total			16187750000

Table 7.4: Approximate cost of other utilities being used

Item	Cost (Per Year)
Process Piping	7359794.00
Electrical Equipment	2886598.00
Instrumentation	5456577.00
Building	5679790.00
Excavation and site preparation	3859794.00
Auxiliaries	10432950.00
Total physical plant	54021960.00
Field Expenses	8536392.00
Engineering	7636292.00
Direct plant	73370105.00
Contractor, Overhead	2256917.00
Contingency	7337732.00
Total fixed investment	90557734.6
Every cost regarding pump	9547392.00
Total	288940028
Total (rounded off)	289,000,000

Table 7.5: Approximate cost of other utilities being used

Staff	No.	Salary per month (Rs.)	Total (Rs.)
Plant Manager	1	200000	200000
Chief Engineer	4	100000	400000
Assistant Engineer	8	75000	600000
Junior Engineer	10	50000	500000
Supervisor	15	35000	525000
Operator	20	25000	500000
Labor	50	10000	500000
Chemist	16	20000	320000
Office Staff	29	25000	725000
Security staff	30	15000	450000
Driver	10	12000	120000
Total			4840000
Annual Total			58080000

The inferences from the analysis of the cost of the project gives us the following highlights:

- The total annual cost for raw materials is: Rs.13,525,622,500
- The total annual operation cost of equipments is: Rs. 1,021,800,000
- The operational cost of utilities is: Rs. 289,000,000
- The total annual amount spend in salary:Rs. 58,080,000
- The total annual revenue obtained from the project is: rs. 16,187,750,000
- The total annual gross profit generated: Rs. 1,345,519,500
- The total annual net profit generated after deduction @ 30% = Rs. 941,863,650
- The total fixed capital at the beginning: Rs. 2,424,500,000
- The pay back period is therefore: 2.57 years.

CONCLUSION

The project reflected the nuances of designing a plant. Even through the design is in the very nascent of its form and no where close in merit with respect to actual plant design, we would like to believe that it is, nonetheless, an appropriate representations of how plant designs should be carried out.

Some of the various short comings faced during the project are as follows:

- Availability of accurate data with respect to the variable used in the plants was very difficult to find.
- A large number of close approximations have been used which makes the project less accurate.
- The design procedure that we are acquainted with at our level of knowledge were not very sufficient to do justice to a complicated plant design.
- Scarcity of availability of concrete data from the industry was also a major hurdle in the successful completion of the project

Some of the takeaways from the experience of the project can be summarized as follows:

- Every plant design project has to involve an extremely rigorous data collection in the first phase of the project
- The plant design should be carried out manually as much as possible as it is, if not anything, a great learning experience for undergraduate students.
- Sometimes it becomes difficult to get past a certain section because of the inherent difficulty of the section. However, in these situations other parts of the project work which have no relation to the part we are stuck in, should be attended to. Parallel flow of task will allow for optimization of time and give time for generation of ideas without stalling the process.

Appendices

GRAPHS USED FOR DESIGN OF HEAT EXCHANGER

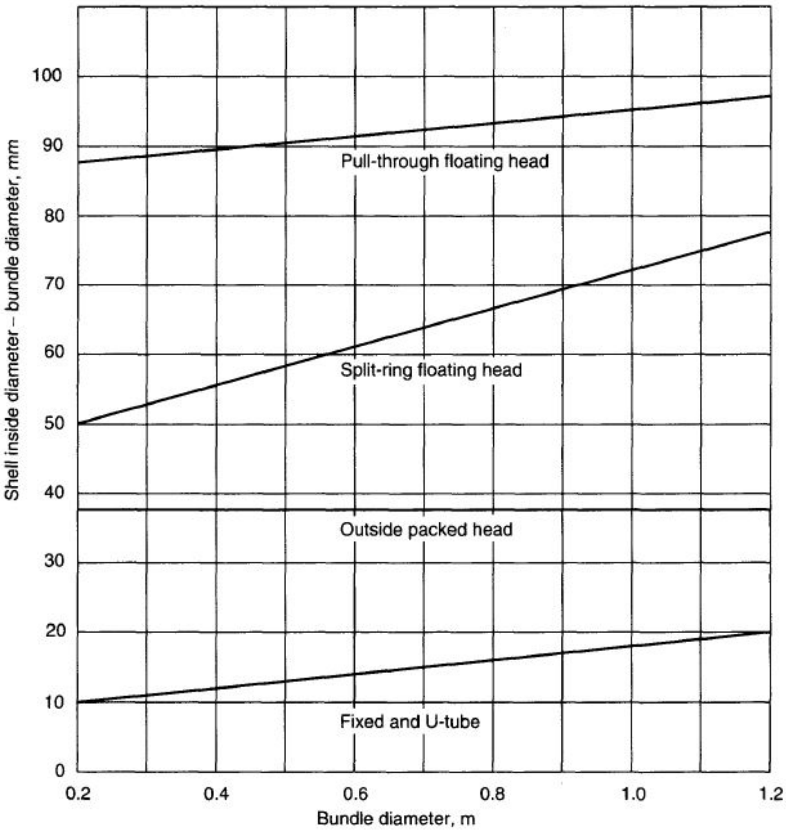
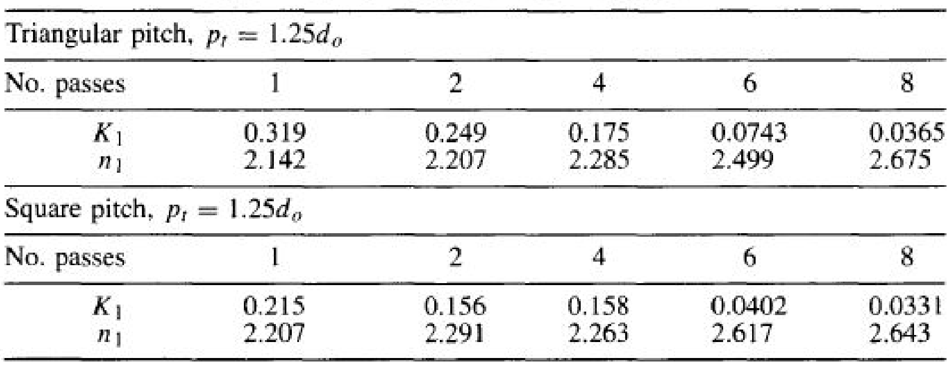


Figure A.1: Finding out clearance using bundle diameter



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 A.2: Table to find the bundle diameter using the number of passes

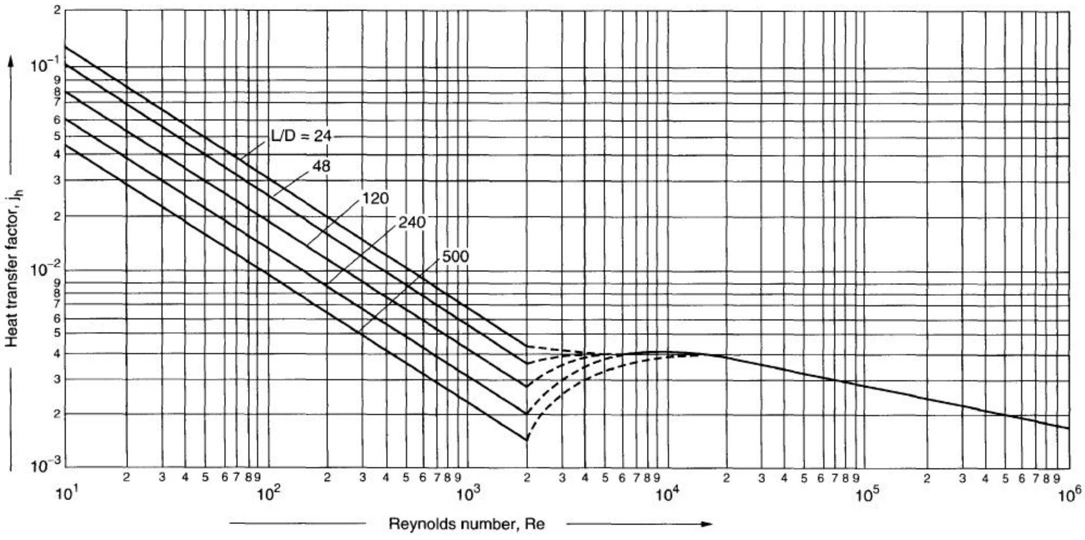


Figure A.3: Chart to find the heat transfer factor on the tube side

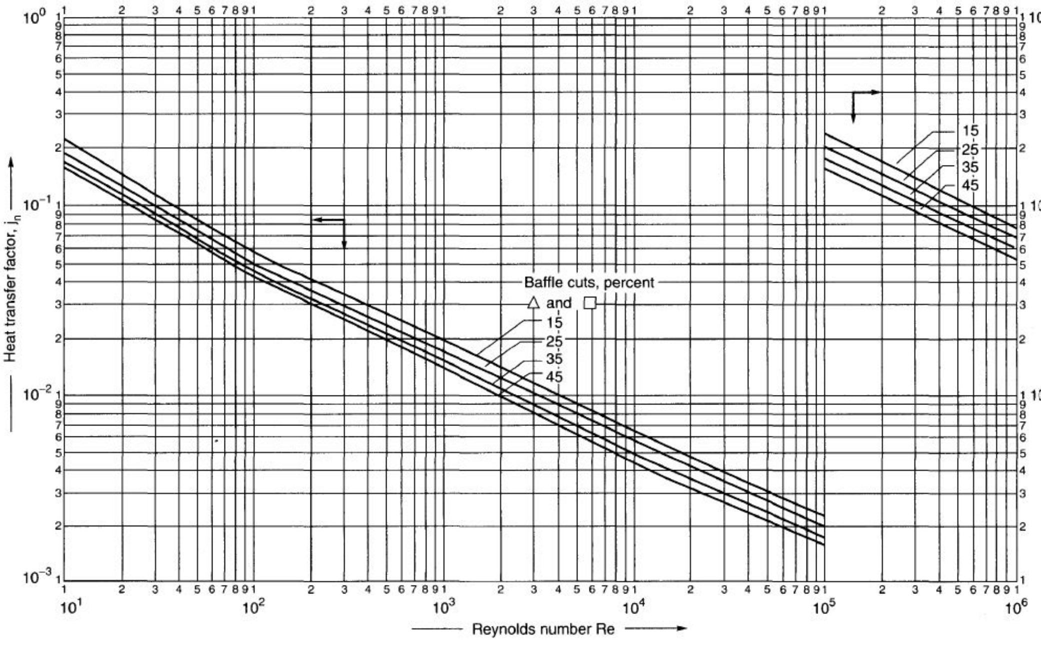


Figure A.4: Chart to find the heat transfer factor on the shell side

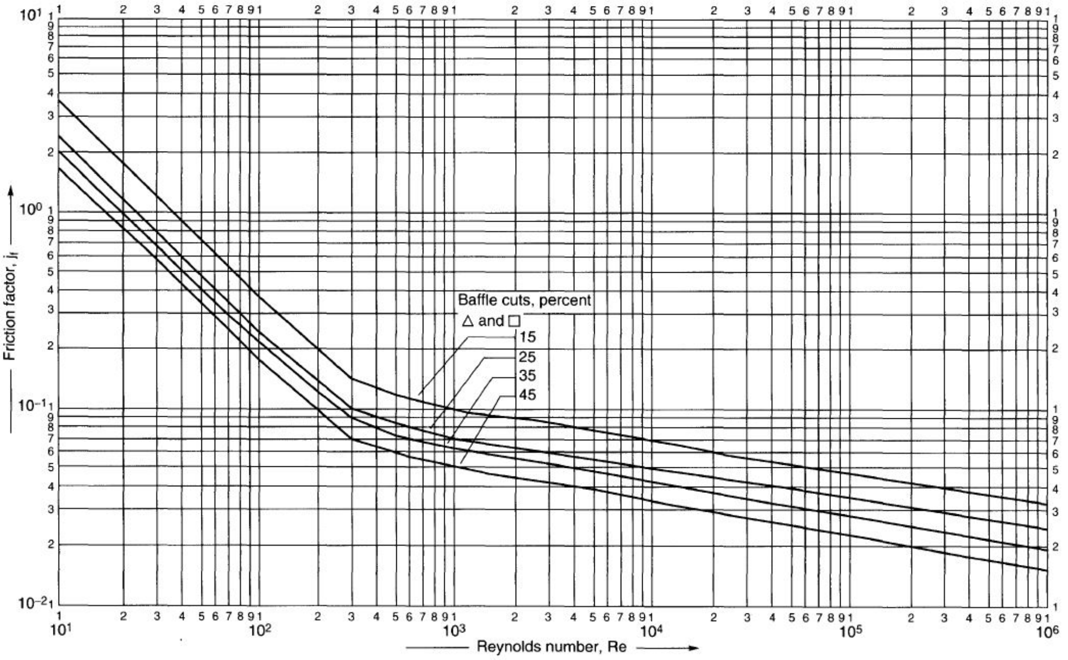


Figure A.5: Chart to find the friction factor on the shell side

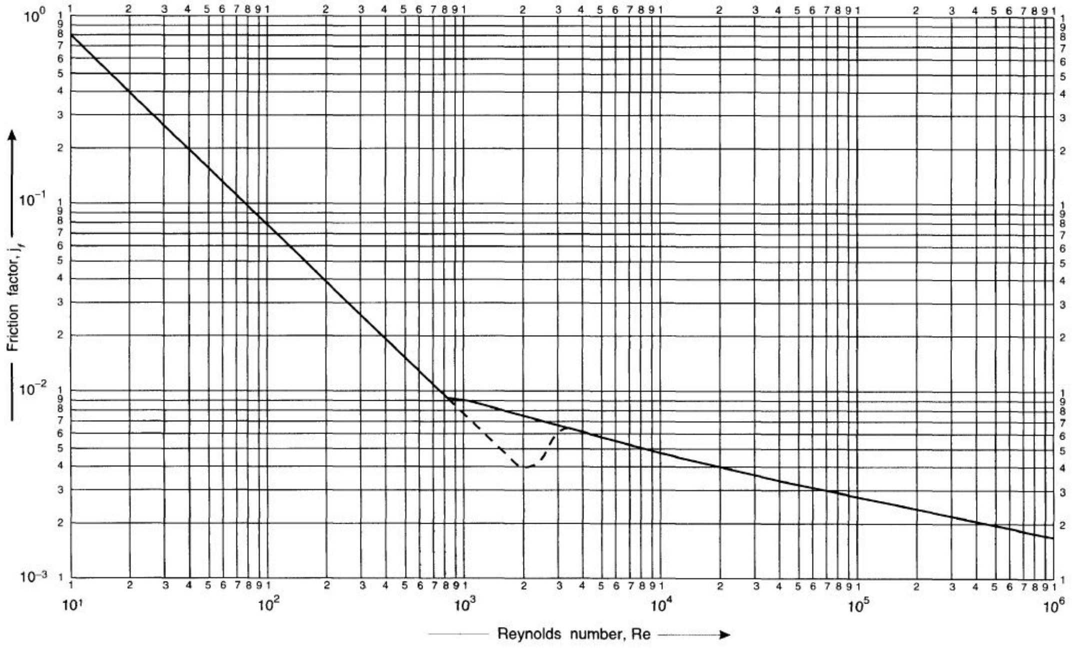


Figure A.6: Chart to find the friction factor on the tube side

GRAPHS USED FOR DESIGN OF DISTILLATION COLUMN

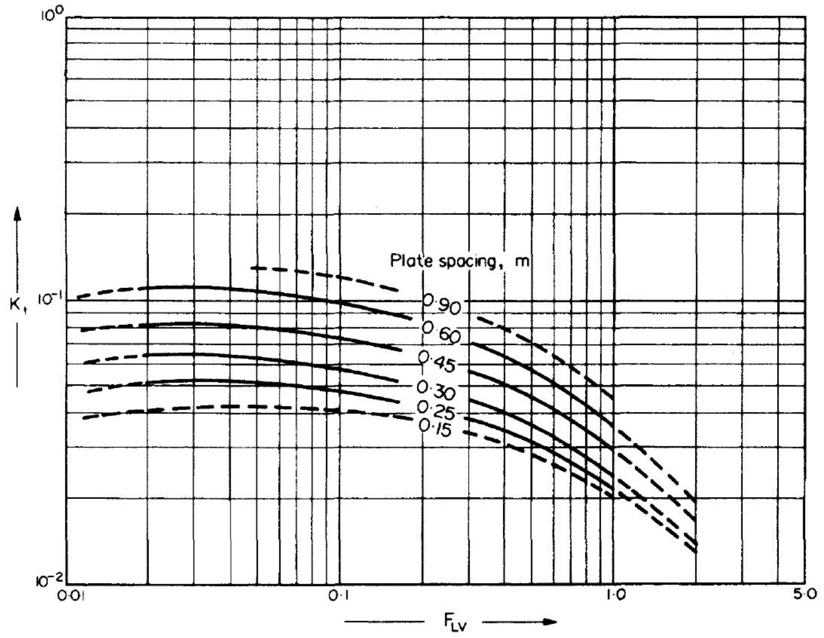


Figure B.1: Chart to find value of K1 based on the tray spacing and value of FLV

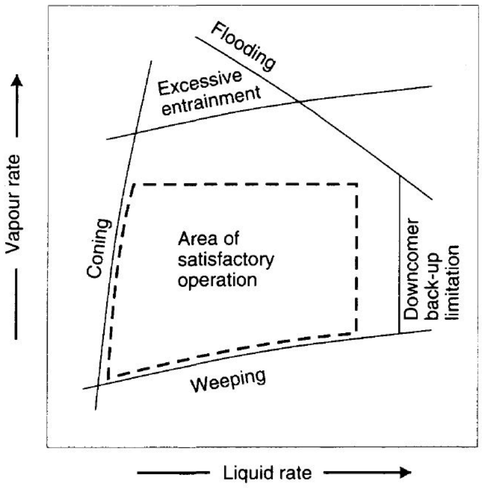


Figure B.2: Chart to find operation range of the distillation column

HEAT EXCHANGER: DETAILED ENGINEERING DRAWING

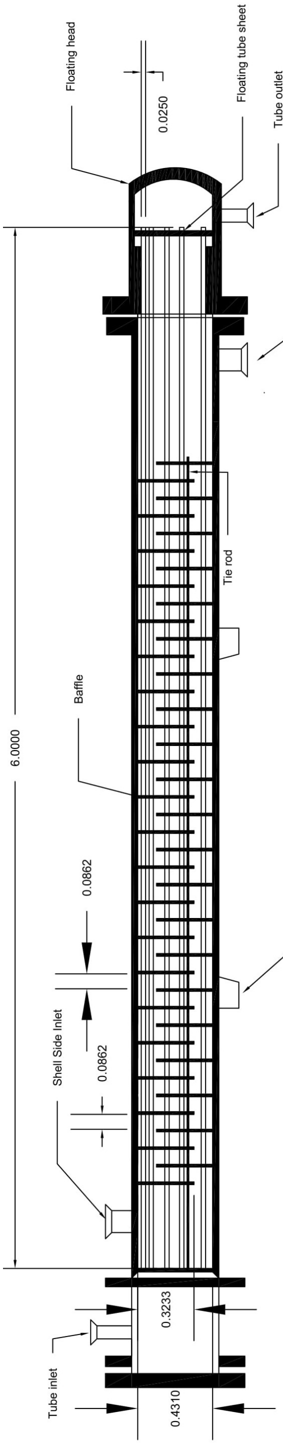


Figure C.1: Designed heat exchanger(all units in metres

FURNACE: DETAILED ENGINEERING DRAWING

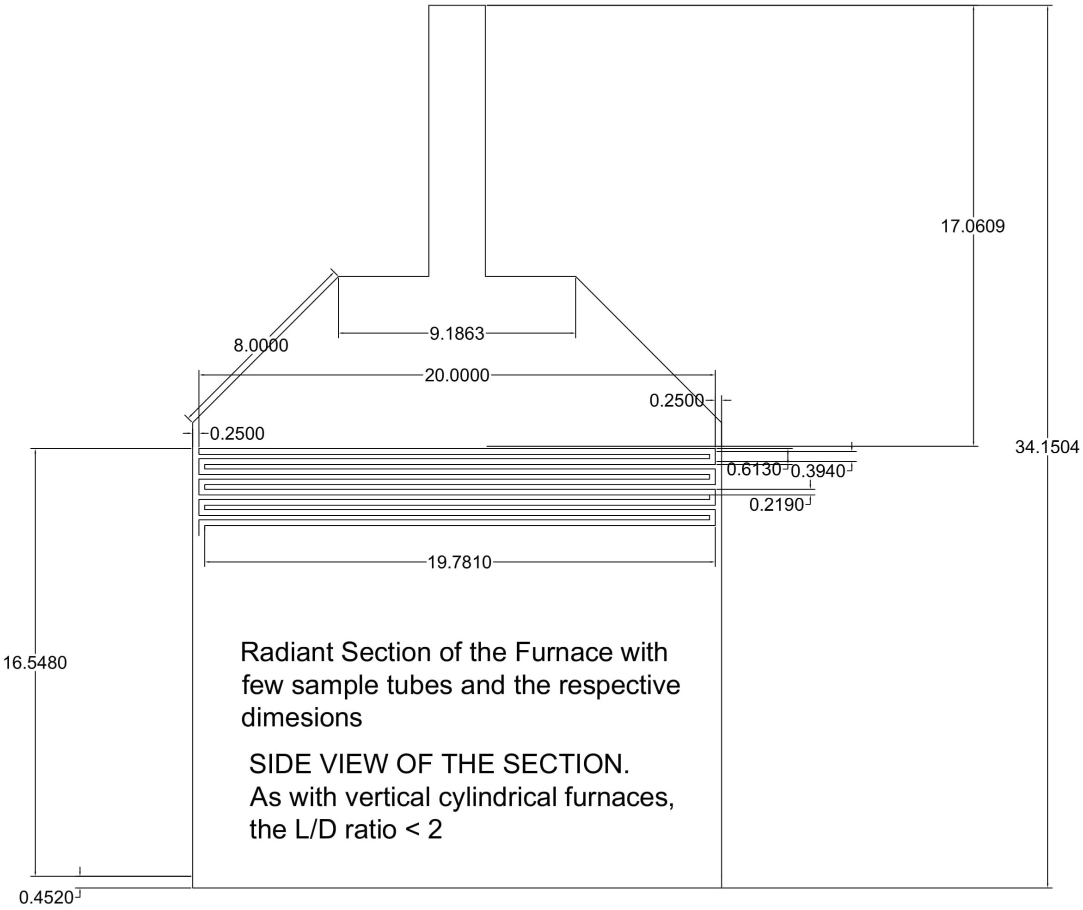


Figure D.1: Designed Furnace (all units in metres)

DISTILLATION COLUMN: DETAILED ENGINEERING DRAWING

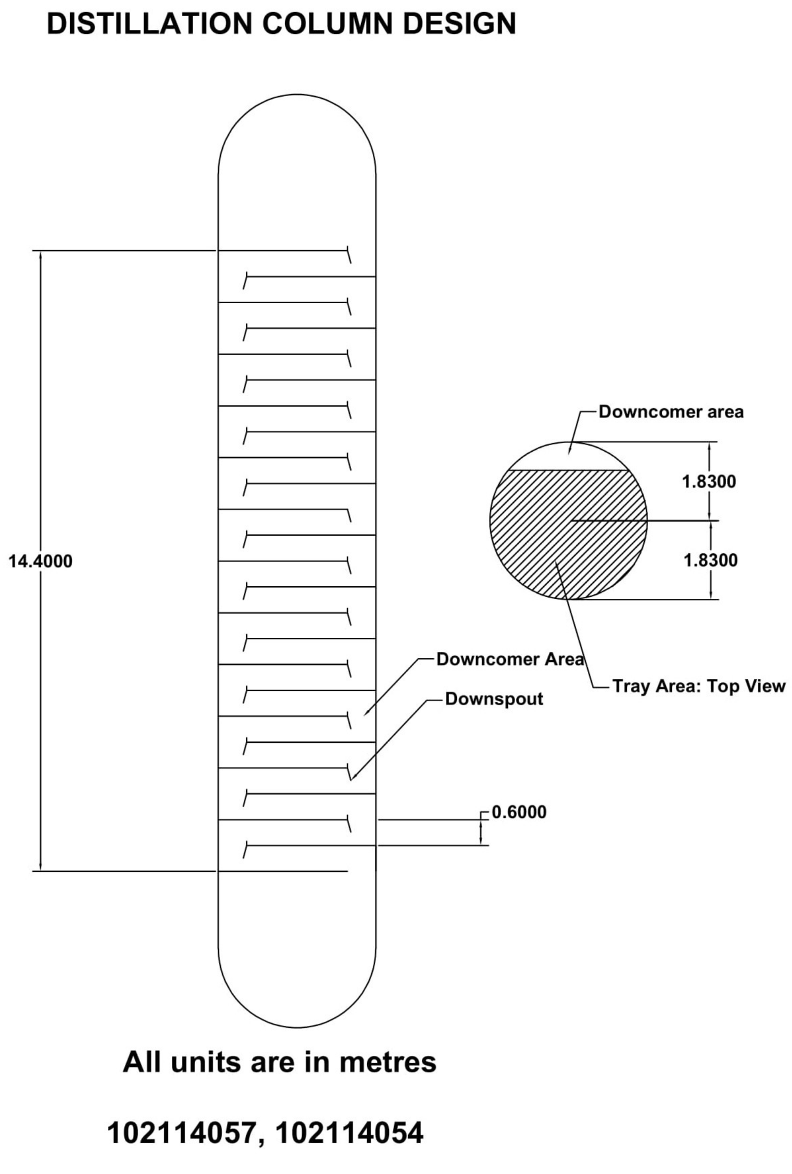


Figure E.1: Designed distillation column

PLANT LAYOUT DIAGRAM

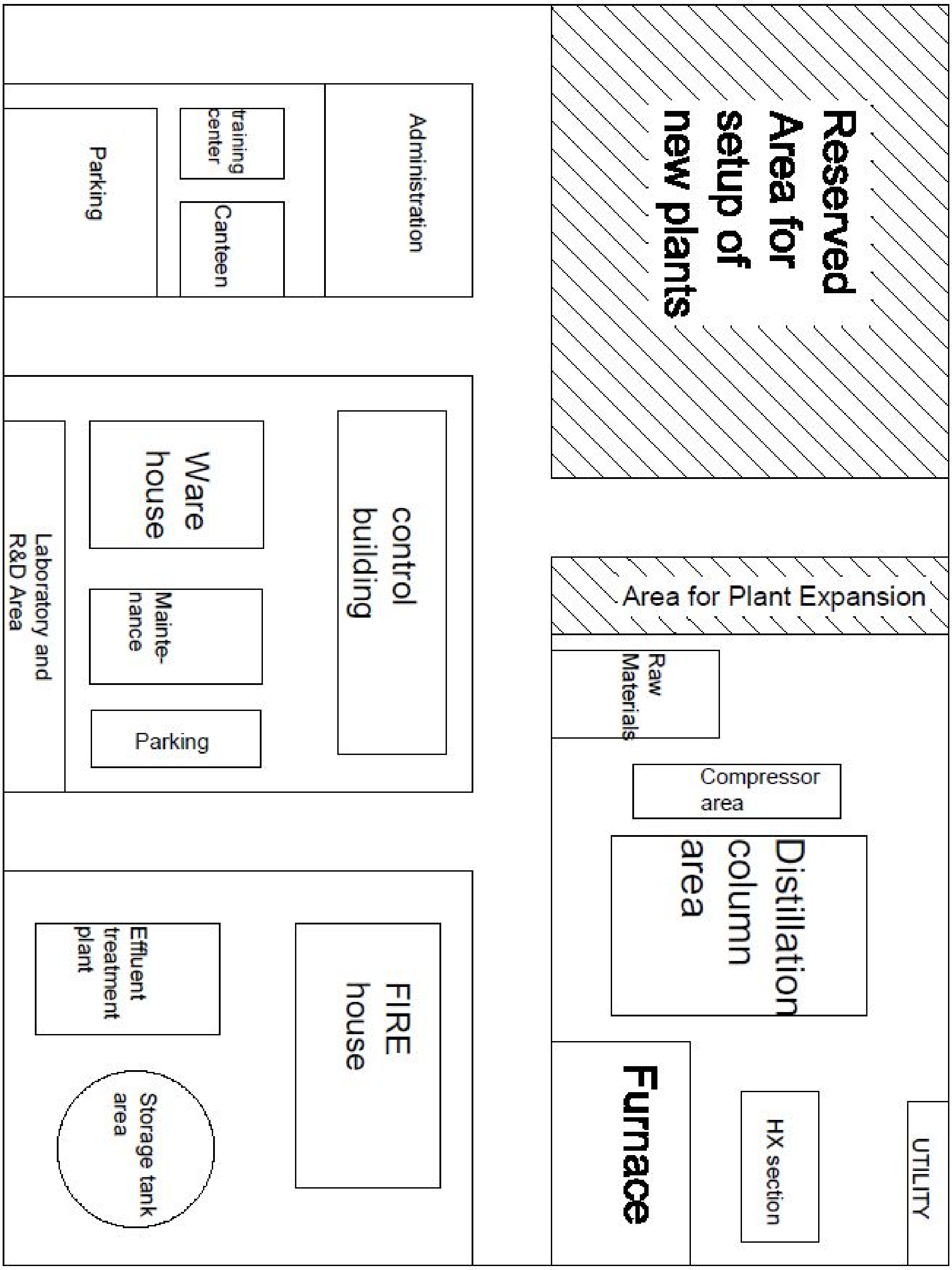


Figure F.1: Plant Layout Design

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