DESIGN OF DISTILLATION COLUMN AND OTHER EQUIPMENT FOR THE PRODUCTION OF FORMALDEHYDE

A Report

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CERTIFICATE

This is to certify that the thesis titled **Design of Distillation Column and Other Equipment for the Production of Formaldehyde** submitted by **Anjali Kumari** ($R900211004$)**, Sonal Bhadauriya** ($R900211028$)** and Prakhar Pathak** ($R900211053$), to the University of Petroleum and Energy Studies, for the award of the degree of **BACHELOR OF TECHNOLOGY** in Chemical Engineering with specialization in Refining & Petrochemicals is a bonafide record of project work carried out by them under my supervision.

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NOMENCLATURE

ξ Moles of Formaldehyde Formed

χ Mole fraction

P   Pressure

T   Temperature

T_c Critical Temperature

P_c Critical Pressure

D   Distillate

B   Bottom

N_i Stream Number

t_r Reduced Temperature

C_P Heat Capacity

ρ_L Density of Liquid

ρ_V Density of Vapour

n_i Number of Moles of Component i

H_f Enthalpy of Feed

H_b Enthalpy of Bottom

H_d Enthalpy of Dstillate

F   Feed

ϕ_b Reboiler Load

ϕ_c Condenser Load

λ Heat of Vaporization

$\dot{m}$ Mass Flow Rate

R   Reflux Ratio

R_m Minimum Reflux Ratio

α Relative Volatility

q   Quality of Feed

ν Constant used in Underwood's Equation

μ Viscosity

v_f Flooding Vapour Velocity

σ Surface Tension

$\dot{Q}$ Volumetric Flow Rate

M_{avg} Average Molecular Weight

θ_{res} Residence Time

ϕ Froth Density at Normal Operating Conditions

U   Overall Heat Transfer Coefficient

k   Thermal Conductivity

B_s Baffle Spacing

C_{OL} Cost of Operating Labour

C_{OM} Cost of Manufacturing

ABSTRACT

This report deals with the designing of Distillation Column and other equipment like reboiler, condenser and heat exchanger aiding the production of Formaldehyde. A plant of capacity 60000 TPA is chosen. Entire Mass Balance and Energy Balance is done for the plant. After finding out the molar flow rates through each stream and heat duties required, the distillation column is designed. In the Distillation Column, minimum Reflux Ratio, minimum Number of Trays, Optimum Feed Stage and Tray design and layout is done. Reboiler and Condenser are also designed. Shell side and tube side Transfer coefficients and Pressure drops are analyzed. In the end, we have tried incorporating the cost estimation and Economics of the plant.

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1   Introduction & Motivation

Chemical process and plant design involves a hierarchy of complex and creative activities including both routine and innovative design. Routine design is largely analytical and is primarily concerned with determining values for the specification and operation of specific units such as reactors, extractors and distillation columns. By contrast, process synthesis, the generation of conceptual flow sheets comprising such units, is a more open-ended activity characterized by a combinatorial large number of feasible alternatives. Finding better flow sheet alternatives has a significant impact on overall process competitiveness. This is a particularly true for separation process synthesis, the selection of separation methods, their inter connection, and their operating parameters. Virtually every chemical process involves the recovery, isolation or purification of products, by products, intermediates, wastes or raw materials [14]. The separation systems to accomplish these tasks often dominate the capital and operating costs of chemical manufacturing processes.

Several equipment approaches to the separations process synthesis problems have been formulated including super structure optimization, evolutionary modification and systematic generation from a known feed composition, desired product compositions and well defined set of separation methods, super structure optimization approaches constructive hypothetical flow sheet which includes all applicable separation methods interconnected in every possible order so as to include all possible separation schemes. Separation synthesis then becomes a problem of systematically stripping away the less desirable parts of this superstructure, simultaneously optimizing the design in operating parameters of the remaining separation methods.

Evolutionary modification starts with an existing flow sheet for a similar separation to which the adaptation is made as necessary to meet the objectives of the specific case at hand. This approach is exemplified by standard flow sheet pattern, for example the schemes for breaking heterogeneous minimum boiling azeotropes for the sequence for extractive distillation, although rarely resulting in novel designs, evolutionary modification is a frequently used separations synthesis technique.

Formaldehyde, the target product of the projects plant, is an organic compound representing the simplest form of the aldehydes. It acts as a synthesis baseline for many other chemical compounds including phenol formaldehyde, urea formaldehyde and melamine resin. The most widely produced grade is formalin (37 wt. percent formaldehyde in water) aqueous solution.

This project deals with a comprehensive study of the Formaldehyde Producing Process using Methnaol that would lead to the design of the main constituent of the plant ; Distillation Column. The plant produces formaldehyde with a production capacity of 60000 TPA. This study will take into consideration various aspects including the entire plants process unit design, process flow diagrams, cost estimations, operation parameters, equipment sizing, construction materials and environment/safety precautions focussing on Distillation Coumn. This project requires the theoretical and practical application of mass transfer, heat transfer, fluid dynamics, unit operations, and reaction kinetics and process control. There are several tasks that are crucial to the completion of the project outlines including mass and energy balances, design of heat exchangers and distillation column and economic analysis.

2 Literature Review

The main purpose of this project is to conduct a comprehensive study that would lead to the design of the main constituent of the plant ; Distillation Column. This study will take into consideration aspects including the entire plants process unit design, process flow diagrams, cost estimations, operation parameters, equipment sizing, construction materials and environment/safety precautions. This project requires the theoretical and practical application of mass transfer, heat transfer, fluid dynamics, unit operations, and reaction kinetics and process control.

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2.1 Alternative Processes

Formaldehyde is to be commercially manufactured on an industrial scale from methanol and air in the presence of a sliver catalyst or the use of a metal oxide catalyst. Alternative processes for production of formaldehyde were studied which are as follows [2] :

1.Metal oxide catalyst process: In this process catalysts like vanadium pentoxide and iron-molybdenum oxide were put to use for the reaction to take place. The iron-molybdenum oxide catalyst makes formaldehyde from the exothermic reaction entirely. The disadvantage of this process design is the need for significantly large equipment to accommodate the increased flow of gases (3 times larger) compared to the original silver catalyst process design. This increase in equipment sizing clashes with economic prospect behind the design costs.

2.Production of Formaldehyde from Methane and Other Hydrocarbon Gases: Production of formaldehyde is through the oxidation of hydrocarbon gases. An increase in the amount produced of formaldehyde is expected in this process. However, the hydrocarbon formaldehyde is usually obtained as dilute solution which is not economically concentrated accompanied by other aldehydes and by-products. The difficulty in this method is in controlling the oxidation of reaction.

2.2 Physical and Chemical Properties

These are the various physical and chemical properties [4]:

Table 1: Physical and Chemical Properties of various Components

Name	Formula	Molecular Weight (kg/kmol)	Boiling Point (C)	λ(KJ/mole)
Methanol	CH_3OH	32.042	64.7	35.27
Oxygen	O_2	31.999	-183	6.82
Air	..	28.851	-194.5	..
Formaldehyde	HCHO	30.026	-19.3	24.48
Hydrogen	H_2	2.016	-252.7	0.904
Water	H_2O	18.015	100	40.656
Formalin	HCHO	30.03	96	..
Silver	Ag	107.8682	1950	1950

Methanol

Its a light, volatile, colorless, clear and flammable liquid. It has a distinctive sweetish smell and close to alcohol in odor and colorlessness. Inhalation of methanol is much more hazardous than mere contact. Due to its low flash point, it forms an explosive mixture with air. The material should be stored in cooled well-ventilated isolated areas [19].

Formalin

This material is a highly toxic material that the ingestion of 30 ml is reported to cause fatal accidents to adult victims. Formaldehyde is flammable in the presence of sparks or open flames. Pure Formaldehyde is not stable, and concentrations of other materials increase over time including formic acid and para formaldehyde solids[10]. A gas at room temperature, formaldehyde is colorless and has a characteristic pungent, irritating odor. Stabilizers for formaldehyde product include hydroxypropyl methylcellulose, Methyl cellulose.

2.3 World Formaldehyde Production & Indian Scenario

In 2012, the global production capacity of formaldehyde surpassed the 46.4 million tones mark. In the same year, the Asian-Pacific region held a share of 56% of the worlds total formaldehyde capacity. It was followed by Europe and North America, accounting for 22% and 15.83% shares, respectively. China was an unrivalled leader in terms of formaldehyde capacity, accounting for over 51% of the total capacity.
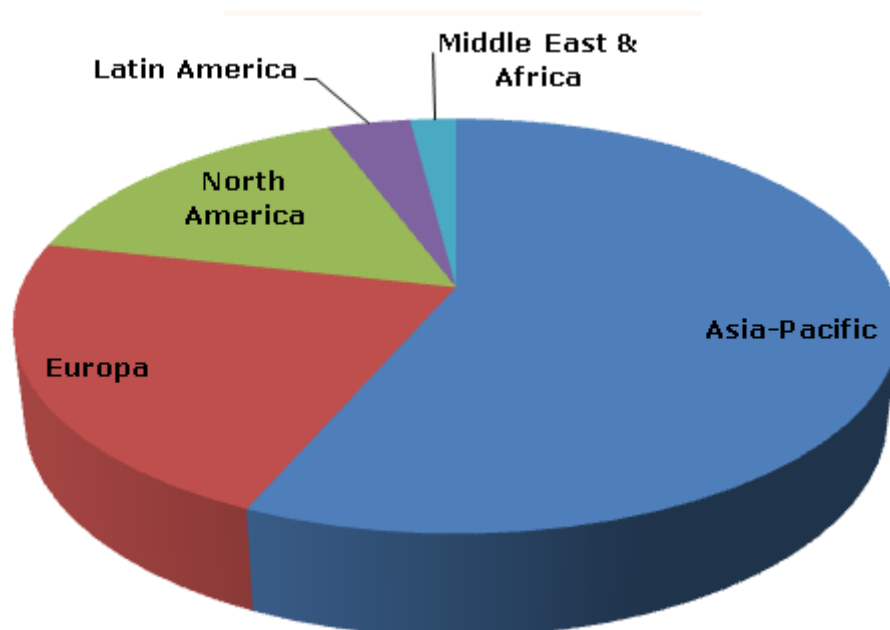
Figure 1: Structure of the world Formaldehyde capacity broken down by region, 2012 [2, 4]

Indian Scenario

Kanoria Chemicals and Industries is the largest producer of formaldehyde in India, with installed capacity of 1.8 lacs TPA(2012). Allied Resins and Chemicals has the second largest formaldehyde producer with installed capacity of 60000 TPA. Hindustan Organic Chemicals, with 30000 TPA capacities, standing third largest Indian formaldehyde producer. Atul Limited has an installed capacity of 20000 TPA of formaldehyde. Venlon Enterprises has installed capacity of 12000 TPA of formaldehyde. Following table gives the capacity of various plants installed in India.

Table 2: Formaldehyde Plants in India

Company	Percent of Total Turnover	Installed Capacity	Sales
Kanoria Chemicals & Industries Ltd.	8	180000	39.53
Allied Resins & Chemicals Ltd.	8	60000	5.48
Hindustan Organic Chemicals Ltd.	4	33000	28.21
Atul Ltd.	1	20000	12.71
Asian Paints	0.6	8100	1.3
Venlon Enterprises	9	12000	10.68
Century Laminating	1	9000	2.05

2.4 Preliminary Cost of Materials

An approximate order of magnitude preliminary analysis of some of the raw materials was done. Following table shows some of these values [11, 4]:

Table 3: Preliminary Cost of Materials (INR*=Indian National Rupee)

Material	Preliminary Cost
Methanol	15250-30500 INR*/ Metric Ton
Formalin	23180-51118 INR*/ Metric Ton
Silver	61000-183000 INR*/ Kilogram
Hydrogen	1830-6100 INR*/ 40 L Cylinder
Deionized Water	1.61 INR*/ L

2.5 Uses

Formaldehyde is mainly used on the construction, automotive and furniture markets. It used to manufacture resins; as an intermediate for synthesizing other chemicals and is also directly used without further processing. Formaldehyde is commercially offered as a 37% to 50% aqueous solution, with 37% being the most widely used grade, which may also contain 0-15% methanol and a polymerization inhibitor. Because of this, the formaldehyde business is characterized by user facilities located near production facilities [19].

Formaldehyde is mainly used as resins, an intermediate and direct use:

Resins
The largest amount of formaldehyde finds application in the manufacture of Resins, viz., Phenol Formaldehyde (PF), Urea Formaldehyde (UF) and Melamine Formaldehyde (MF) - which finds applications in laminates, plywood, MDF, particle board and hard boards. UF, MF and PF resins accounted for approximately 66to produce curable molding materials, as raw materials for surface coatings, as binders for foundry sand

Intermediate

Formaldehyde is also used for synthesizing chemicals like:

Penta-erythritol, Hexamine, Para formaldehyde, 1,4-Butanediol, Polyacetal Resins, Methylene Diplineylene Di-isocynates (MDI), Trimethylolpropane, Neopentylglycol.

Direct Use

Formaldehyde is also used directly for the following:

Mirror finishing and electroplating. Preservation and disinfection. Film development in photography industry

3 Objectives

1. To design a distillation column pertaining to the given industrial problem.

   - Vessel Design
   - Tower Internals

2. Designing of other equipment aiding the production of formaldehyde.

   - Reboiler
   - Condenser
   - Heat Exchanger

3. To conduct an Order-of-the-Magnitude Cost Estimation Survey of the plant

4   Methodology

The design revolves around the production of formaldehyde from methanol. It acts as a synthesis baseline for many other chemical compounds including phenol formaldehyde, urea formaldehyde and melamine resin. The molecular weight, density, melting point, boiling point, and heat capacities are taken in account to establish the mass and heat balance equations. This separation is done in a tray column for which different tray types will be analyzed. Some additional original data on semi-technical columns will also be analyzed and further utilized to attain the goal. After all these data acquisition, vapour-liquid equilibrium characteristics [17] (indicated by the shape of the equilibrium curve) of the mixture will determine the number of stages required for the separation. The number of trays required to achieve the given separation also depends upon the amount of reflux used. Using more reflux decreases the number of plates required and using less reflux increases the number of plates required. Also the tower internals will be designed according to the given operating conditions. Heat exchangers are also designed for optimum production with low wastage of energy.

Further the work may be extended to design other units along with this to achieve overall production of the chemical considered(formaldehyde).

5   Process Flow Diagram
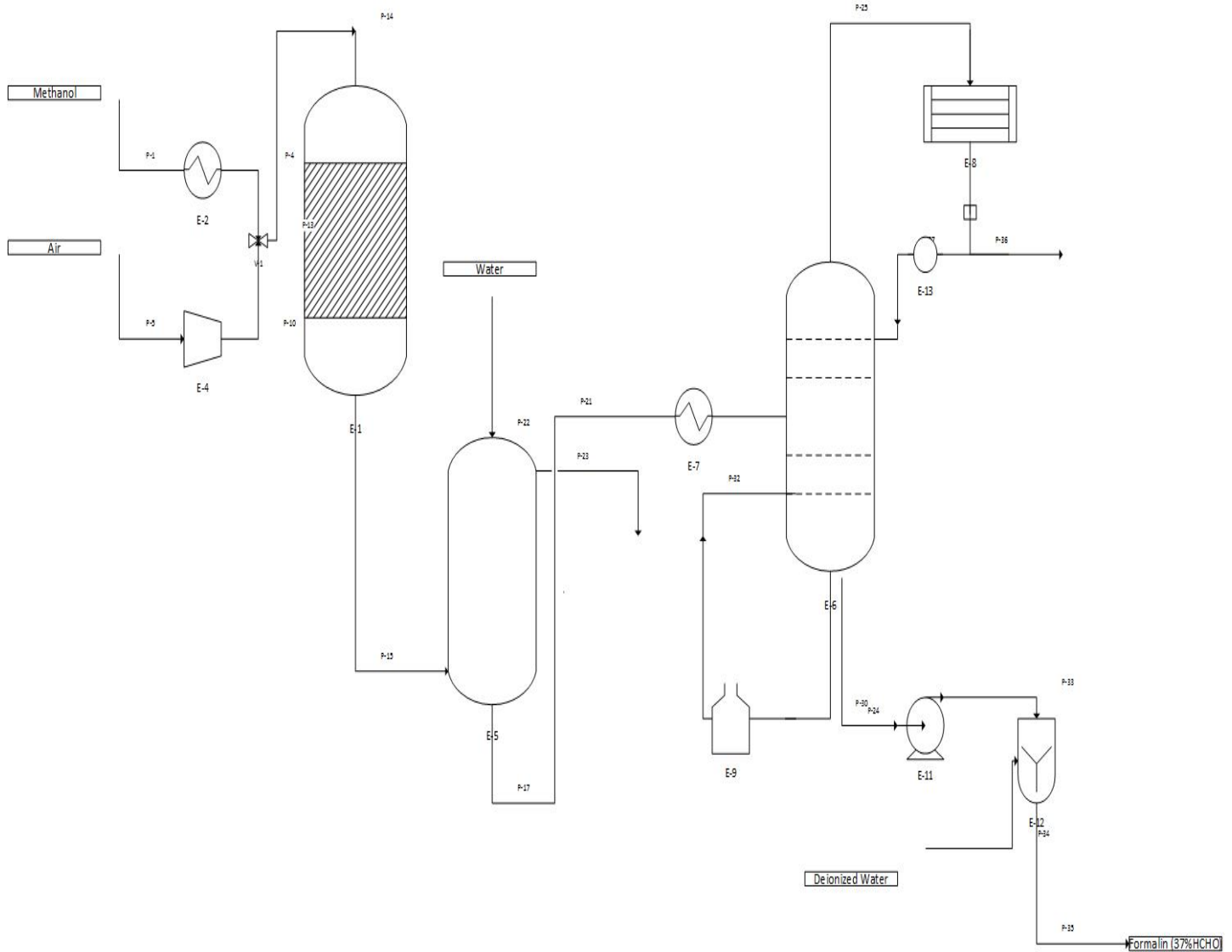
Figure 2: Process Flow Diagram

The Process Flow Diagram showing the overall layout of the plant. The flowsheet has been prepared using Microsoft Visio [15].

6 Mass Balance & Energy Balance

This is a full detailed chapter presenting the Mass Balance and Energy Balance for the projects plant of producing formaldehyde from methanol. The analysis and calculations were done manually.

6.1 Mass Balance

Mass Balance around the entire plant was carried out. All required parameters from the problem statement including; conversion, selectivity, temperature, pressure and production capacity were implemented in the mass balance. The following process flow diagram (PFD) of the formaldehyde plant is the reference for unit designation and stream numbering. The methanol feed input is the basis of calculation throughout the chapter. The amount of input basis of methanol was $n_3 = 10 kmol/hr$.

Information provided in the statement problem:

- Overall conversion of methanol: 0.874

- Selectivity of desired reaction to undesired reaction $= 9$

- Production of formaldehyde needed $= 60000$ ton per year

- The inlet temperature to the reactor 343 C

- The outlet temperature from the reactor 200 C

- Recycled temperature and pressure is 68.3 C and 1.2 atm respectively.

- Pressure of the absorber is 138 KPa with formaldehyde absorption recovery of 99%.

- Exist liquid stream from absorber is heated to 102 C.

Definitions of all abbreviations used in our calculations:

- N: Molar flow-rate (Kmol/hr.)

- M : Methanol

- W: Water

- H_2: hydrogen

- N_2: nitrogen

- F : Formaldehyde

- O_2 : Oxygen

- χ , y : Mole fraction

- ξ_1 : Moles of Formaldehyde formed in reaction 1

$$CH_3OH + 0.5O_2 \rightarrow HCHO + H_2O$$
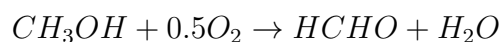
- ξ_2: Moles of Formaldehyde formed in reaction 2, Side Reaction

$$CH_3OH \rightarrow HCHO + H_2$$

- N_M: Methanol flow rate, similarly for the rest components.

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6.1.1   First Run

- **Mass Balance around the Reactor**

A heated mixture of Methanol and air (N8) is sent to the reactor at 343 C. Two reactions occur:

$$CH_3OH + 0.5O_2 \rightarrow HCHO + H_2O$$

Desired Reaction

$$CH_3OH \rightarrow HCHO + H_2$$
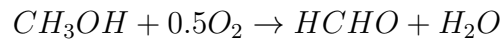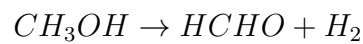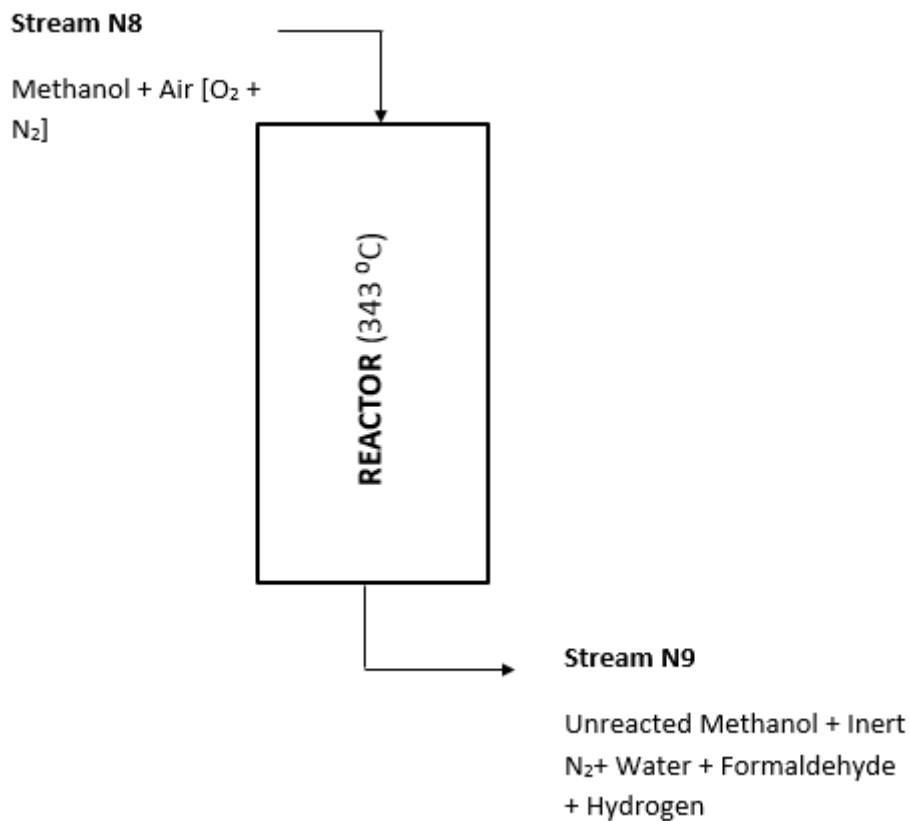
SideReaction

Figure 3: Mass Balance around the Reactor

After the reaction, we will get stream N9 which is a mixture of unreacted methanol and oxygen, Formaldehyde, Hydrogen and Inert nitrogen. We are considering all the oxygen gets consumed in the reaction.

Applying Material Balance:

$$N_{M9} = N_{M8} - \xi_1 - \xi_2 \tag{1}$$

$$N_{O_29} = N_{O_28} - 0.5\xi_1 \tag{2}$$

$$N_{H_29} = \xi_2 \tag{3}$$

$$N_{H_2O9} = \xi_1 \tag{4}$$

$$N_{F9} = \xi_1 + \xi_2 \tag{5}$$

Now, conversion is 0.874 and selectivity is 9.

From (1) :

$$\frac{N_{M8} - N_{M9}}{N_{M8}} = \frac{\xi_1 + \xi_2}{10} = 0.874$$

$$\xi_1 + \xi_2 = 10 * 0.874 = 8.74 \tag{6}$$

Selectivity= 9

Therefore,

$$\frac{\xi_1}{\xi_2} = 9 \tag{7}$$

Solving (6) and (7) :

$$\xi_1 = 7.866 kmol/hr$$
$$\xi_2 = 0.874 kmol/hr$$

Thus,

$$N_{M9} = 10 - 7.866 - 0.894 = 1.26 kmol/hr$$
$$N_{O_28} = 0.5 * 7.866 = 3.933 kmol/hr$$
$$N_{N_28} = N_{N_29} = \frac{N_{O_28}}{0.21} * 0.79 = 14.796 kmol/hr$$
$$N_{H_29} = \xi_2 = 0.874 kmol/hr$$
$$N_{H_2O9} = \xi_1 = 7.866 kmol/hr$$
$$N_{F9} = \xi_1 + \xi_2 = 8.74 kmol/hr$$

Composition of Stream 8:

Total Moles=$N_{M8} + N_{O_28} + N_{N_28} = 10 + 3.933 + 14.796 = 28.729 kmol/hr$

Mole Fractions:

$$\chi_{M8} = \frac{10}{28.729} = 0.3481$$

$$\chi_{O_28} = \frac{3.933}{28.729} = 0.1361$$

$$\chi_{N_28} = \frac{14.796}{28.729} = 0.515$$

Composition of Stream 9:

Total Moles=$N_{M9} + N_{F9} + N_{H_29} + N_{N_29} + N_{H_2O9}$

$$= 1.26 + 8.74 + 0.874 + 14.796 + 7.866 = 35.536 kmol/hr$$

Mole Fractions:

$$\chi_{M9} = \frac{1.26}{33.536} = 0.0375$$

$$\chi_{F9} = \frac{8.74}{33.536} = 0.261$$

$$\chi_{H_29} = \frac{0.874}{33.536} = 0.0261$$

$$\chi_{N_29} = \frac{14.796}{33.536} = 0.441$$

$$\chi_{H_2O9} = \frac{7.866}{33.536} = 0.234$$

- **Mass Balance around the Absorber**

  Stream N9 after throttling becomes N10 (165 C). It goes to the absorber where the recovery of Formaldehyde is 99%. The solvent used is pure water (N11) at 14 C. It absorbs all the methanol and 99 % Formaldehyde. The solubility of formaldehyde and methanol in water is evaluated at 89.5 C (Average of the two incoming streams). Solubility of Formaldehyde at 89.5 C= 0.78 kg Formaldehyde in 1 kg water

  In kmol Formaldehyde per kmol water units:

  $$= \frac{\frac{0.78 kg formaldehyde}{30.02}}{\frac{1 kg water}{18}} = 0.468 kmol formaldehyde/kmol water$$

  This means 0.468 kmol of formaldehyde gets absorbed by 1 kmol of water. Hence, to absorb 99% of 8.74 kmol/hr (from stream 10) of Formaldehyde we require=$\frac{8.74}{0.468} = 18.675 kmol water/hr$

  Threfore $N11$ (Minimum)= $18.675 kmol/hr$ Now, Solubility of Methanol at 89.5 C is 0.20259 kmol Methanol/kmol of water

  For 18.675 Kmol of Water/hr, Methanol absorbed = $0.20259 * 18.675 = 3.78 Kmol/hr$.
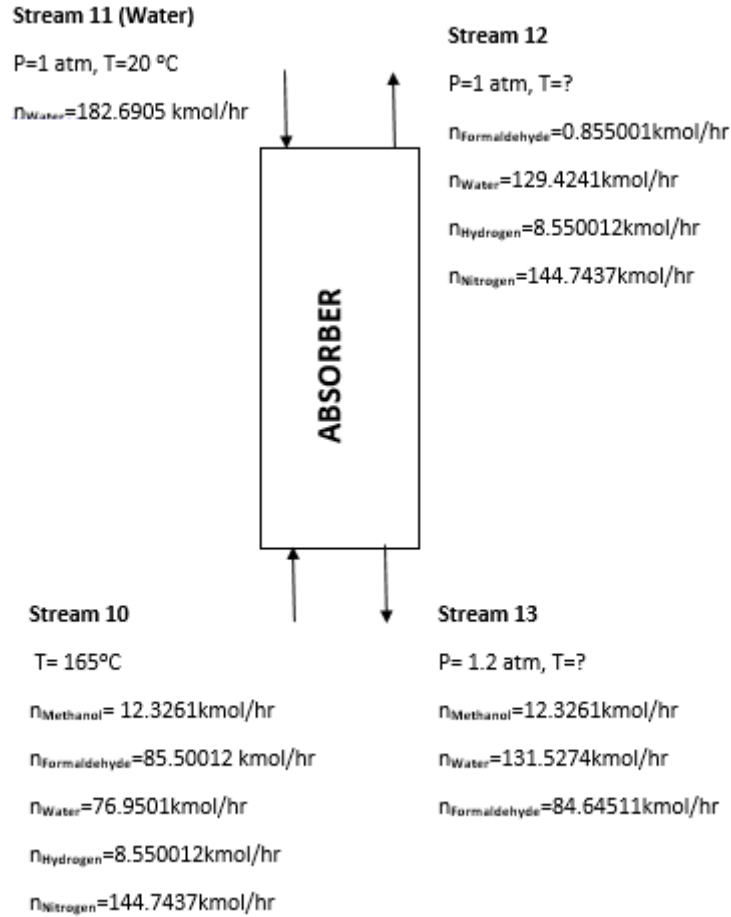
Figure 4: Mass Balance around the Absorber

Therefore, all methanol will get absorbed by N11 $(18.675 Kmol/hr)$. All Methanol will dissolve in water and NO Methanol in the off-gas. Therefore, Stream 12 will contain methanol, formaldehyde, Nitrogen and water.

$$N_{N_2 12} = N_{N_2 10} = 14.796 kmol/hr$$

$$N_{H_2 12} = N_{H_2 10} = 0.874 kmol/hr$$

$N_{F12} = N_{F10} * 0.01 = 0.0874 kmol/hr$ (Amount of Formaldehyde not absorbed)

For water, Assuming Raoults Law to be applicable:

$$y_{H_2O,12} = \frac{P^{vapor}(89.5C)}{P^{Total}}$$

Vapor Pressure of Water at 89.5 (From Antoines Equation):

$$P^{vapor} = exp[16.3872 - \frac{3885.7}{89.5 + 230.17}] = 68.8 kPa$$

$$y_{H_2O} = \frac{68.8}{138} = 0.496$$

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The amount of water going to the vapour phase (N12) will be 0.496 of the total moles of water present in the absorber. Therefore,

$$N_{H_2O,12} = (18.675 + 7.866) * 0.498 = 13.164 kmol/hr$$

Hence, $N_{H_2O,13} = (18.675 + 7.866) - 13.164 = 13.378 kmol/hr$

Also,

$$N_{M13} = 1.26 kmol/hr$$

$N_{F13} = 0.99 * 8.74 = 8.6526 kmol/hr$ (All the methanol goes to N13)

Total number of moles in Stream 12: $N_{F12} + N_{N_212} + N_{H_212} + NH_2O, 12$

$$= 0.0874 + 14.796 + 0.874 + 13.164 = 28.9214 kmol/hr$$

Total number of moles in Stream 13: $N_{F13} + N_{M13} + N_{H_2O,13}$

$$= 8.6526 + 1.26 + 13.378 = 23.29 kmol/hr$$

Composition of Stream N10 (Gas Inlet Stream):

Total Moles= $33.536 kmol/hr$ (Same as N9)

$$\chi_{M10} = \frac{1.26}{33.536} = 0.0375$$

$$\chi_{F10} = \frac{8.74}{33.536} = 0.261$$

$$\chi_{H_210} = \frac{0.874}{33.536} = 0.0261$$

$$\chi_{N_210} = \frac{14.796}{33.536} = 0.441$$

$$\chi_{H_2O10} = \frac{7.866}{33.536} = 0.234$$

Composition of Stream N11 (Liquid Inlet Stream):

Total Moles= $18.675 kmol/hr$

Mole Fraction=1 (Pure)

Composition of Stream N12 (Gas Outlet Stream):

Total Moles= $28.9214 kmol/hr$

$$\chi_{F12} = \frac{0.0874}{28.9214} = 0.003022$$

$$\chi_{N_212} = \frac{14.796}{28.9214} = 0.511$$

$$\chi_{H_212} = \frac{0.874}{28.9214} = 0.302$$

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$$\chi_{H_2O12} = \frac{13.164}{28.9214} = 0.455$$

Composition of Stream N13 (Liquid Outlet Stream):

Total Moles= $23.29 kmol/hr$

$$\chi_{F13} = \frac{8.6526}{23.29} = 0.371$$

$$\chi_{M13} = \frac{1.24}{23.29} = 0.0541$$

$$\chi_{H_2O13} = \frac{13.378}{23.29} = 0.574$$

- **Mass Balance around the Distillation Column**

  The stream liquid stream N13 coming out of the absorber is preheated (becomes N14, composition is considered to be same) to be fed to the Distillation Column at 102 C . Some of the assumptions made are the methanol is considered to be the Light-key component and water as the Heavy-Key Component. The fractional recovery of methanol is 99.7% (In Distillate) and that of water is 99% (Present in the Bottom with the Formaldehyde). Remaining amounts of the methanol and water goes to the bottom and distillate respectively.

  Distillate, $D = N15$

  Bottom,$B = N17$

  Now,
  $$D_M = 0.997 * \text{Moles of Methanol Present in } N14$$
  $$D_M = 0.997 * 1.26 = 1.25622 kmol/hr$$

  Hence, $B_M = 1.26 - 1.25622 = 0.0038 kmol/hr$ Now, calculation for the bottom

  $$B_{H_2O} = 0.99 * \text{Moles of Water present in } N14$$

  Hence, $D_{H_2O} = 13.378 - 13.244 = 0.1338 kmol/hr$

  $$B_F = \text{ Same as N14} = 8.6526 kmol/hr$$

Total Distillate N15$=D_M + D_{H_2O}$

$$= 1.25622 + 0.1338 = 1.39 kmol/hr$$

Mole Fractions:

$$y_M = \frac{1.25622}{1.39} = 0.903$$

$$y_{H_2O} = \frac{0.1338}{1.369} = 0.096$$

Total Bottom N17$=B_F + B_{H_2O} + B_M$

$$= 8.6526 + 13.244 + 0.0038 = 21.9 kmol/hr$$

15
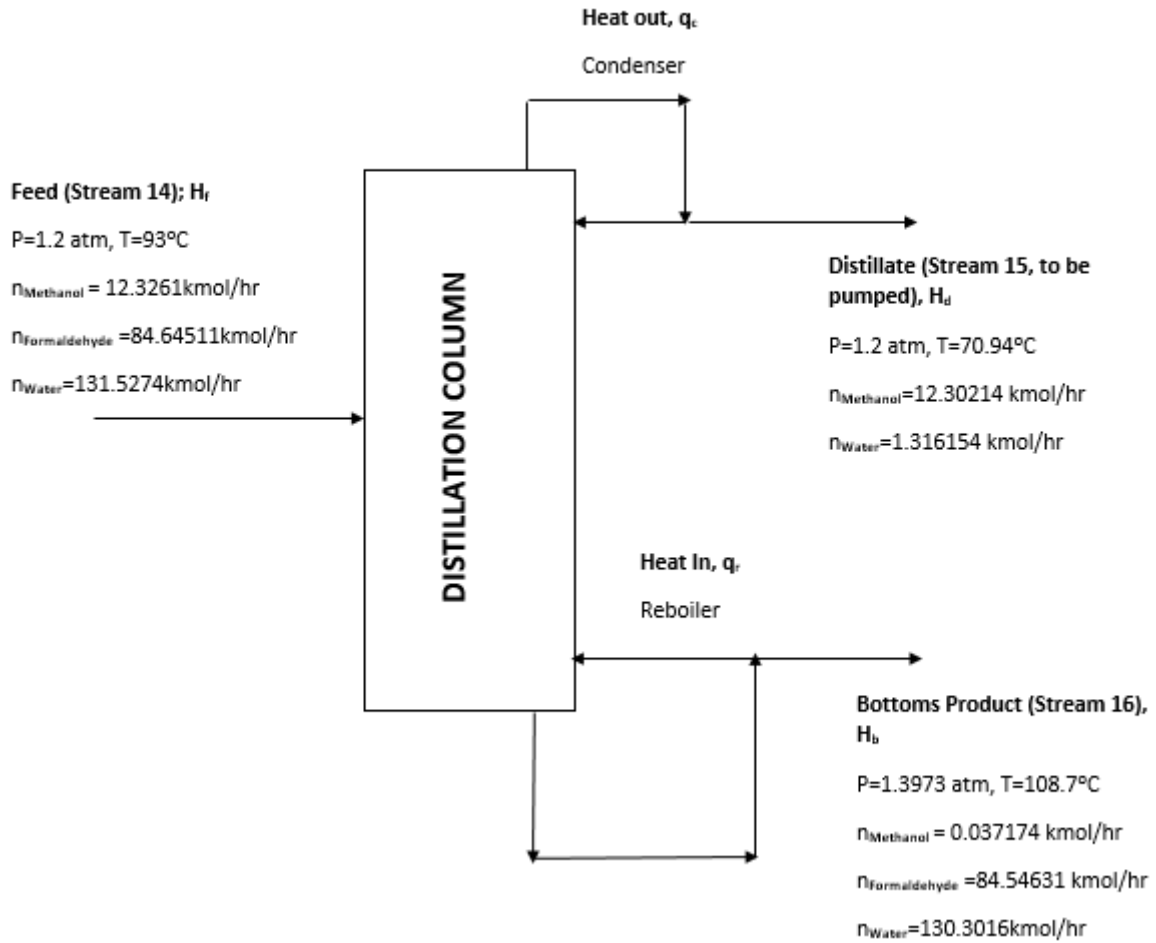
Figure 5: Mass Balance around the Distillation Column

Mole Fractions:

$$\chi_F = \frac{8.6526}{21.9} = 0.395$$

$$\chi_{H_2O} = \frac{13.244}{21.9} = 0.604$$

$$\chi_M = \frac{0.0038}{21.9} = 0.000173$$

In the bottom, to find the mass fraction of Formaldehyde (Methanol is present in a very low quantity, therefore it can be considered as a mixture of Formaldehyde in water). The composition of the Bottom Stream is given in Table 4.

Total Mass in Bottom Stream $= 259.80 + 238.392 + 0.1217 = 498.31 kg$

The mass percent of formaldehyde formed is around 52.1%.

Table 4: Composition of Bottom Stream(First Run)

Component	Moles (n)	Molecular Weight (M)	Mass (n*M)	Mass Fraction
Formaldehyde	8.6526	30	259.80	0.521
Water	13.244	18	238.392	0.478
Methanol	0.0038	32.042	0.1217	0.000244

6.1.2 Second Run

For the second run, we are considering the recycle stream N15, coming from the Distillation Column and mixing it with fresh Methanol Stream N2. The input basis for methanol is still 10 kmol/hr in Stream N3.

- **Mass balance around mixing point of streams 2, 3 and 15:**

  Now the recycle stream N15 is mixed with the pure Methanol stream N2 to keep the total input of methanol to the reactor be 10 kmol/hr. Thus, on applying mass balance on mixing tee for streams, 2, 3 and 15, we get

  $$N_{M3} = N_{M15} + N_{M2}$$

  $$N_{M2} = N_{M3} - N_{M15} = 10 - 1.25622 = 8.74378 kmol/hr$$

  $$N_{H_2O3} = 0.1338 kmol/hr$$

  Therefore Total moles now in stream $N3 = 10 + 0.1338 = 10.1338 kmol/hr$

- **Mass balance around mixing point of streams 6, 7 and 8:**

  Composition of N6=Composition of N3

  $$\chi_{M3} = \chi_{M6} = \frac{10}{10.1338} = 0.9868$$

  $$\chi_{H_2O3} = \chi_{H_2O6} = \frac{0.1338}{10.1338} = 0.0132$$

  From first run,we got

  $$N_{O_21} = 0.5 * 7.866 = 3.933 kmol/hr$$

  $$N_{N_21} = \frac{N_{O_28}}{0.21} * 0.79 = 14.796 kmol/hr$$

  $$N_7 = N_5 = N_1 = N_{O_21} + N_{N_21} = 3.933 + 14.796 = 18.729 kmol/hr$$

  Total Number of Moles in stream $N_8 = $ Moles in $N_6 + $ Moles in N_7

  $$= 10.13387 + 18.729 = 28.863 kmol/hr$$

- **Mass Balance around the Reactor**

  The feed to the reactor is stream N8 (28.863 kmol/hr) at 343 C .

  Composition:

  $$N_M = 10 kmol/hr$$
  $$N_{H_2O} = 0.1338 kmol/hr$$
  $$N_{O_2} = 3.933 kmol/hr$$
  $$N_{N_2} = 14.796 kmol/hr$$

  Since the conversion and the selectivity is the same and the amount of methanol entering is also same, thus the composition of Formaldehyde, hydrogen, nitrogen, unreacted Methanol will remain the same.

  Only the water composition will change.

  $$N_{H_2O9}(SecondRun) = 0.1338 + \xi_1 = 0.1338 + 7.866 = 7.999 kmol/hr$$

- **Mass Balance around the Absorber**

  The feed to the absorber is Stream N10 at 165 C and 138 kPa.

  $$N_{F10}(SecondRun) = N_{F10}(FirstRun) = 8.74 kmol/hr$$

  Solubility of Formaldehyde at 89.5 C= 0.78 kg Formaldehyde in 1 kg water

  In kmol Formaldehyde per kmol water units:

  $$= \frac{\frac{0.78 kg formaldehyde}{30.02}}{\frac{1 kg water}{18}} = 0.468 kmol formaldehyde/kmol water$$

  This means 0.468 kmol of formaldehyde gets absorbed by 1 kmol of water. Hence, to absorb 99% of 8.74 kmol/hr (from stream 10) of Formaldehyde we require$= \frac{8.74}{0.468} = 18.675 kmol water/hr$

  Threfore $N11$ (Minimum, Second Run)= $18.675 kmol/hr$ Now, Solubility of Methanol at 89.5 C is 0.20259 kmol Methanol/kmol of water

  For 18.675 Kmol of Water/hr, Methanol absorbed $= 0.20259 * 18.675 = 3.78 Kmol/hr$.

  Therefore, all methanol will get absorbed by N11 ($18.675 Kmol/hr$) . All Methanol will dissolve in water and NO Methanol in the off-gas. Therefore, Stream 12 will contain methanol, formaldehyde, Nitrogen and water.

  $$N_{N_212} = N_{N_210} = 14.796 kmol/hr$$
  $$N_{H_212} = N_{H_210} = 0.874 kmol/hr$$

  $N_{F12} = N_{F10} * 0.01 = 0.0874 kmol/hr$ (Amount of Formaldehyde not absorbed)

  For water, Assuming Raoults Law to be applicable:

  $$y_{H_2O,12} = \frac{P^{vapor}(89.5C)}{P^{Total}}$$

Vapor Pressure of Water at 89.5 (From Antoines Equation):

$$P^{vapor} = exp[16.3872 - \frac{3885.7}{89.5 + 230.17}] = 68.8 kPa$$

$$y_{H_2O} = \frac{68.8}{138} = 0.496$$

The amount of water going to the vapour phase (N12) will be 0.496 of the total moles of water present in the absorber. Therefore,

$$N_{H_2O,12} = (18.675 + 7.999) * 0.498 = 13.23 kmol/hr$$

Hence, $N_{H_2O,13} = (18.675 + 7.999) - 13.23 = 13.44 kmol/hr$
Also,

$$N_{M13} = 1.26 kmol/hr$$

$N_{F13} = 0.99 * 8.74 = 8.6526 kmol/hr$ (All the methanol goes to N13)
Total number of moles in Stream 12 (Second Run):

$$= N_{F12} + N_{N_212} + N_{H_212} + NH_2O, 12$$
$$= 0.0874 + 14.796 + 0.874 + 13.23 = 28.98 kmol/hr$$

Total number of moles in Stream 13 (Second Run):

$$= N_{F13} + N_{M13} + N_{H_2O,13}$$
$$= 8.6526 + 1.26 + 13.44 = 23.358 kmol/hr$$

- **Mass Balance around the Distillation Column**

  Distillate $D = N15$
  Bottom $B = N17$

  Now, $D_M = 0.997 * $ Moles of Methanol Present in $N14$

  $$D_M = 0.997 * 1.26 = 1.25622 kmol/hr$$

  hence, $B_M = 1.26 - 1.25622 = 0.0038 kmol/hr$

  Now, calculation for the bottom

  $B_{H_2O} = 0.99 * $ Moles of Water present in $N14$

  $$B_{H_2O} = 0.99 * 13.44 = 13.31 kmol/hr$$

  Hence, $D_{H_2O} = 13.44 - 13.31 = 0.13 kmol/hr$
  $B_F = $ Same as $N14 = 8.6526 kmol/hr$

Table 5: Composition of Bottom Stream(Second Run)

Component	Moles (n)	Molecular Weight (M)	Mass (n*M)	Mass Fraction
Formaldehyde	8.6526	30	259.80	0.52
Water	13.31	18	239.58	0.479
Methanol	0.0038	32.042	0.1217	0.000243

Total Distillate N15$=D_M + D_{H_2O}$

$$= 1.25622 + 0.13 = 1.38622 kmol/hr$$

Mole Fractions:

$$y_M = \frac{1.25622}{1.38622} = 0.9062$$

$$y_{H_2O} = \frac{0.1338}{1.38622} = 0.095$$

Total Bottom N17$=B_F + B_{H_2O} + B_M$

$$= 8.6526 + 13.31 + 0.0038 = 21.9664 kmol/hr$$

Mole Fractions:

$$\chi_F = \frac{8.6526}{21.9664} = 0.393$$

$$\chi_{H_2O} = \frac{13.31}{21.9664} = 0.605$$

$$\chi_M = \frac{0.0038}{21.9664} = 0.000172$$

In the bottom, to find the mass fraction of Formaldehyde (Methanol is present in a very low quantity, therefore it can be considered as a mixture of Formaldehyde in water). The composition of the Bottom Stream is given in Table 5.

Total Mass in Bottom Stream $= 259.80 + 239.58 + 0.1217 = 499.5 kg$

The mass percent of formaldehyde formed is around 52%.

- **Mass Balance around Stream N17, N18 and N19:**

$$M_{F17} = \frac{259.8 \text{kg/hr of formaldehyde}}{499.5 \text{kg solution}}$$

LetM_{18} be x kgH_2O /hr

Now, for $M_{19} = 37 wt\%$of formaldehyde

$$\frac{259.8}{499.5 + x} = 0.37$$

$$x = 202.66 \text{kg of } H_2O \text{ /hr}$$

6.2 Material Balance Data Sheets

1. **Initial Mass Balance (before scaling) on 10 kmol Methanol/hour basis:**

   On the basis of the material balance calculations done before, following data was obtained.

Table 6: Mass Balance Data sheet (before Scaling) on 10 kmol Methanol /hr basis

Stream No	Methanol	Oxygen	Formaldehyde	Water	Hydrogen	Nitrogen	Sum
1	0	3.933	0	0	0	14.796	18.729
2	8.7466	0	0	0	0	0	8.7466
3	10	0	0	0.13378	0	0	10.13378
4	10	0	0	0.13378	0	0	10.13378
5	0	3.933	0	0	0	14.796	18.729
6	10	0	0	0.13378	0	0	10.13378
7	0	3.933	0	0	0	14.796	18.729
8	10	3.933	0	0.13378	0	14.796	28.86278
9	1.26	0	8.74	7.866	0.874	14.796	33.66978
10	1.26	0	8.74	7.866	0.874	14.796	33.66978
11	0	0	0	18.675	0	0	18.675
12	0	0	0.0874	13.23	0.874	14.796	28.9874
13	1.26	0	8.6526	13.445	0	0	23.3576
14	1.26	0	8.6526	13.445	0	0	23.3576
15	1.25755	0	0	0.13454	0	0	1.39209
16	0.0038	0	8.6526	13.3197	0	0	21.966
17	0.0038	0	8.6526	13.3197	0	0	21.966
18	0	0	0	11.16667	0	0	11.16667
19	0.0038	0	8.6526	24.48637	0	0	33.13267
20	0.0038	0	8.6526	24.48637	0	0	33.13267

2. **Mass Balance(before scaling) on kg/hr unit:**

Each of the component in each of the stream is divided by its molecular weight to get its composition on kg/hr basis. Molecular weights of the components are as follows:

Methanol = 32 kg/kmol

Oxygen = 32 kg/kmol

Formaldehyde=30 kg/kmol

Water=18 kg/kmol

Hydrogen=2 kg/kmol

Nitrogen=28 kg/kmol

Table 7: Mass Balance Data sheet (before Scaling) on kg /hr basis

Stream No	Methanol	Oxygen	Formaldehyde	Water	Hydrogen	Nitrogen	Sum
1	0	125.856	0	0	0	414.288	540.144
2	279.8912	0	0	0	0	0	279.8912
3	320	0	0	2.40804	0	0	322.408
4	320	0	0	2.40804	0	0	322.408
5	0	125.856	0	0	0	414.288	540.144
6	320	0	0	2.40804	0	0	322.408
7	0	125.856	0	0	0	414.288	540.144
8	320	125.856	0	2.40804	0	414.288	862.552
9	40.32	0	262.2	141.588	1.748	414.288	860.144
10	40.32	0	262.2	141.588	1.748	414.288	860.144
11	0	0	0	336.15	0	0	336.15
12	0	0	2.622	238.14	1.748	414.288	656.798
13	40.32	0	259.578	242.01	0	0	541.908
14	40.32	0	259.578	242.01	0	0	541.908
15	40.2416	0	0	2.42172	0	0	42.66332
16	0.1216	0	259.578	239.7546	0	0	499.1512
17	0.1216	0	259.578	239.7546	0	0	499.1512
18	0	0	0	201.0001	0	0	201.0001
19	0.1216	0	259.578	440.7547	0	0	700.1513
20	0.1216	0	259.578	440.7547	0	0	700.1513

3. **Mass Balance (before scaling) on ton/year mass unit:**

The calculations made in the above table was in kg/hr basis it is converted to tonnes per year by multiplying by 8.76 as

$$1kg/hr = 365 * \frac{24}{1000} tons/year$$
$$= 8.76 tons/year$$

Table 8: Mass Balance Data Sheet (before scaling) on ton/year mass unit

Stream No	Methanol	Oxygen	Formaldehyde	Water	Hydrogen	Nitrogen	Sum
1	0	1102.499	0	0	0	3629.163	4731.661
2	2451.847	0	0	0	0	0	2451.847
3	2803.2	0	0	21.09443	0	0	2824.294
4	2803.2	0	0	21.09443	0	0	2824.294
5	0	1102.499	0	0	0	3629.163	4731.661
6	2803.2	0	0	21.09443	0	0	2824.294
7	0	1102.499	0	0	0	3629.163	4731.661
8	2803.2	1102.499	0	21.09443	0	3629.163	7555.956
9	353.2032	0	2296.872	1240.311	15.31248	3629.163	7534.861
10	353.2032	0	2296.872	1240.311	15.31248	3629.163	7534.861
11	0	0	0	2944.674	0	0	2944.674
12	0	0	22.96872	2086.106	15.31248	3629.163	5753.55
13	353.2032	0	2273.903	2120.008	0	0	4747.114
14	353.2032	0	2273.903	2120.008	0	0	4747.114
15	352.5164	0	0	21.21427	0	0	373.7307
16	1.065216	0	2273.903	2100.25	0	0	4372.565
17	1.065216	0	2273.903	2100.25	0	0	4372.565
18	0	0	0	1760.761	0	0	1760.761
19	1.065216	0	2273.903	3861.011	0	0	6133.325
20	1.065216	0	2273.903	3861.011	0	0	6133.325

4. **Mass Balance (after scaling) on ton/year mass unit:**

Since the production capacity was fixed to be 60000 tons per year, the mass flow rate of stream 20 should be 60000 tonnes/year. Therefore, each of the components mass flow rate has been scaled up to meet the above requirement.

Table 9: Mass Balance Data Sheet (after scaling) on ton/year mass unit

Stream No	Methanol	Oxygen	Formaldehyde	Water	Hydrogen	Nitrogen	Sum
1	0	10785.33	0	0	0	35502.73	46288.06
2	23985.49	0	0	0	0	0	23985.49
3	27422.65	0	0	206.3588	0	0	27629
4	27422.65	0	0	206.3588	0	0	27629
5	0	10785.33	0	0	0	35502.73	46288.06
6	27422.65	0	0	206.3588	0	0	27629
7	0	10785.33	0	0	0	35502.73	46288.06
8	27422.65	10785.33	0	206.3588	0	35502.73	73917.06
9	3455.253	0	22469.43	12133.49	149.7962	35502.73	73710.7
10	3455.253	0	22469.43	12133.49	149.7962	35502.73	73710.7
11	0	0	0	28806.63	0	0	28806.63
12	0	0	224.6943	20407.59	149.7962	35502.73	56284.81
13	3455.253	0	22244.74	20739.23	0	0	46439.22
14	3455.253	0	22244.74	20739.23	0	0	46439.22
15	3448.535	0	0	207.5312	0	0	3656.066
16	10.42061	0	22244.74	20545.95	0	0	42775.15
17	10.42061	0	22244.74	20545.95	0	0	42775.15
18	0	0	0	17224.85	0	0	17224.85
19	10.42061	0	22244.74	37770.81	0	0	60000
20	10.42061	0	22244.74	37770.81	0	0	60000

5. **Mass Balance (after scaling) on kg/hr mass unit:**

   After we scaled up each of the components mass flow rate, we got all the data in tons per year.
   For the energy balance calculations, each of the mass flow rates are converted to kg/hr.

Table 10: Mass Balance Data Sheet (after scaling) on kg/hr mass unit

Stream No	Methanol	Oxygen	Formaldehyde	Water	Hydrogen	Nitrogen	Sum
1	0	1231.202	0	0	0	4052.823	5284.025
2	2738.07	0	0	0	0	0	2738.07
3	3130.439	0	0	23.55695	0	0	3153.996
4	3130.439	0	0	23.55695	0	0	3153.996
5	0	1231.202	0	0	0	4052.823	5284.025
6	3130.439	0	0	23.55695	0	0	3153.996
7	0	1231.202	0	0	0	4052.823	5284.025
8	3130.439	1231.202	0	23.55695	0	4052.823	8438.021
9	394.4353	0	2565.003	1385.102	17.10002	4052.823	8414.464
10	394.4353	0	2565.003	1385.102	17.10002	4052.823	8414.464
11	0	0	0	3288.428	0	0	3288.428
12	0	0	25.65003	2329.634	17.10002	4052.823	6425.207
13	394.4353	0	2539.353	2367.492	0	0	5301.281
14	394.4353	0	2539.353	2367.492	0	0	5301.281
15	393.6684	0	0	23.69077	0	0	417.3591
16	1.189567	0	2539.353	2345.429	0	0	4883.007
17	1.189567	0	2539.353	2345.429	0	0	4883.007
18	0	0	0	1966.308	0	0	1966.308
19	1.189567	0	2539.353	4311.736	0	0	6849.315
20	1.189567	0	2539.353	4311.736	0	0	6849.315

6. **Mass Balance (after scaling) on kmol/hr mass unit:**

We have also converted all the mass flow rates to molar flow rates by dividing them by their corresponding molecular weights.

Table 11: Mass Balance Data Sheet (after scaling) on kmol/hr mass unit

Stream No	Methanol	Oxygen	Formaldehyde	Water	Hydrogen	Nitrogen	Sum
1	0	38.47505	0	0	0	144.7437	183.2187
2	85.56468	0	0	0	0	0	85.56468
3	97.82622	0	0	1.308719	0	0	99.13494
4	97.82622	0	0	1.308719	0	0	99.13494
5	0	38.47505	0	0	0	144.7437	183.2187
6	97.82622	0	0	1.308719	0	0	99.13494
7	0	38.47505	0	0	0	144.7437	183.2187
8	97.82622	38.47505	0	1.308719	0	144.7437	282.3537
9	12.3261	0	85.50012	76.9501	8.550012	144.7437	328.07
10	12.3261	0	85.50012	76.9501	8.550012	144.7437	328.07
11	0	0	0	182.6905	0	0	182.6905
12	0	0	0.855001	129.4241	8.550012	144.7437	283.5728
13	12.3261	0	84.64511	131.5274	0	0	228.4986
14	12.3261	0	84.64511	131.5274	0	0	228.4986
15	12.30214	0	0	1.316154	0	0	13.61829
16	0.037174	0	84.64511	130.3016	0	0	214.8851
17	0.037174	0	84.64511	130.3016	0	0	214.8851
18	0	0	0	109.2393	0	0	109.2393
19	0.037174	0	84.64511	239.5409	0	0	324.1244
20	0.037174	0	84.64511	239.5409	0	0	324.1244

6.3 Energy Balance

Energy balance refers to the calculation of heat duties and energy consumed for each plant component. It depends on calculating the heat capacity (C_P) of each substance present in the system. Table 12 serves as reference to the upcoming calculations of the plants energy balance.[11]

Heat Capacity C_P (J/kmol.K) for liquid phase (for all compounds except Hydrogen) is calculated as per the following equation:

$$C_P = C_1 + C_2T + C_3T^2 + C_4T^3 + C_5T^4$$

For Hydrogen in liquid state, following equation is used:

$$C_P = \frac{C_1{}^2}{T} - C_2 - 2C_1C_3T - C_1C_4T^2 - C_3{}^{\frac{2}{3}}T^3 - \frac{C_3C_4}{2}T^4 - C_4{}^{\frac{2}{5}}T^5$$

$T = 1 - t_r$ where t_r is the reduced temperature.
For ideal gas specific Heat calculation (for all the compounds except gaseous Hydrogen):

$$C_P = C_1 + C_2\left[\frac{\frac{C_3}{T}}{\sinh\frac{C_3}{T}}\right]^2 + C_4\left[\frac{\frac{C_5}{T}}{\cosh\frac{C_5}{T}}\right]^2$$

where T is in Kelvin.
For Hydrogen gas:

$$C_P = C_1 + C_2\ln T + \frac{C_3}{T} + C_4T$$

Referring to Table 12 , all the heat capacities are calculated. The reference temperature is taken as 298.15K (25 C) for all the calculations. Energy Balance is done for each individual component (Mixing Tees, Pumps, Heat Exchangers and Compressor along with Reactor, Absorber and Distillation Column).

Table 12: Coefficients for the calculation of Heat Capacity

Substance	Molecular Weight	Phase	C1	C2	C3	C4	C5	Tmin (K)	CP at Tmin (*E-05)	Tmax(K)	CP at Tmax (*E-05)
Methanol	32.042	Gas	0.3925E+05	0.8790 E+05	1.9165E+03	0.5365 E+05	896.7	200	0.3980	1500	1.0533
			-3.6223E+02	0.9379	0	0	175.47	0.7112	400	1.1097	400	1.1097
Formaldehyde	30.026	Gas	0.3327 E+05	0.4954 E+05	1.8666 E+03	0.2808 E+05	934.9	50	0.3327	1500	0.7113
			2.83E+01	0	0	0	204	0.6767	234	0.6852	234	0.6852
Oxygen	31.999	Gas	0.2910 E+05	0.1004 E+05	2.5265 E+03	0.0936 E+05	1153.8	50	0.2910	1500	0.3653
			-6.1523E+03	1.1392E+02	-9.2382E-01	2.7963E-03	54.36	0.5365	142	0.9066	142	0.9066
Nitrogen	28.014	Gas	0.2911 E+05	0.0861 E+05	1.7016 E+03	0.0010 E+05	909.79	50	0.2911	1500	0.3484
			-1.2281E+04	2.4800E+02	-2.218	7.4902E-03	63.15	0.5593	112	0.7960	112	0.7960
Water	18.015	Gas	0.3336 E+05	0.2679 E+05	2.6105 E+03	0.0890 E+05	1169	100	0.3336	2273.15	0.5276
			-2.0901E+03	8.1250	-1.4116E-02	9.3701E-06	273.16	0.7615	533.15	0.8939	533.15	0.8939
Hydrogen	2.016	Gas	0.2762 E+05	0.0956 E+05	2.4660 E+03	0.0376	567.6	250	0.2843	1500	0.3225
			6.7659E+03	-1.2363E+02	4.7827E+02	0	13.95	0.1262	32	1.3122	32	1.3122

- **Mixing Point between Stream 2, 15 and 3:**

  Energy coming in should be equal to the energy going out.

  Stream 15 (Recycle)

  T=70.94 °C, P=1.2 atm

  $n_{Methanol}$ = 12.30214 kmol

  n_{Water} = 1.316154 kmol

  Stream 3

  T=? , P=1 atm

  $n_{Methanol}$ = 97.82622 kmol

  n_{Water} = 1.316154 kmol

  Stream 2 (Pure Methanol)

  T=25°C, P=1 atm

  $n_{Methanol}$ = 85.56468 kmol

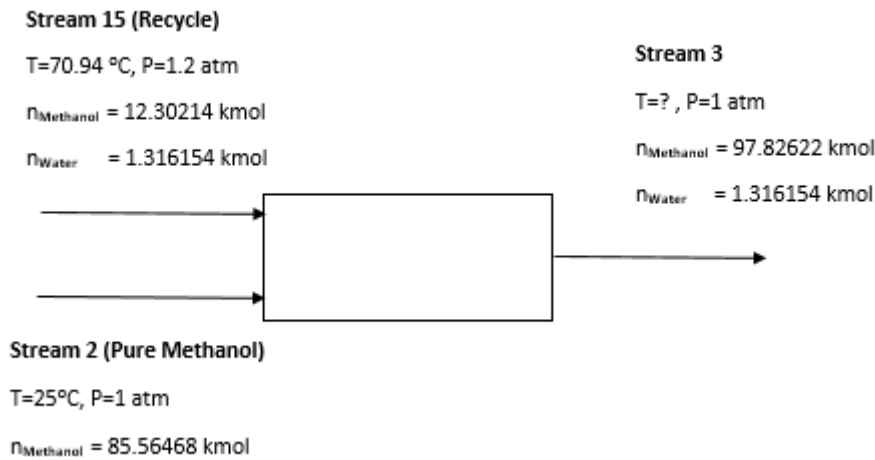 Figure 6: Mixing Point between Stream 2, 15 and 3

$$n_{2,methanol} \int_{298.15}^{298.15} C_{p,methanol} dT + n_{15,water} \int_{298.15}^{343.94} C_{p,water} dT + n_{15,methanol} \int_{298.15}^{343.94} C_{p,methanol} dT$$

$$= n_{3,water} \int_{298.15}^{T} C_{p,water} dT + n_{3,methanol} \int_{298.15}^{T} C_{p,methanol} dT$$

Refer to Table 12 to get the respective values for $C_1, C_2, C_3, C_4 and C_5$. We will use the liquid phase values as the compounds at the given pressure and temperature conditions are in liquid state.

Using MATLAB to solve the above equation [5] , we get: T= 302.8276K (29.6776 C) The temperature of stream 3 is 29.6776 C.

- **Pump 1**

  Mass Flow Rate (Stream 3):

  Methanol $= 3130.439 kg/hr$

  Water$= 23.55695 kg/hr$

  Total $= 3153.996 kg/hr$

  In mole fraction (Stream 3):

$$\chi_{methanol} = \frac{97.82622}{99.13494} = 0.9867$$

$$\chi_{water} = \frac{1.308719}{99.13494} = 0.0133$$

Using Dortmund Data Bank (DIPPR105 EQN) [1]:

At $302.8276K$,

$$\rho_{methanol} = 784.34 kg/m^3$$

$$\rho_{water} = 996.505 kg/m^3$$

$$\rho_{mix} = \sum \rho_i \chi_i = 786.95 kg/m^3$$

Now, $P(in), P1 = 1atm = 101325Pa$

$$P(out), P2 = 3atm = 303975Pa$$

In a pump, generally suction velocity(v_1)=Discharge velocity (v_2).

Applying Bernoullis Equation:

$$\frac{P_1}{\rho g} + \alpha \frac{v_1^2}{2g} + z_1 + W_{pump} = \frac{P_2}{\rho g} + \alpha \frac{v_2^2}{2g} + z_2 + \text{losses}$$

Also $z_1 = z_2$

Assuming no losses, we get W_{pump} as:

$$W_{pump} = 26.3138 mHead$$

The efficiency of the pump is assumed to be 75%. Therefore Required Work $= W_{pump}/0.75$

$$= 26.3138/0.75 = 35.0851m$$

$$Power(in kJ/hr) = Required Work * Total mass flowrate * g$$
$$= 35.0851 * 3153.996 * 9.81 = 1085.55 kJ/hr$$

- **Heat Exchanger E1:**

  In the Heat Exchanger, phase change is also taking place. Therefore, in addition to the sensible heat change, there is a latent heat of vaporization also which is to be accounted.

  Heat of Vaporization of Methanol [4], $H_v, Methanol = 35278000 J/kmol$

  Heat of Vaporization of Water [4] , $H_v, Water = 40.7E + 06 J/kmol$

$$Q_{in} = \sum nC_P \triangle T$$
$$= (nC_P \triangle T)_{methanol} + (nC_P \triangle T)_{Water}$$

$$Q_{in} = 1.308719 \left[\int_{302.82}^{373.15} (C_P, l)_{water} dT + H_{V,Water} + \int_{373.15}^{423.15} (C_P, g)_{water} dT \right]$$
$$+ 97.82622 \left[\int_{302.82}^{337.7} (C_P, l)_{methanol} dT + H_{V,methanol} + \int_{337.7}^{423.15} (C_P, g)_{methanol} dT \right]$$
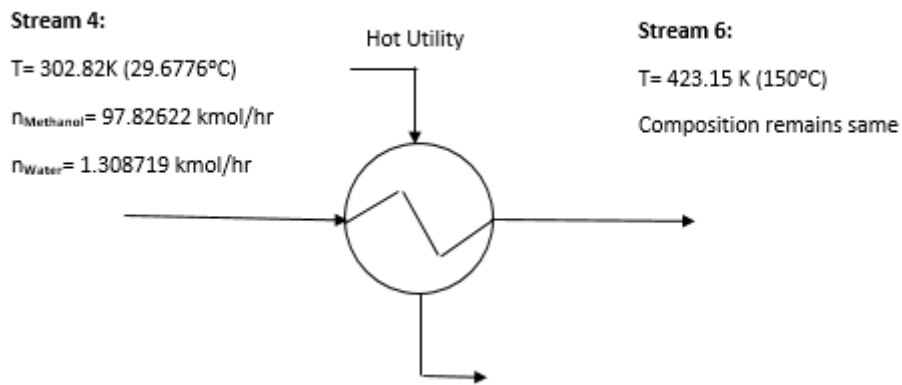
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Figure 7: Heat Exchanger E1

Using MATLAB Heat Input is computed [5],

$$Q_{in} = 4.2014 * 10^6 kJ/hr$$

Therefore, Hot Utility Consumption $= 4.2014 * 10^6 kJ/hr$

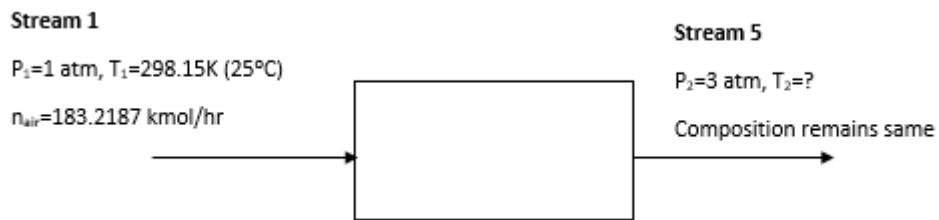Figure 8: Compressor C1

- **Heat Exchanger E2:**

  Following the same method as Heat Exchanger E1,

  $$Q_{in} = \sum nC_P \triangle T$$

  $$= (nC_P \triangle T)_{Nitrogen} + (nC_P \triangle T)_{Oxygen}$$

  $$Q_{in} = 144.7437 \left[\int^{423.15} (C_P)_{Nitrogen} dT \right] + 38.47505 \left[\int^{423.15} (C_P)_{Oxygen} dT \right]$$

Stream 5

P=3 atm, T=407.8604K (134.7°C)

$n_{Nitrogen}$ = 144.7437 kmol/hr

n_{Oxygen} =38.47505kmol/hr

n_{Total}=183.2187 kmol/hr

Hot Utility

Stream 7
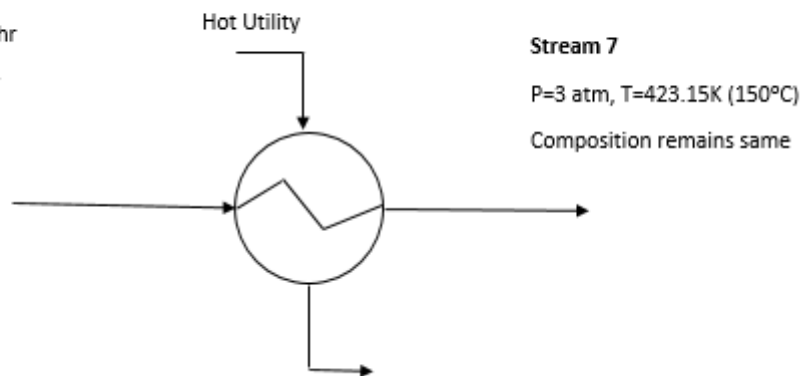
P=3 atm, T=423.15K (150°C)

Composition remains same

Figure 9: Heat Exchanger E2

Using MATLAB Heat Input is computed [5] ,

$$Q_{in} = 82658kJ/hr$$

Therefore, Hot utility consumption=$Q_{in} = 82658kJ/hr$.

- **Mixing point between Streams 6, 7 and 8:**

  The temperature of stream number 6 is same as the temperature of stream number 7, so stream 8 will also has the same temperature (150 C).

  Stream 6

  P=3atm, T=423.15K (150ºC)

  $n_{Methanol}$=97.82622kmol/hr

  n_{Water}=1.308719kmol/hr

  Stream 8

  $n_{Methanol}$=97.82622kmol/hr

  n_{Water}=1.308719kmol/hr

  n_{Oxygen}=38.47505kmol/hr

  $n_{Nitrogen}$=144.7437kmol/hr

  Stream 7

  P=3atm, T=423.15K (150ºC)

  n_{Oxygen}=38.47505kmol/hr

  $n_{Nitrogen}$=144.7437kmol/hr

Figure 10: Mixing point between Streams 6, 7 and 8

- **Reactor**

  Stream 8 enters the reactor and stream 9 comes out. Oxygen is the limiting reactant.

Table 13: Energy Balance around the Reactor

Species	Stream 8 (n_in) kmol/hr	Hin	Stream 9 (n_in) kmol/hr	Hout
Methanol	97.82622	H1	kmol/hr	H5
Oxygen	38.47505	H2	0	H6
Nitrogen	144.7437	H3	144.7437	H7
Formaldehyde	-	-	85.50012	H8
Hydrogen	-	-	8.550012	H9
Water	1.308719	H4	76.9501	H10

Now,

$$H_1 = \int_{298.15}^{423.15} (C_P)_{Methanol,gas} dT = 4.2954 * 10^7 J/hr$$

$$H_2 = \int_{298.15}^{423.15} (C_P)_{Oxygen,gas} dT = 3.7205 * 10^5 J/hr$$

$$H_3 = \int_{298.15}^{423.15} (C_P)_{Nitrogen,gas} dT = 3.6499 * 10^5 J/hr$$

$$H_4 = \int_{298.15}^{423.15} (C_P)_{Water,gas} dT = 4.8071 * 10^7 J/hr$$

$$H_5 = \int_{298.15}^{616.15} (C_P)_{Methanol,gas} dT = 5.4759 * 10^7 J/hr$$

$$H_6 = \int_{298.15}^{616.15} (C_P)_{Oxygen,gas} dT = 9.7570 * 10^6 J/hr$$

$$H_7 = \int_{298.15}^{616.15} (C_P)_{Nitrogen,gas} dT = 9.3827 * 10^6 J/hr$$

$$H_8 = \int_{298.15}^{616.15} (C_P)_{Formaldehyde,gas} dT = 1.323 * 10^7 J/hr$$

$$H_9 = \int_{298.15}^{616.15} (C_P)_{Hydrogenl,gas} dT = 9.29 * 10^6 J/hr$$

$$H_{10} = \int_{298.15}^{616.15} (C_P)_{Water,gas} dT = 5.4913 * 10^7 J/hr$$

The first (main) reaction is exothermic with a yield of $7.866 kmol/hr$ and second reaction is endothermic with a yield of $0.874 kmol/hr$.

Total Formaldehyde formed: $85.50012 kmol/hr$

From first reaction which is exothermic, formaldehyde formed is $= \frac{7.866}{8.74} * 85.50012 = 76.9 kmol/hr$

Therefore from the second reaction which is endothermic, formaldehyde produced $= 85.50012 - 76.9 = 8.5852 kmol/hr$

Heat of formation of the first reaction$= -156000 kJ/kmol$

Heat of formation of the second reaction$= 85000 kJ/kmol$

Net Heat released due to both the reactions$= (-156000 * 76.9) + (85000 * 8.5852) = -11266658 kJ/hr$

Change in enthalpy around the reactor $=$ Heat of Formation $+ H(out) - H(in)$

$$= -11266658 kJ/hr + H(out) - H(in)$$

- **Heat Exchanger after the Reactor E3:**

  This heat exchanger installed after the reactor brings the product temperature down to 200 C from 343 C.

  These are the enthalpies at the end of the reactor before entering the Heat Exchanger (Cold Utility will be used to bring the temperature down):

$$H_5 = \int_{298.15}^{616.15} (C_P)_{Methanol,gas} dT = 5.4759 * 10^7 J/hr$$

$$H_6 = \int_{298.15}^{616.15} (C_P)_{Oxygen,gas} dT = 9.7570 * 10^6 J/hr$$

$$H_7 = \int_{298.15}^{616.15} (C_P)_{Nitrogen,gas} dT = 9.3827 * 10^6 J/hr$$

$$H_8 = \int_{298.15}^{616.15} (C_P)_{Formaldehyde,gas} dT = 1.323 * 10^7 J/hr$$

$$H_9 = \int_{298.15}^{616.15} (C_P)_{Hydrogen,gas} dT = 9.29 * 10^6 J/hr$$

$$H_{10} = \int_{298.15}^{616.15} (C_P)_{Water,gas} dT = 5.4913 * 10^7 J/hr$$

Now, these are the enthalpies at the end of the reactor and the heat exchanger, Stream 9 (bringing the temperature down to 200 C):

$$H_{9,Methanol} = \int_{298.15}^{473.15} (C_P)_{Methanol,gas} dT = 4.5736 * 10^7 J/hr$$

$$H_{9,Oxygen} = \int_{298.15}^{473.15} (C_P)_{Oxygen,gas} dT = 5.247 * 10^6 J/hr$$

$$H_{9,Nitrogen} = \int_{298.15}^{473.15} (C_P)_{Nitrogen,gas} dT = 5.1194 * 10^6 J/hr$$

$$H_{9,Formaldehyde} = \int_{298.15}^{473.15} (C_P)_{Formaldehyde,gas} dT = 6.7819 * 10^6 J/hr$$

$$H_{9,Hydrogen} = \int_{298.15}^{473.15} (C_P)_{Hydrogen,gas} dT = 5.0974 * 10^6 J/hr$$

$$H_{9,Water} = \int_{298.15}^{473.15} (C_P)_{Water,gas} dT = 4.9805 * 10^7 J/hr$$

Net Cold Utility Consumption= $\sum n_{out} H_{out} - \sum n_{in} H_{in}$

- **Throttle**

  Throttle is used to reduce the temperature; its calculation depends on the difference in pressure ($\triangle P$) of the inlet and outlet of the reactor. This leads to the need for the reactors dimensions.Since our project doesn't deal with the reactor's design. Therefore, the temperature after the throttle was decided to be chosen 165 C (from literature reference) in order to continue the energy balance around the absorber.

Stream 11 (Water)

P=1 atm, T=20 °C

n_{Water}=182.6905 kmol/hr

Stream 12

P=1 atm, T=?

$n_{Formaldehyde}$=0.855001kmol/hr

n_{Water}=129.4241kmol/hr

$n_{Hydrogen}$=8.550012kmol/hr

$n_{Nitrogen}$=144.7437kmol/hr

ABSORBER

Stream 10

T= 165°C

$n_{Methanol}$= 12.3261kmol/hr

$n_{Formaldehyde}$=85.50012 kmol/hr

n_{Water}=76.9501kmol/hr

$n_{Hydrogen}$=8.550012kmol/hr

$n_{Nitrogen}$=144.7437kmol/hr

Stream 13

P= 1.2 atm, T=?

$n_{Methanol}$=12.3261kmol/hr

n_{Water}=131.5274kmol/hr

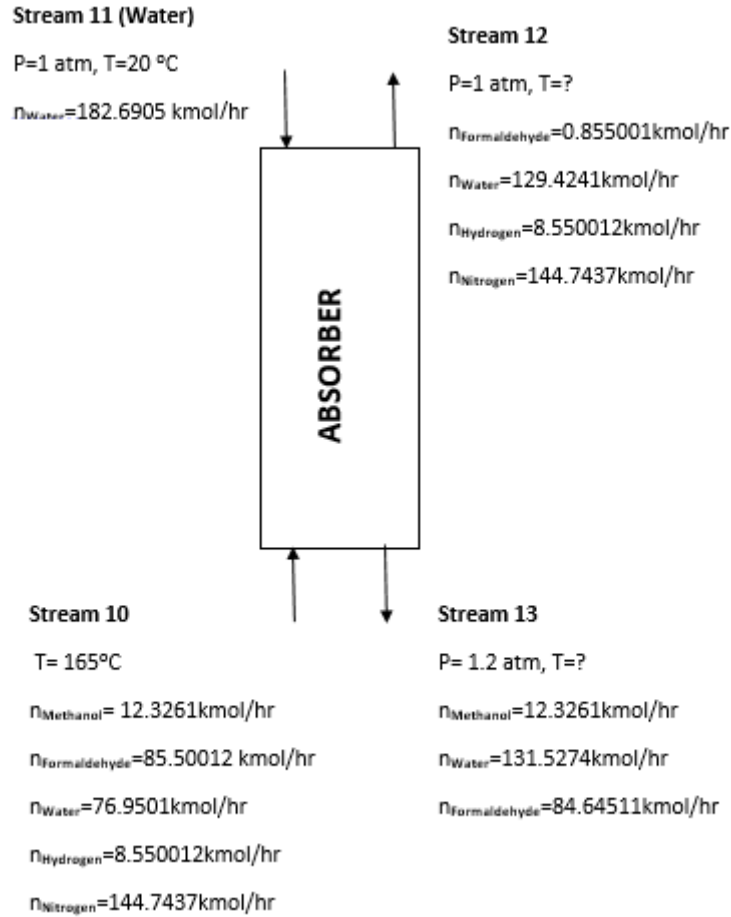$n_{Formaldehyde}$=84.64511kmol/hr

Figure 11: Energy Balance around the Absorber

- **Absorber**

  We have four streams, the temperature of the two inlets streams are 20 C and 165 C for reaction product and water stream respectively.

  Now, for energy flow computation, we know that

  Input + Generation=Output + Consumption+ Accumulation (All are in terms of Rate) [12]

  There is no generation, consumption or accumulation of heat inside the absorber.

  Therefore, Input=Output

  $$\sum n_{out} H_{out} = \sum n_{in} H_{in}$$

  Therefore,

  $$H_{in} = H(Stream 11) + H(Stream 10)$$

  $$H_{in} = \int_{298.15}^{293.15} \sum n_{i,11} C_{Pi,11} dT + \int_{298.15}^{438.15} \sum n_{i,10} C_{Pi,10} dT$$

  Also,

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$$H_{out} = H(Stream12) + H(Stream13)$$

$$H_{out} = \int_{298.15}^{T} \sum n_{i,12} C_{Pi,12} dT + \int_{298.15}^{T} \sum n_{i,13} C_{Pi,13} dT$$

Equating $H(in) = H(out)$ and solving for T, we get

$$T = 362.31K$$

- **Heat Exchanger E3:**

Stream 13

P=1.2 atm, T=362.31K

$n_{Methanol}$ = 12.3261kmol/hr

$n_{Formaldehyde}$ =84.64511kmol/hr

n_{Water}=131.5274kmol/hr

Hot Utility

Stream 14

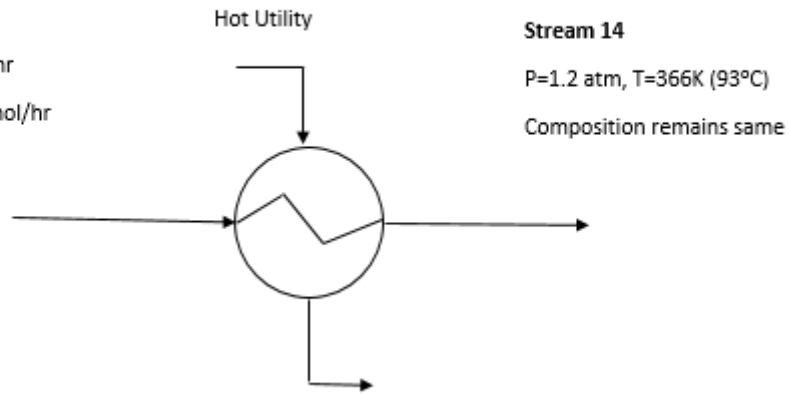P=1.2 atm, T=366K (93°C)

Composition remains same

Figure 12: Energy Balance around the Heat Exchanger E3

$$Q_{in} = \sum nC_P \triangle T$$
$$= (nC_P \triangle T)_{Methanol} + (nC_P \triangle T)_{Water} + (nC_P \triangle T)_{Formaldehyde}$$

- **Distillation Column**

  For given feed and product streams, only one of the heat effects, ϕ_b or ϕ_c is independent and subject to choice by the designer. In our project, ϕ_c is chosen to correspond to the desired reflux ratio and moles of overhead vapor. Then ϕ_b is calculated from the following overall Energy Balance equation:

  $$FH_f + \phi_b = DH_d + BH_b + \phi_c + \phi_{losses}$$

  The reflux ratio is taken to be around 8.671.
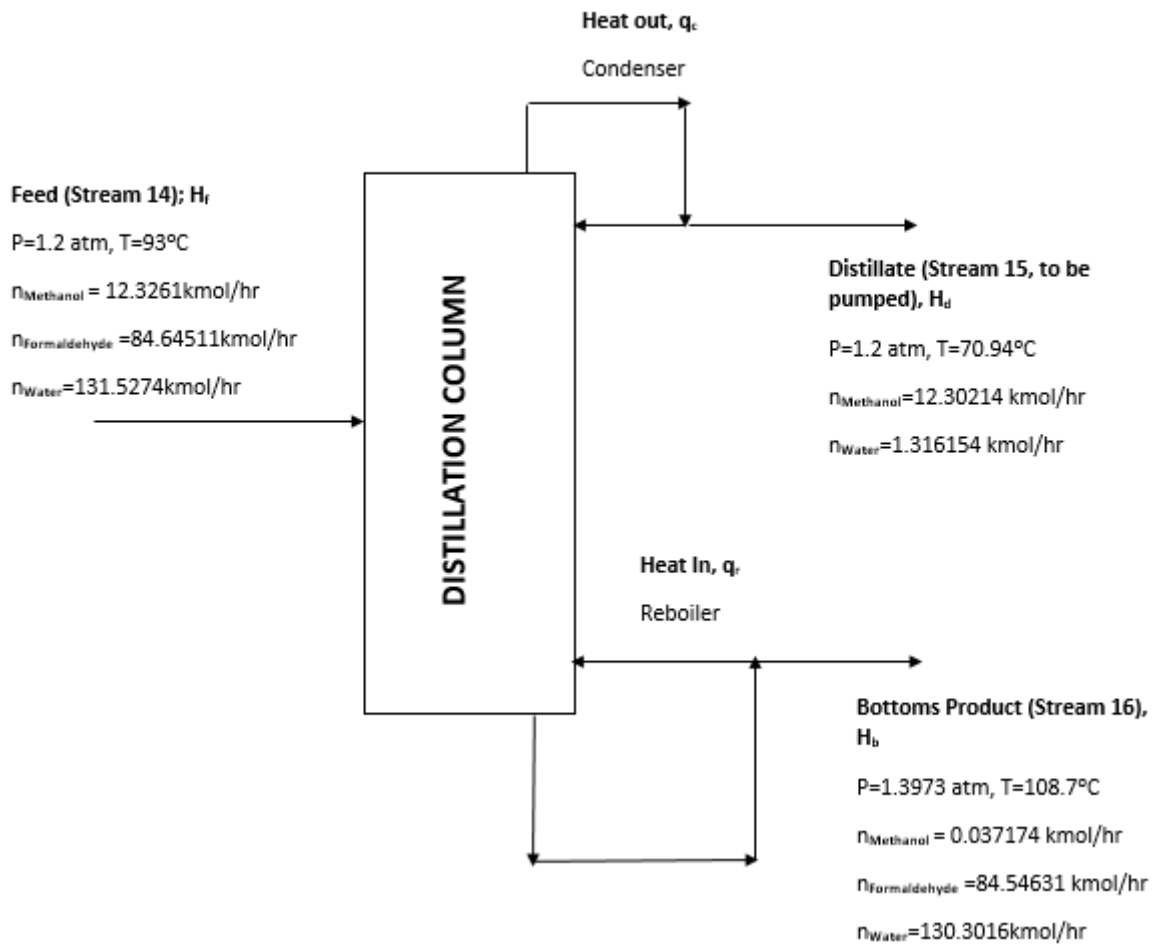

Figure 13: Energy Balance around the Distillation Column

Setting the reference Temperature as 344.09 K for enthalpy calculations.

$$T_o = 344.09K$$

Heat duty of reboiler is found out by calculating the condenser heat load first and then substituting it in the overall heat balance equation.

Condenser Heat Load is given by : $\phi_c = (R+1)D\lambda$

where λ =Latent Heat of vaporization of the mixture at the dew point [4]

To find the λ_{avg}, we will use Watson's Equation:

$$\frac{\lambda_2}{\lambda_1} = [\frac{T_c - T}{T_c - T_1}]^{0.38}$$

where $\lambda_2 = Heat of vaporization at T$

$$\lambda_1 = Heat of Vaporization at T1$$

$$T_c = Critical Temperature$$

Properties of Components of Distillate:

Table 14: Properties of Components of Distillate

Component	y_i	$T_c(K)$	$T_1(K)$	$\lambda at T_1$, kJ/kmol
Methanol	0.9033	512.6	338.15	35278
Water	0.0967	647.15	373.15	40650

Using Watson's Equation:

$$\lambda_{methanol} = 35278 \times [\frac{512.6 - 344.09}{512.6 - 338.15}]^0.38 = 34713.529kJ/kmol at 344.09K$$

$$\lambda_{water} = 40650 \times [\frac{647.15 - 344.09}{512.6 - 338.15}]^0.38 = 42167.834kJ/kmol at 344.09K$$

$$\lambda_{avg} = \sum y_i\lambda_i = 35434.360kJ/kmol$$

Now, Heat Duty of Condensation: $\phi_c = (8.671 + 1)13.615 \times 35434.360 = 4666794.552kJ/hr = 1296.331kW$

Now, $T_o = 344.09K$

Therefore, $DH_d = 0$

Enthalpy of Feed: FH_f

$$FH_f = n_{methanol,14} \int_{344.09}^{366.15} C_{P,methanol}dT + n_{water,14} \int_{344.09}^{366.15} C_{P,water}dT + n_{formaldehyde,14} \int_{344.09}^{366.15} C_{P,formaldehyde}$$

Refering to Table 12, values of cheat capacity coefficients is taken and FH_f is calculated.

$$FH_f = 105.214 kW$$

Enthalpy of Bottom: BH_b

$$BH_b = n_{water} \int_{344.09}^{381.85} C_{P,water} dT + n_{formaldehyde} \int_{344.09}^{381.85} C_{P,formaldehyde} dT + n_{methanol} \int_{344.09}^{381.85} C_{P,methanol} dT$$
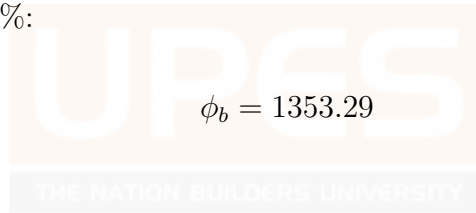
On calculating, $BH_b = 97.7363 kW$

Therefore, Substituting in the overall heat balance equation:

$$105.214 + \phi_b = 97.7363 + 1296.331 + \phi_{losses}$$

$$\phi_b = 1288.85 + \phi_{losses}$$

$$\phi_b = 1.05 \times 1288.85 = 1353.295 kW$$

Considering losses to be 5%:

$$\phi_b = 1353.29$$

7   Design of Distillation Column

Distillation is a process that separates two or more components into an overhead distillate and bottoms. The bottoms product is almost exclusively liquid, while the distillate may be liquid or a vapor or both. The separation process requires three things. [3]. First, a second phase must be formed so that both liquid and vapor phases are present and can contact each other on each stage within a separation column. Secondly, the components have different volatilities so that they will partition between the two phases to different extent. Lastly, the two phases can be separated by gravity or other mechanical means. Distillation differs from absorption and stripping in that the second phase is created by thermal means

The distillation column contains one feed stream and two product streams. The feed contains a mole percent of the light component, Z_F. The product stream exiting the top has a composition of X_D of the light component [8]. The product stream leaving the bottom contains a composition of X_b of the light component. The column is broken in two sections. The top section is referred to as the rectifying section. The bottom section is known as the stripping section. The top product stream passes through a total condenser. This effectively condenses all of the vapor distillate to liquid.

The calculations include a mass balance of the distillation unit, average physical properties of the components and relative volatilities. The minimum reflux ratio of the column is obtained through Underwood's equations and the minimum number of trays is obtained through Fenske's Equation. The diameter of the column is sized in the rectifying section and the stripping section. The layout of the sieve trays and their hydrodynamic effects are then obtained in a detailed fashion for the top and bottom sections.

Table 15: Composition of Feed, Distillate and Bottom Streams

Component	Feed		Distillate		Bottom	
	Moles(kmol/hr)	Mole Fraction	Moles(kmol/hr)	Mole Fraction	Moles(kmol/hr)	Mole Fraction
Methanol	12.3261	0.0539	12.30214	0.9031	0.037174	1.72*10E-04
Water	131.5274	0.5756	1.316154	0.0969	130.3016	0.6063
Formaldehyde	84.64511	0.3704	−		84.54631	0.3934

In our system :

Heavy Key: Water

Light Key: Methanol

Non-key component: Formaldehyde
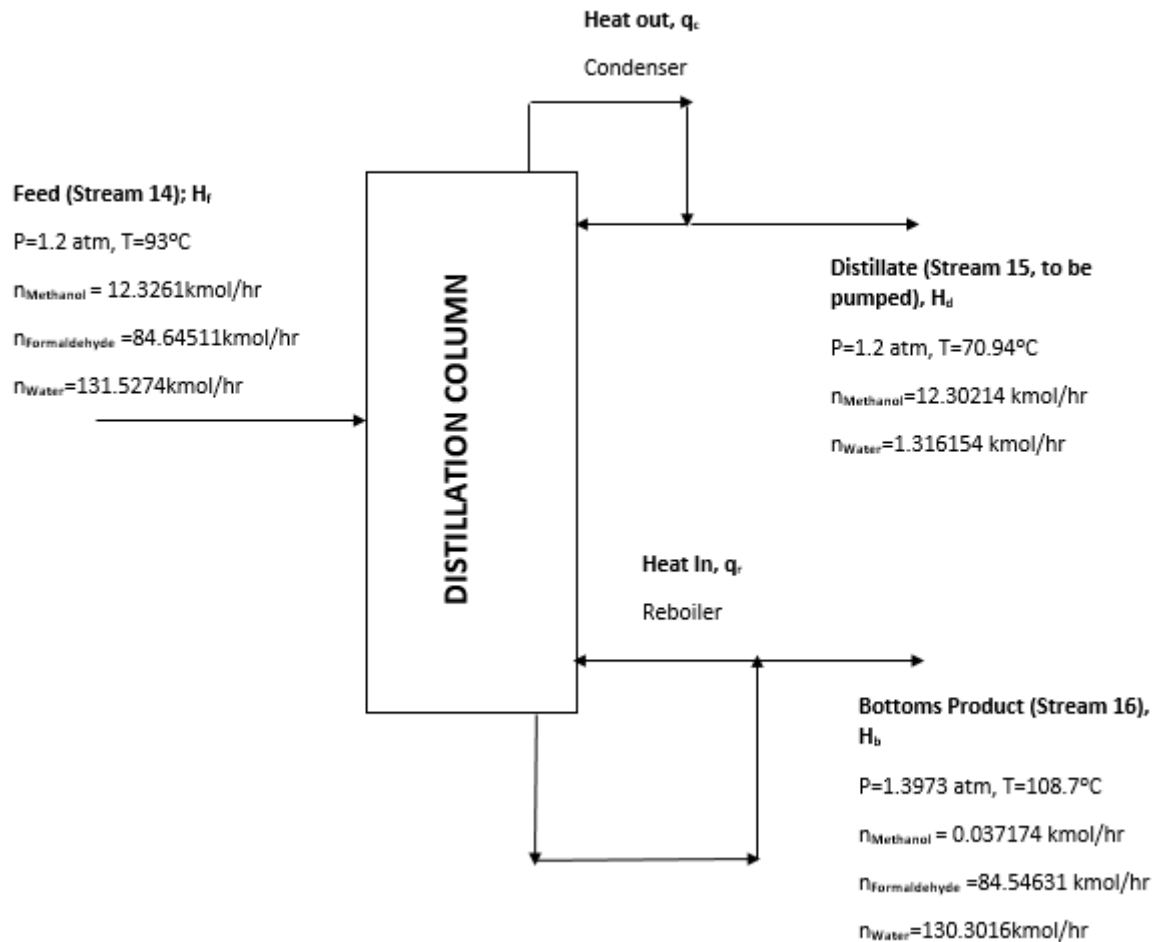
7.1   Selection of Operating Pressure

Operating Pressure in distillation column is fixed by temoerature of the cheapest coolest medium(cooling water), which must be able to condense the over head distillate vapor. For this, bubble point of distillate (in case of total condenser) should be sufficiently greater than cooling water temperature [13] .

Keeping project economics in mind, we have taken our cooling water temperature at $32C(305K)$,which is easily available.

From the VLE Data [17], bubble point of distillate at various pressures:

At $P = 1atm$,

$$T = 339.22K$$

Heat out, q_c

Condenser

Feed (Stream 14); H_f

P=1.2 atm, T=93°C

$n_{Methanol}$ = 12.3261kmol/hr

$n_{Formaldehyde}$ =84.64511kmol/hr

n_{Water}=131.5274kmol/hr

DISTILLATION COLUMN

Distillate (Stream 15, to be pumped), H_d

P=1.2 atm, T=70.94°C

$n_{Methanol}$=12.30214 kmol/hr

n_{Water}=1.316154 kmol/hr

Heat In, q_r

Reboiler

Bottoms Product (Stream 16), H_b

P=1.3973 atm, T=108.7°C

$n_{Methanol}$ = 0.037174 kmol/hr

$n_{Formaldehyde}$ =84.54631 kmol/hr

n_{Water}=130.3016kmol/hr

Figure 14: Distillation Column

At $P = 1.1atm$,

$$T = 341.67K$$

At $P = 1.2atm$,

$$T = 343.94K$$

At $P = 1.3atm$

$$T = 346.06K$$

As we keep on increasing the pressure, our operating costs keep on increasing. At atmospheric pressure, we will get minimum operating costs but the bubble point of distillate willl not be sufficiently greaterthan the coolant.

Therefore, assessing the choices, $P = 1.2atm$, seems optimum.

Generally, as studied in the literature, a column operationg under such conditions has a pressure drop of around $20kPa(0.1973atm)$.

Therefore,

$$P_{Bottom} = P_{Top} + \triangle p$$

$$P_{Bottom} = 1.2 + 0.1973 = 1.3973 atm$$

7.2   FUG(Fenske-Underwood-Gilliland) Method

It is the most widely used short cut method for determining number of equilibrium stages for multi-component distillation.Using Fenske's Equation, we find the number of minimum number of theroetical plates required [3] . After that, Underwood's Method is used to find the Minimum reflux ratio. Gilliland's equation is finally used to get the actual number of theoretical plates required using the minimum number of plates required and minimum reflux ratio.

- **Determination of Minimum Number of Trays:Fenske's Equation**

  The Fenske's Equation in used to find the minimum number of theoretical stages required for desired separation.

  Fenske's Equation:

  $$N_m = \frac{\ln\left[\left(\frac{x_{LK}}{x_{HK}}\right)_d \left(\frac{x_{HK}}{x_{LK}}\right)_b\right]}{\ln(\alpha_{LK/HK,avg})}$$

  where $\alpha_{LK/HK,avg}$ =Average Relative Volatility of Light Key with respect to Heavy key

  $(x_{LK}, X_{HK})_d$ =Mole Fraction of Light key and Heavy key in Distillate

  $(x_{LK}, X_{HK})_b$ =Mole Fraction of Light key and Heavy key in Bottom

  Also, generally
  $$\alpha_{LK/HK,avg} = (\alpha_{LK/HK,Top} \times \alpha_{LK/HK,Bottom})^{\frac{1}{2}}$$

  Now,
  $$\alpha_{i/j} = \frac{\left(\frac{\text{Mole Fraction of i in vapour}}{\text{Mole Fraction of i in liquid}}\right)}{\left(\frac{\text{Mole Fraction of j in vapour}}{\text{Mole Fraction of j in liquid}}\right)}$$

  LK=Methanol(M) and HK=WaterW

  Now, for $\alpha_{M/W,top}$, referring to the VLE Diagram Figure 15:

  From Figure 15, $\alpha_{M/W,top} = 2.509$

  For $\alpha_{M/W,bottom}$, referring to the VLE Diagram Figure 16:

  From Figure 16, $\alpha_{M/W,bottom} = 7.7650$
  Therefore,
  $$\alpha_{LK/HK,avg} = (2.509 \times 7.7650)^{\frac{1}{2}} = 4.41$$

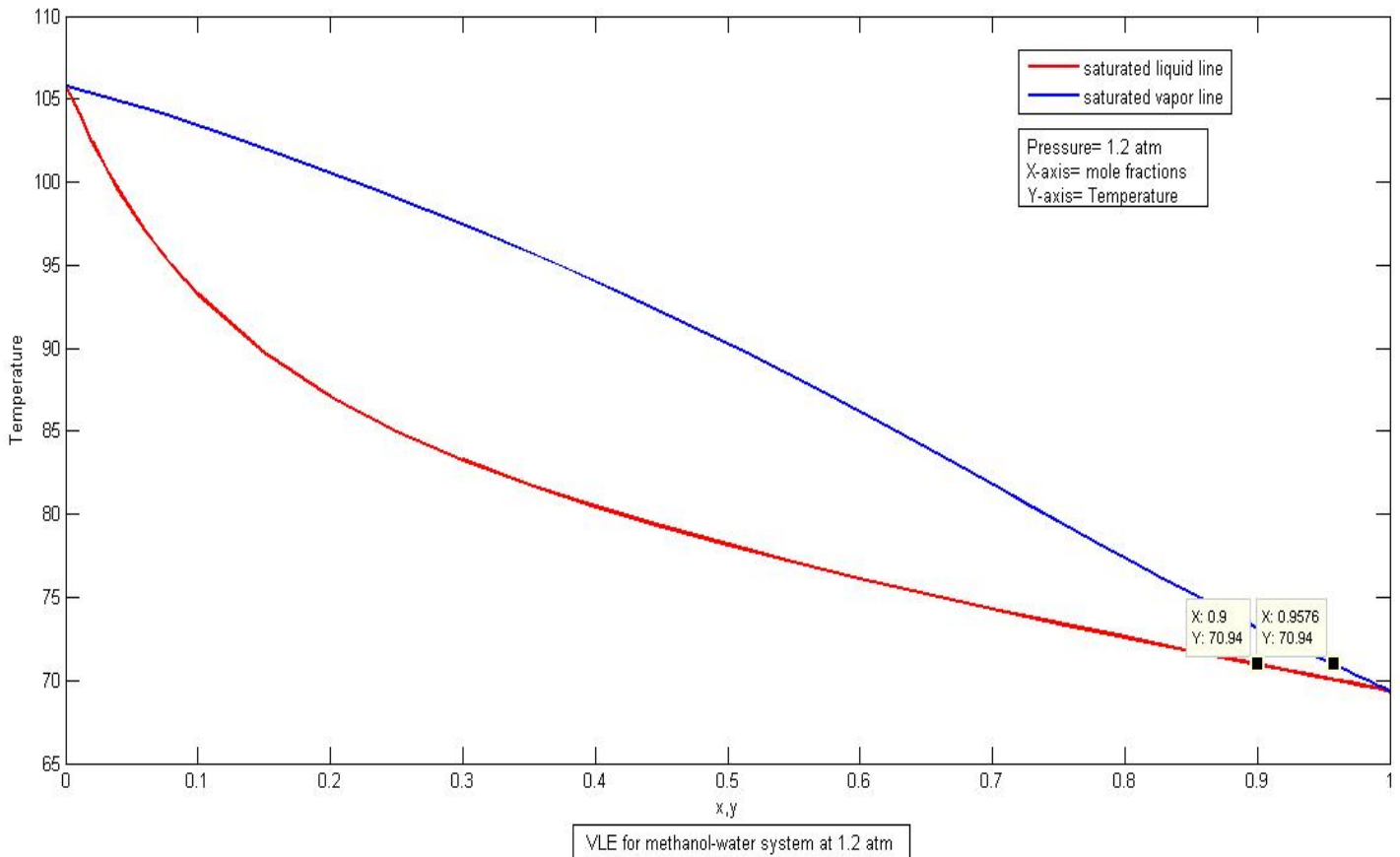 From Kister [6] , this condition holds true only if following condition is satified:

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Figure 15: VLE Diagram for Methanol-Water system at 70.94 C and 1.2 atm [17]

$$\frac{\alpha_{top} - \alpha_{bottom}}{\alpha_{top} + \alpha_{bottom}} \leq 0.1 \ln\left[\frac{\alpha_{top} + \alpha_{bottom}}{2}\right]$$

By substituting the values of $\alpha_{top} and \alpha bottom$, the above condition holds true. Therfore we can use $\alpha_{avg} = 4.41$

Thus, substituting all the values in Fenske's Equation , we get

$$N_m = \frac{\ln\left[\left(\frac{0.903}{0.0969}\right)\left(\frac{0.6063}{1.72*10^{-4}}\right)\right]}{\ln 4.41}$$

$$= 7.007 \approx 7$$

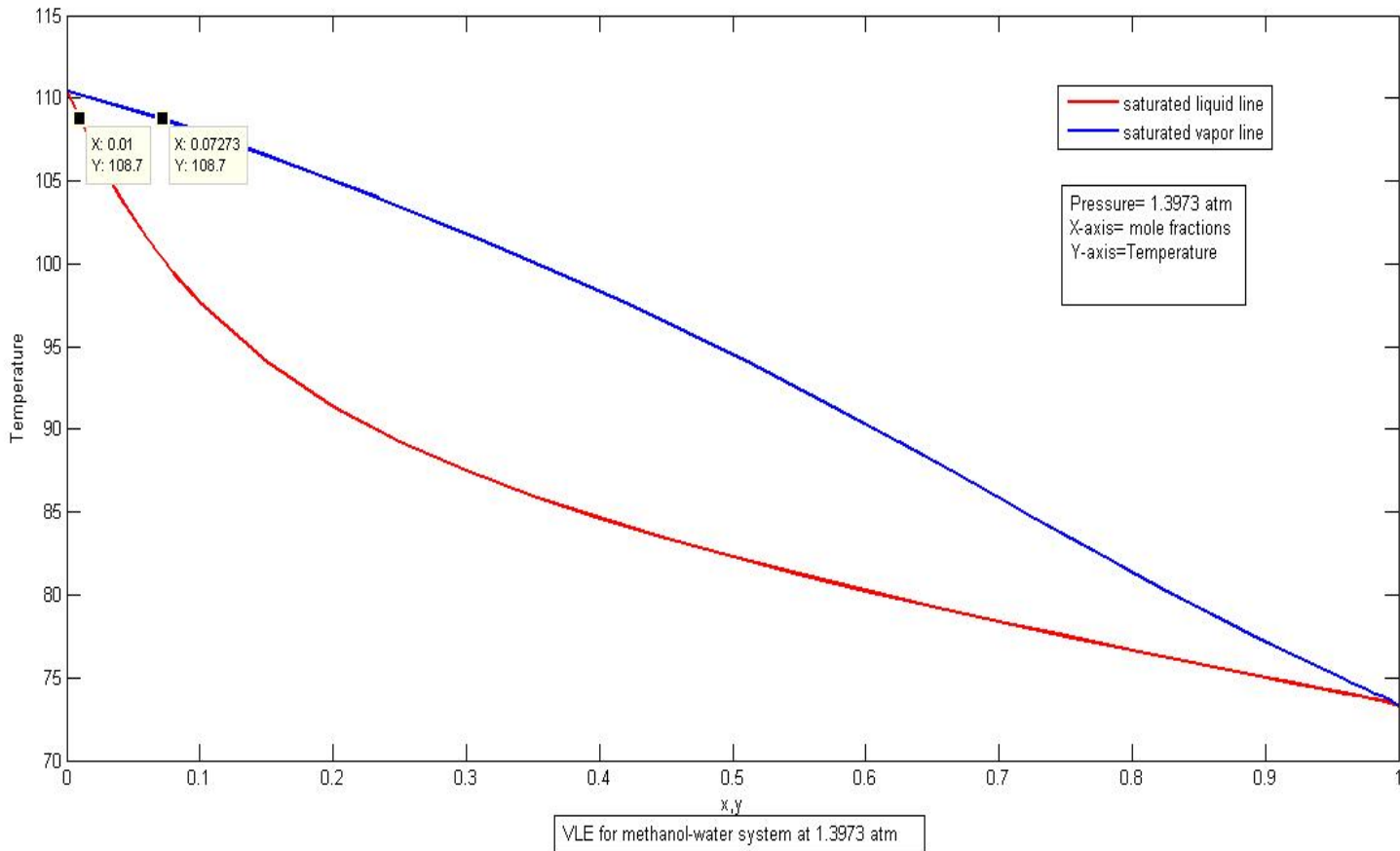The Minimum number of Trays (Theoretical) required are 7 trays.

Figure 16: VLE Diagram for Methanol-Water system at 108.7 C and 1.3973atm [17]

- **Determination of Minimum Reflux Ratio: Underwood's Equation**

  The Underwood's Equation is used to calculate the Minimum Reflux Ratio required.

  Underwood's Equation:

  $$\sum \frac{\alpha_i x_{id}}{\alpha_i - \nu} = R_m + 1$$

  $$\sum \frac{\alpha_i x_{if}}{\alpha_i - \nu} = 1 - q$$

  where, α_i = Average Relative Volatility of each component i with respect to Heavy Key(here, Water)

  x_{id} = Mole Fraction of component i in Distillate

  x_{if} = Mole Fraction of component i in Feed

  q = Quality of feed (Here, Saturated Liquid $q = 1$)

  ν = Constant whose value lies between the relative volatility if heavy key and light key

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Now, to find the relative volatilities of formaldehyde and methanol with respect to water at mean column temperature of 89.8 C:

For $\alpha_{F/W}$, referring to the VLE Diagram Figure :

From Figure , $\alpha_{F/W} = 2.21$

For $\alpha_{M/W}$ at $89.8C$, referring to the VLE Diagram Figure :

$$\alpha_{M/W} = 5.48$$

$$\alpha_{W/W} = 1$$

Now, applying the Underwood's Equation to calculate ν:

Equation to be used:

$$\sum \frac{\alpha_i x_{if}}{\alpha_i - \nu} = 1 - q$$

$$q = 1 (\text{Saturated Liquid})$$

Therefore,

$$\sum \frac{\alpha_i x_{if}}{\alpha_i - \nu} = 0$$

Table 16: Relative Volatilities (wrt to Water) and the Composition in the Feed Stream

Component	α	Mole fraction in feed
Methanol	5.48	0.0539
Water	1	0.5756
Formaldehyde	2.21	0.3704

$$\Rightarrow \frac{5.48 \times 0.0539}{5.48 - \nu} + \frac{1 \times 0.5756}{1 - \nu} + \frac{2.21 \times 0.3704}{2.21 - \nu} = 0$$

Solving the above equation using MATLAB:

We get $\nu = 4.8367, 0.9853$.

The value 4.8367 lies between $(1, 5.48)$.

Therefore, value of $\nu = 4.8367$.

Now, to find R_m,

$$\sum \frac{\alpha_i x_{id}}{\alpha_i - \nu} = R_m + 1$$

Now, $R_m + 1 = 7.67$

$$R_m = 6.67$$

Table 17: Relative Volatilities (wrt to Water) and the Composition in the Distillate Stream

Component	x_d	α	$\frac{\alpha_i x_{id}}{\alpha_i - \nu}$
Water	0.0969	1	-0.0252
Methanol	0.903	5.48	7.6922

Minimum Reflux Ratio= 6.67

Assuming the actual Reflux Ratio to be 1.3 times R_m.

$$R = 1.3 \times R_m$$

$$R = 1.3 \times 6.67 = 8.671$$

- **Determination of Actual Number of Theoretical Plates: Gilliland's Equation:**

  In Gilliland's Equation, we find $X = \frac{R - R_m}{R+1}$, corresponding to which we read $Y = \frac{N - N_m}{N+1}$. Since we know all other values other than N, we can easily calculate N.
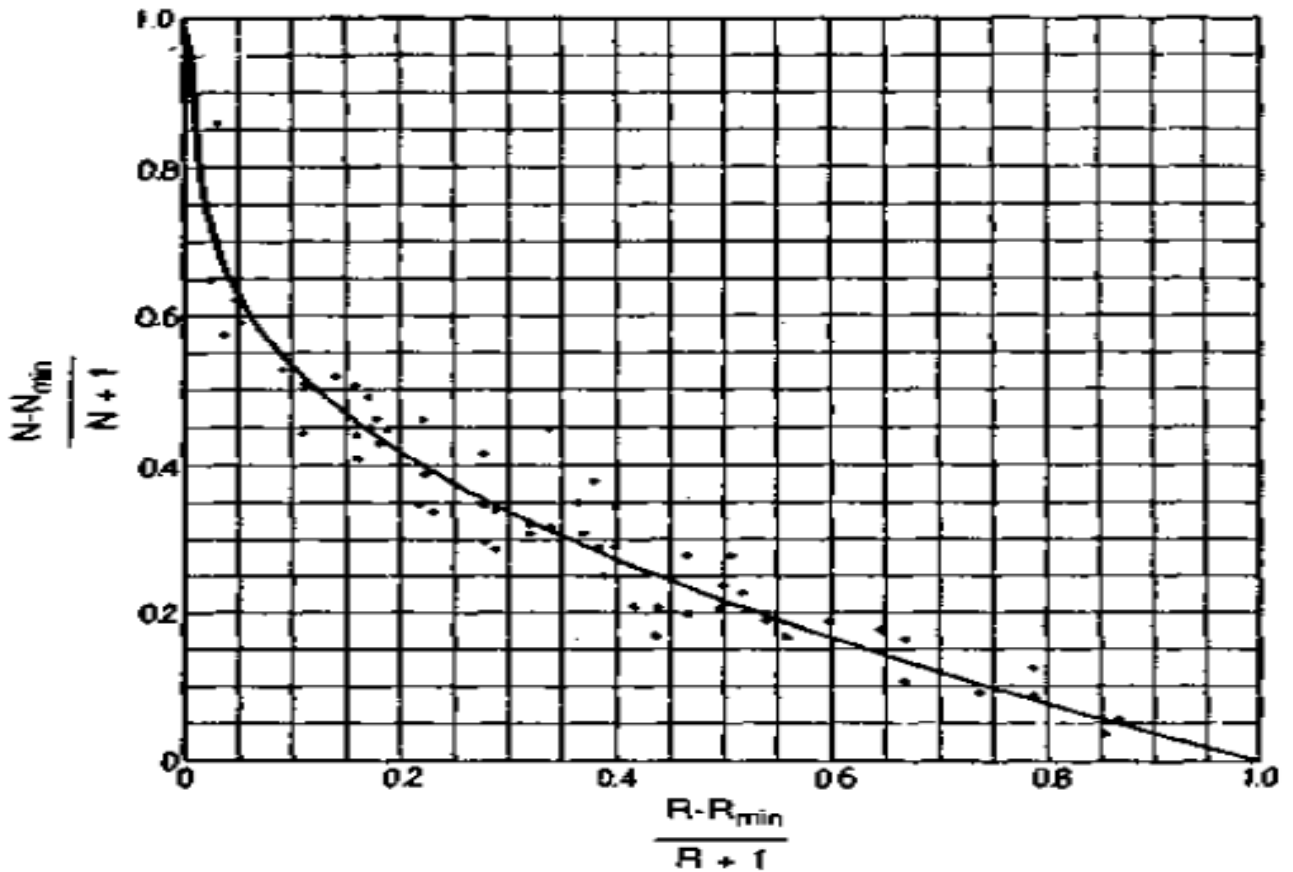
Figure 17: Gilliland's Correlation [6]

$$R_m = 6.67$$

$$N_m = 7$$

$$R = 8.671$$

Now,

$$X = 0.2069$$

Correspondingly, we get $Y = 0.43$

$$\frac{N - N_m}{N + 1} = 0.43$$

Solving for N, we get $N = 10.44$ stages.

7.3 Molokanov Approach

Molokanov Equation is used to calculate the actual number of theoretical plates required (same as FUG Method).This method is much more accuarte than Gilliland's Correlation. It gives much precise results. Therefore it is preferable to use the results computed from thus equation.

Molokanov Correlation:

$$Y = 1 - \exp\left[\left(\frac{1 + 54.4X}{11 + 117.2X}\right)\left(\frac{X - 1}{X^{0.5}}\right)\right]$$

where,

$$X = \frac{R - R_m}{R + 1}$$

Now, $R_m = 6.67, R = 8.671$
Substituting these values, we get $X = 0.2069$

$$X = 0.2069$$

Substituting X to get the values of Y:

$$Y = 0.45326$$

Now,

$$Y = \frac{N_m + Y}{1 - Y}$$

Therefore, we get $N = 9.39$ stages.
This values is comparable to the number of stages calculated using FUG Method.

7.4 Optimum Feed Stage

To get the location of the optimum feed stage, we use the following equation:

$$\frac{N_{Top}}{N_{Bottom}} = \frac{\ln\left[\frac{(D/F)_{Methanol}}{(D/F)_{Water}}\right]}{\ln\left[\frac{(F/B)_{Methanol}}{(F/D)_{Water}}\right]}$$

Now,

$$(\frac{D}{F})_{Methanol} = \frac{12.30214}{12.3261} = 0.9980$$

$$(\frac{D}{F})_{Water} = \frac{1.3161}{131.5274} = 0.01$$

$$(\frac{F}{B})_{Methanol} = \frac{12.3261}{0.037174} = 331.605$$

$$(\frac{F}{D})_{Water} = \frac{131.5274}{130.3016} = 1.009$$

On solving we get,

$$\frac{N_{Top}}{N_{Bottom}} = 3.83749$$

Now, optimum feed stage location (from the top) is given by:

$$\frac{N_{Total}}{1 + \frac{N_{Bottom}}{N_{Top}}} = \text{Optimum Stage from the Top}$$

Taking $N_{Total} = 10.44$ Stages(Using Gilliland's Correlation)

$$\frac{10.44}{1 + \frac{1}{3.83749}} = 8.28 \approx 8$$

Thus feed should be introduced on **stage 8** fom the top.

7.5 Determination of Sieve Tray Tower Diameter

Sieve tray column is decided to be used in the design. This decision is based upon the compatibility of this tray type with our methanol- formaldehyde-water separation process. These features include moderate capacity, relatively high efficiency, low cost, low fouling tendency and low maintenance requirements. We will start with the calculation of the vapor flooding velocity, the operating velocity and finally sizing the actual diameter of the column. This approach is to be applied to the rectifying section and extended to the stripping section of the column.

- **Rectifying Section Diameter**

  Flooding velocity through sieve tray tower can be determined by the following equation:

  $$v_f = C_f(\frac{\sigma}{0.02})^{0.2}(\frac{\rho_L - \rho_V}{\rho_v})^{0.5}$$

  where,

v_f = Flooding Vapour velocity based on net cross-sectional area, m/s

C_f = Flooding Constant which can be obtained from Figure

σ = Surface Tension of Liquid, N/m

ρ_V = Density of vapour, kg/m^3

ρ_L = Density of liquid, kg/m^3

F_{LV} =Liquid-vapour flow factor= $\frac{L_W}{V_W}\left(\frac{\rho_V}{\rho_L}\right)^{0.5}$

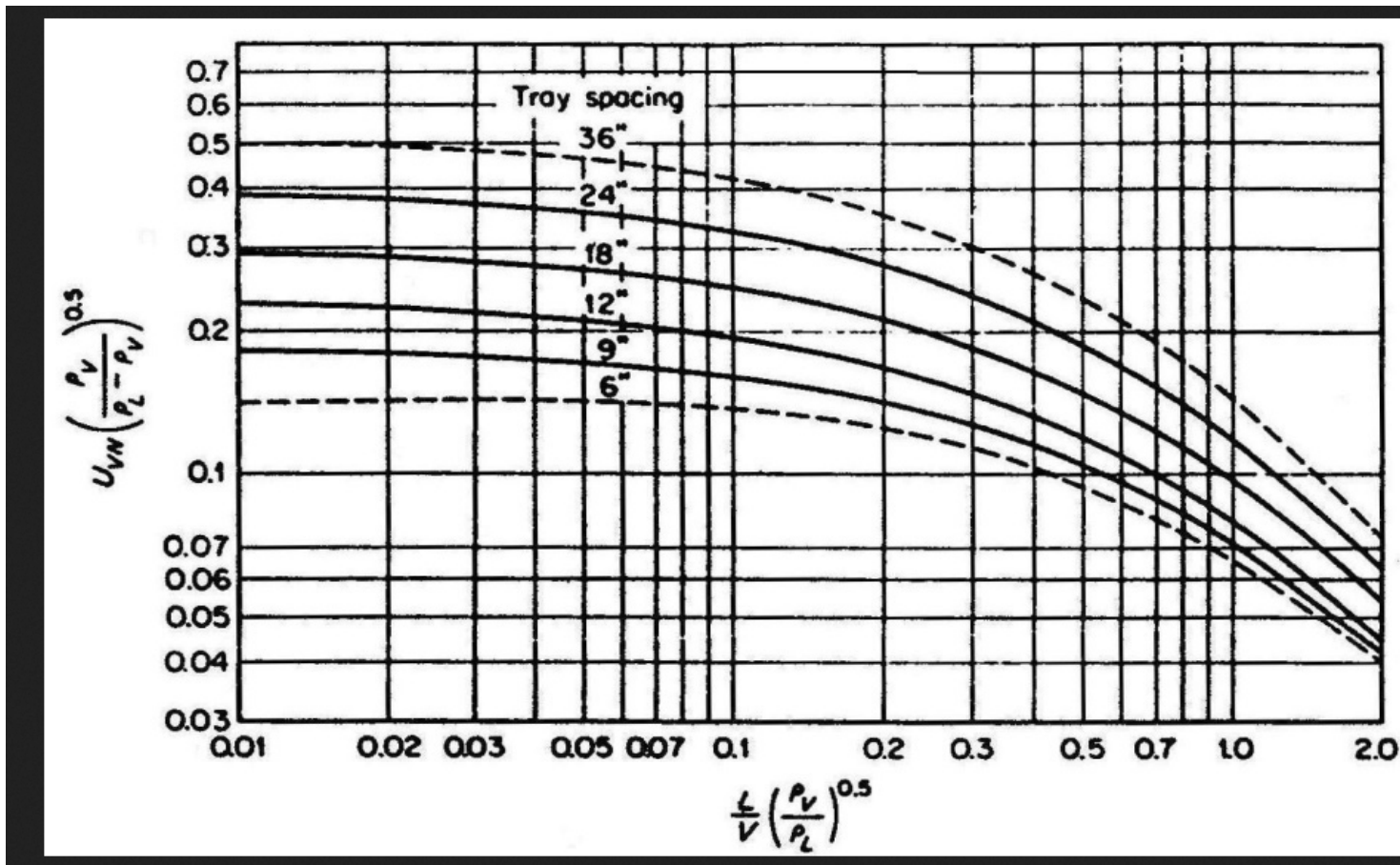L_W =Mass flow rate, kg/s

V_W =Mass flow rate,kg/s

Figure 18: Flooding Velocity Correlation for Sieve Tray Tower [6]

Now, Molar flow rate of vapour and liquid at the top in the rectifying section:

$$L = RD$$

$$L = 8.671 \times 13.61829$$

$$L = 118.08 kmol/hr$$

Also, $V = (R+1)D$

$$V = 7.76 \times 13.61829$$

$$V = 131.5256 kmol/hr$$

Density of vapour at top, ρ_V-

$$\rho_V = \frac{P \times M_{avg}}{RT}$$

$$M_{avg} = 32 \times 0.9033 + 18 \times 0.0966$$
$$M_{avg} = 30.6444$$

$$\rho_V = \frac{30.6444}{344.09} \frac{273}{22.414}$$
$$\rho_V = 1.0847 kg/m^3$$

Density of liquid at top, ρ_L-

$$\rho_L = \frac{1}{\sum \frac{w_i}{\rho_i}}$$

$$w_{methanol} = 0.9432$$

$$w_{water} = 0.0567$$

$$\rho_{methanol at 70.94°C=742.266 kg/m^3}$$

$$\rho_{water at 70.94°C=968.435 kg/m^3}$$

$$\rho_L = \frac{1}{\frac{0.9432}{742.266} + \frac{0.0567}{968.435}}$$

$$\rho_L = 752.303 kg/m^3$$

Now, $F_{LV} = \frac{L_W}{V_W} \left(\frac{\rho_V}{\rho_L} \right)^{0.5}$

$$\frac{L_W}{V_W} = \frac{L}{V} = \frac{118.08}{131.5256}$$

$F_{LV} = 0.897 \times \frac{1.0847}{752.303}^{0.5} = 0.034$

For the first trial calculations, let the tray spacing be $12 inch$ or $0.3m$. Referrring to Figure 18:

$$C_f = 0.24$$

From Online Dortmund Data Bank [1] :

$$\sigma_{methanol} = 18.2343 \times 10^{-03} N/m$$

$$\sigma_{waterl} = 64.1097 \times 10^{-03} N/m$$

Therefore

$$\sigma_{top} = 0.02266 N/m$$

Using the Figure 18 and the equation :

$$v_f = 1.75290 m/s$$

Let actual vapour velocity through tower be 85% of the flooding velocity.

$$v = 0.85 \times v_f$$

$$v = 0.85 \times 1.75290 = 1.4899 m/s$$

Volumetric flow rate of vapor at top:

$$\dot{Q} = \frac{V \times M_{avg}}{\rho_V} = \frac{131.5256 \times 30.6444}{1.0847} = 3715.795 m^3/hr = 1.032 m^3/s$$

Net Area required at the top:

$$A_n = \frac{\dot{Q}}{v} = \frac{1.032}{1.4899} = 0.6926 m^2$$

Let downcomer area be 10% of the net area:

$$A_d = 0.10 A_n$$

$$A_n = A_c - A_d$$

where, A_c =Inside cross sectional area of the tower

$$0.90 A_c = 0.6926$$
$$A_c = 0.7695 m^2$$

Therefore Inside diameter of the column at the top:

$$D_i(top) = \sqrt{\frac{4 \times A_c}{\pi}}$$

$$D_i(top) = 0.9898 m$$

- **Stripping Section Diameter**

Following the same procedure as we did for the rectifying section:

Molar flow rates of vapour and liquid in the stripping section:

$$\overline{L} = L + Fq = L + F = 118.08 + 228.4986 = 346.582 kmol/hr$$

$$\overline{V} = F(q-1) + V = V = 131.5256 kmol/hr$$

Therefore $\frac{\overline{L}}{\overline{V}} = \frac{118.08}{131.5256} = 0.8972$

Density of vapour at bottom, ρ_V-

$$\rho_V = \frac{P \times M_{avg}}{RT}$$

$$M_{avg} = 30 \times 0.3934 + 32 \times 0.000172 + 18 \times 0.6063$$
$$M_{avg} = 22.7239$$

$$\rho_V = \frac{141.581 \times 22.7239}{381.85} \frac{273}{22.414 \times 101.325}$$
$$\rho_V = 1.012 kg/m^3$$

Density of liquid at bottom, ρ_L-

$$\rho_L = \frac{1}{\sum \frac{w_i}{\rho_i}}$$

$$w_{methanol} = 2.436 \times 10^{-04}$$
$$w_{water} = 0.4803$$
$$w_{formaldehyde} = 0.5194$$
$$\rho_{methanol at 108.7°C=699.482kg/m^3}$$
$$\rho_{water at 108.7°C=940.744kg/m^3}$$
$$\rho_{formaldehyde at 108.7°C=766.706kg/m^3}$$

$$\rho_L = \frac{1}{\frac{0.4803}{940.744} + \frac{2.436 \times 10^{-04}}{699.482} + \frac{0.5194}{766.706}}$$

$$\rho_L = 841.75 kg/m^3$$

Now, $F_{LV} = \frac{L_W}{V_W} \left(\frac{\rho_V}{\rho_L}\right)^{0.5}$

$F_{LV} = 0.8972 \times \frac{1.012}{841.75}^{0.5} = 0.031$

Let the tray spacing be $12 inch or 0.3m.$

Referrring to Figure 18:

$$C_f = 0.245$$

From Online Dortmund Data Bank [1]:

$$\sigma_{methanol} = 14.8202 \times 10^{-03} N/m$$

$$\sigma_{water} = 57.1663 \times 10^{-03} N/m$$

$$\sigma_{formaldehyde} = 0.00132417 N/m$$

Therefore
$$\sigma_{bottom} = 0.03518 N/m$$

Using the Figure 18 and the equation :

$$v_f = 2.06520 m/s$$

Let actual vapour velocity through tower be 85% of the flooding velocity.

$$v = 0.85 \times v_f$$

$$v = 0.85 \times 2.06520 = 1.755427 m/s$$

Volumetric flow rate of vapor at top:

$$\dot{Q} = \frac{\overline{V} \times M_{avg}}{\rho_V} = \frac{131.5256 \times 22.7239}{1.0847} = 2953.332 m^3/hr = 0.82037 m^3/s$$

Net Area required at the top:

$$A_n = \frac{\dot{Q}}{v} = \frac{0.82037}{1.755427} = 0.46733 m^2$$

Let downcomer area be 10% of the net area:

$$A_d = 0.10 A_n$$

$$A_n = A_c - A_d$$

where, $A_c =$ Inside cross sectional area of the tower

$$0.90 A_c = 0.46733$$

$$A_c = 0.5193m^2$$

Therefore Inside diameter of the column at the bottom:

$$D_i(bottom) = \sqrt{\frac{4 \times A_c}{\pi}}$$

$$D_i(bottom) = 0.8131m$$

Since both the top and bottom diameters are less than 1m, tray spacing of $12inch \approx 0.3m$ is acceptable.

7.6 Tray Efficiencies and Column Height

Since the Diameters of the rectifying section and the stripping section are different, a slight change in the tray efficiency is to be considered in the column design[9]. The efficiency of the trays is to be determined using OConnell Correlation which is estimated the efficiency as a function of the product of the feed liquid viscosity and the relative volatility of the key components. This correlation is based on test data from 31 plant columns, including hydrocarbon, chlorinated hydrocarbon, and alcohol separation columns.

The viscosity is evaluated at the average arithmetic temperature between the column top and bottom temperatures.

The Lockett's Equation Form for O'Conell correlation:

$$E_{OC} = 0.492(\mu_L \alpha)^{-0.245}$$

- **Rectifying Section Efficiency**

  Viscosity of Feed(at $89.82°C$):

  Weight Fractions of components in the feed:

  From Table 10:
  $$w_{methanol} = \frac{394.4353}{5301.281} = 0.0744$$
  $$w_{water} = \frac{2367.492}{5301.281} = 0.4465$$
  $$w_{formaldehyde} = \frac{2539.353}{5301.281} = 0.479$$

  Also, from Dortmund Data Bank [1] , viscosities at $89.82°C$:

  $$\mu_{methanol} = 0.20944cP$$

  $$\mu_{water} = 0.30583cP$$

  $$\mu_{formaldehyde} = 0.00546cP$$

Therefore $\mu_{feed} = \sum w_i \mu_i$
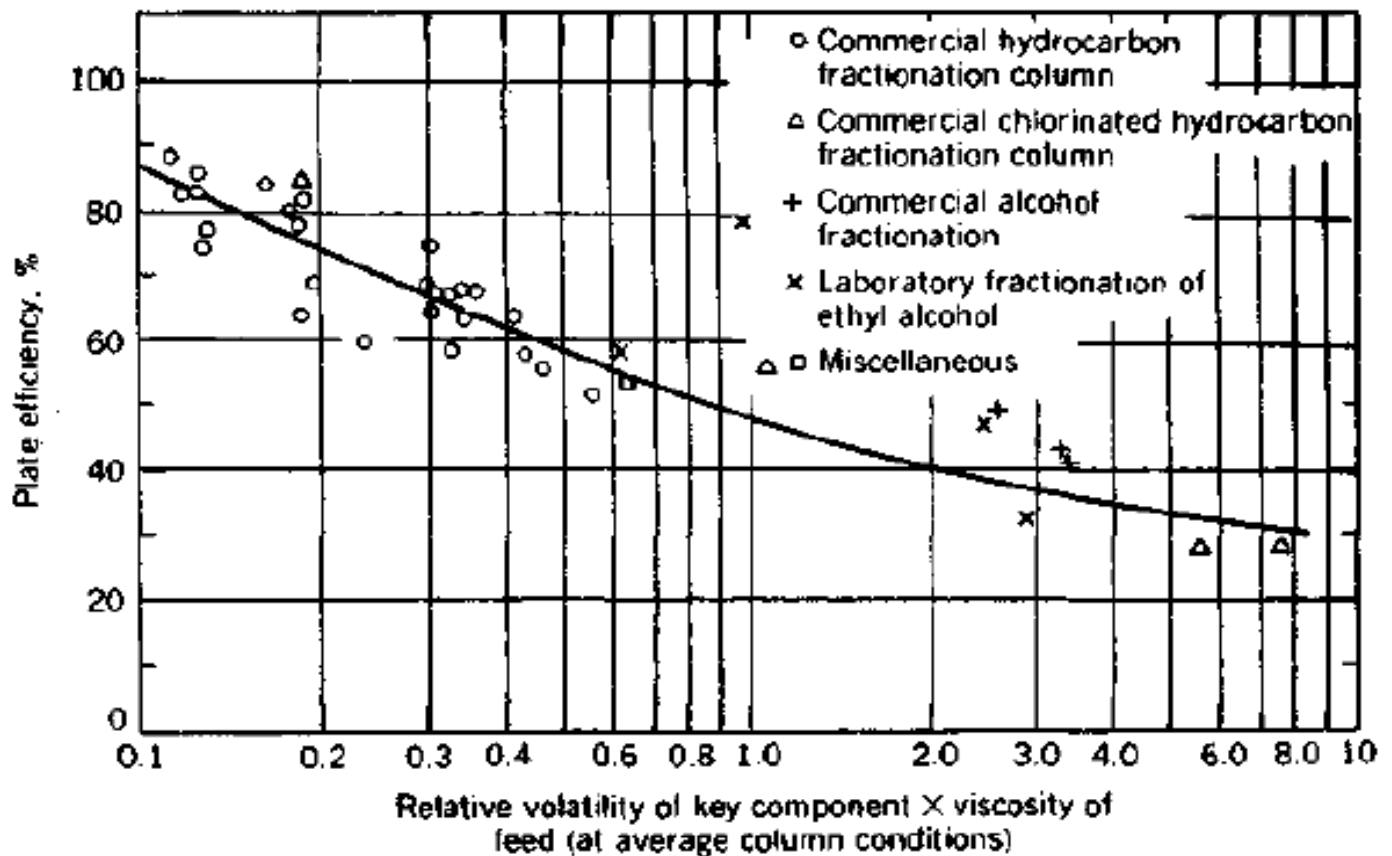
$$\mu_{feed} = 0.15239cP$$

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Figure 19: O'Conell Correlation for Tray Efficiency [6]

Now, $\alpha_{M/W,Top} = 2.509$

Therefore, $(\mu \times \alpha_{M/W,Top}) = 0.3823$

Using Lockett's Equation Form [6] :

Tray Efficiency: 0.622

Using Figure 19

Tray Efficiency: 63%

The efficiencies calculated by the two methods agree well with each other.

Actual Number of trays in the Rectifying section: $\frac{N_{Top}}{TrayEfficiency}$

$$N_{Top,actual} = \frac{8.28}{0.622} = 13.31 \approx 14$$

- **Stripping Section Efficiency**

  Using the same method stated above:

  $$\mu_{feed} = 0.15239 cP$$

  $$\alpha_{M/W,Bottom} = 7.7650$$

Therefore, $(\mu_{feed} \times \alpha_{M/W,Bottom}) = 1.183$

Using Lockett's Equation Form:

Tray Efficiency: 0.472

Using Figure 19

Tray Efficiency: 48%

The efficiencies calculated by the two methods agree well with each other. Now,

Actual Number of trays in the Stripping section: $\frac{N_{Bottom}}{TrayEfficiency}$

$$N_{Bottom,actual} = \frac{2.158}{0.472} = 4.572 \approx 5$$

- **Column Height**

  The column height depends on the tray spacing. In our design method, we have fixed our tray spacing to be $12 inches \approx 0.3m$. [3] A safety factor of 10% is to be considered for the final design height.

  Total Number of Stages= $14 + 5 + 1$(Reboiler Stage)

$$N_{Total} = 20$$

Therefore, Column Height= $TraySpacing \times N_{Total} \times SafetyFactor$

$$ColumnHeight = 0.3048 \times 20 \times 1.1$$

$$ColumnHeight = 6.7056m$$

7.7 Selection of Liquid Flow Pattern over the Sieve Tray

After finding the tower diameter, liquid flow pattern over sieve tray is decided. Liquid flowpatterns depends on the liquid flow rate and tower diameter. [16]

Following table can be used for the selection of liquid flow pattern:

Table 18: Selection of Liquid Flow Pattern

| Column Diameter,m	Range of liquid flow rate, m^3/s			
	Reverse Flow	Cross Flow (Single Pass)	Double Pass	Cascade Double Pass
0.9	$0 - 1.90 \times 10^{-03}$	$1.90 \times 10^{-03} - 0.0126$		
1.2	$0 - 2.50 \times 10^{-03}$	$2.50 \times 10^{-03} - 0.0190$		
1.8	$0 - 3.15 \times 10^{-03}$	$3.15 \times 10^{-03} - 0.0250$	0.025-0.0440	
2.4	$0 - 3.15 \times 10^{-03}$	$3.15 \times 10^{-03} - 0.0315$	0.0315-0.050	
3	$0 - 3.15 \times 10^{-03}$	$3.15 \times 10^{-03} - 0.0315$	0.0315-0.057	0.057-0.0880
3.66	$0 - 3.15 \times 10^{-03}$	$3.15 \times 10^{-03} - 0.0315$	0.0315-0.063	0.063-0.1000
4.6	$0 - 3.15 \times 10^{-03}$	$3.15 \times 10^{-03} - 0.0315$	0.0315-0.070	0.070-0.1136
6	$0 - 3.15 \times 10^{-03}$	$3.15 \times 10^{-03} - 0.0315$	0.0315-0.070	0.070-0.1260
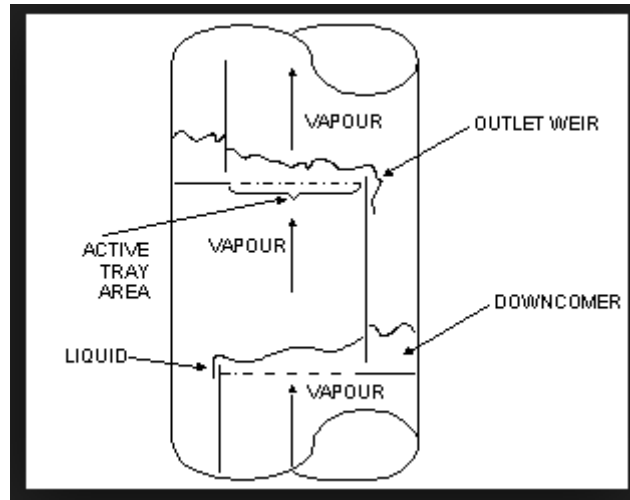
Figure 20: Cross Flow Liquid Flow Pattern

Now, Volumetric Flow rate of Liquids:
For Rectifying Section,

$$\dot{Q}_L = \frac{LM_{avg}}{\rho_L}$$

$$\dot{Q} = \frac{118.08 \times 30.644}{752.303} = 6.96 m^3/hr = 1.96 \times 10^{-03}$$

For Stripping Section,

$$\dot{Q}_L = \frac{\overline{L}M_{avg}}{\rho_L}$$

$$= \frac{346.582 \times 22.7239}{841.75} = 9.3563 m^3/hr = 2.598 \times 10^{-03}$$

Referring to Table 18, single pass tray can be used for both sections.

7.8 Tray Layout and Hydraulics

This section deals with deciding the tray layout and understanding the hydraulics associated with it. It comprises of finding the amount of liquid entrained and desigining the downcomer. We will be using a straight segmental vertical downcomer and a weir. The decision to use a segmental straight downcomer is due to its simple geometry, low cost. Also because it utilizes most of the column area for the large downflow in our system and the ease at which its operated and maintained.[6]

- **Rectifying Section**

  Diameter of the column:$0.9898m$

  – **Entrainment at a Flooding Point of 85% :**
    Fractional Entrainment ψ is found out using the Fair's Entrainment Correlation. Abscissa $(F_{LV} = \frac{L}{G} \left(\frac{\rho_G}{\rho_L}\right)^{0.5})$ is found out and for the corresponding percent of flood, fractional entrainment is found out.
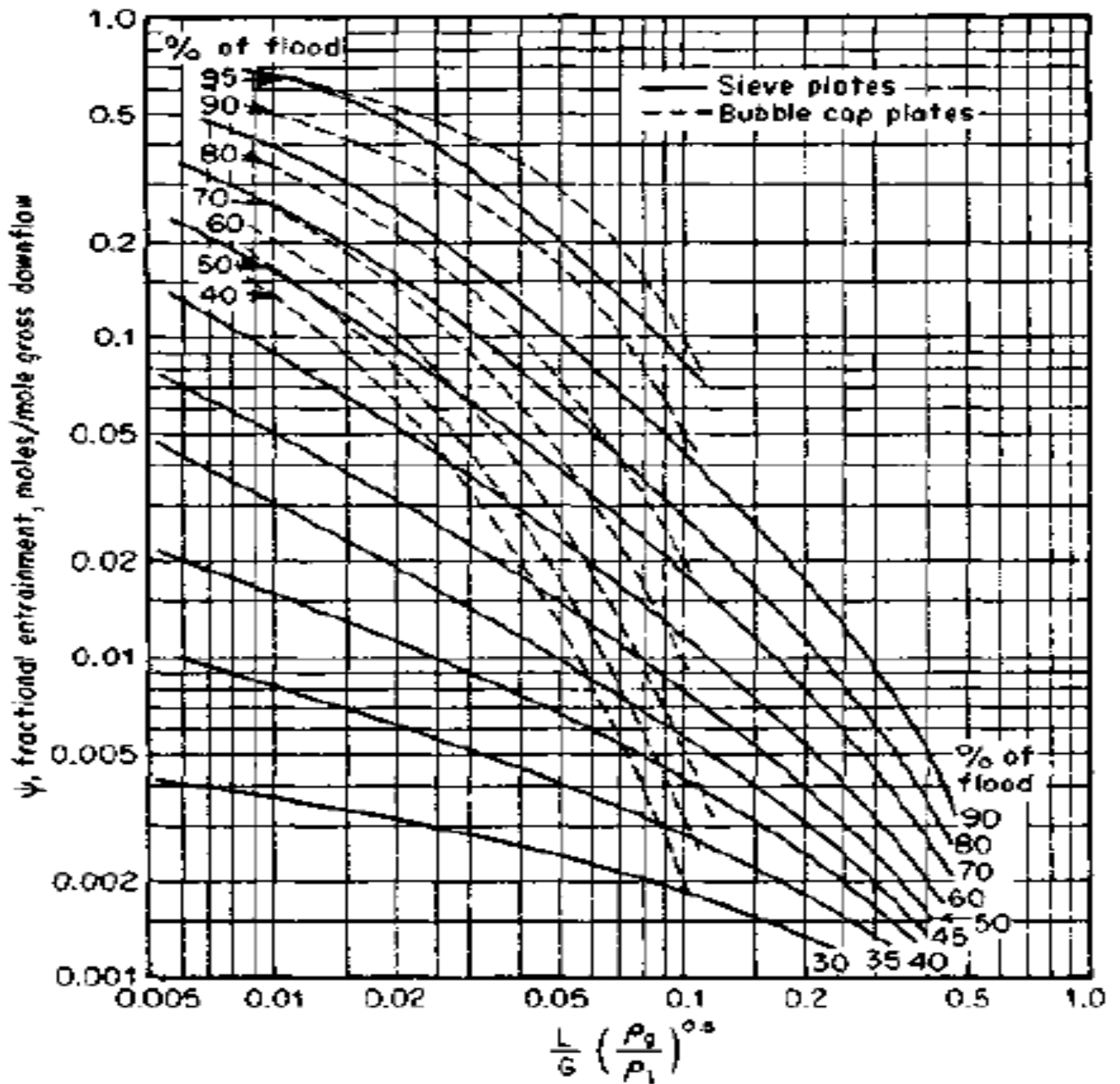
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Figure 21: Fair's Entrainement Correlation [6]

$$F_{LV} = 0.897 \times \frac{1.0847}{752.303}^{0.5} = 0.034$$

Now, $F_{LV} = 0.034$

Percent Flood= $\frac{\text{Actual Vapour Velocity based on net area}}{\text{Flooding Velocity}} \times 100$

We have taken the actual velocity to be 85% of the flooding velocity. Its a standard value to assume.

Therefore, for a percent flood of 85%-

$$\psi = 0.13 = 13\%$$

– **Entrained Liquid:**

$$L_{entrained} = L_e = \frac{\psi}{1 - \psi} \times L$$

$$L_e = \frac{0.13}{1 - 0.13} \times 118.08 = 17.644 kmol/hr$$

– **Amount Entrained on Top:**

$$L + L_{entrained} = L + L_e = 118.08 + 17.644 = 135.724 kmol/hr$$

– **Column Cross Sectional Area:**

$$A_{Total} = \frac{\pi \times Dia^2}{4} = \frac{\pi \times 0.9898^2}{4}$$

$$A_{Total} = 0.76945 m^2$$

– **Downcomer Area:**

$$A_d = (1 - \eta) \times A_{Total}$$

η denotes the fraction of tray area excluding the downcomer portion. Generally its value lies between 0.8 to 0.95. Assuming the value to be 0.9.

$$A_d = (1 - 0.9) \times 0.76945 = 0.076945 m^2$$

– **Weir Length and Height**
For a distillation column operating above atmospheric pressure, weir height is kept in between 40 mm to 90 mm; while 40mm to 50 mm is more common.
Therefore, Weir height is kept as $h_w = 45mm$.

Weir Length is normally kept in between 0.6 to 0.85 times the column diameter. For $\frac{A_d}{A_{Total}} = 0.1$, 0.727 can be used as the multiplying factor.
Therefore, Weir Length= $0.727 \times 0.9898 = 0.7196 m = 28.331 inch = 2.361 ft$

– **Active Area of the Tray:**

$$\eta = 0.9$$

$$A_{Active} = A_{Total} \times (1 - 2(1 - \eta))$$

Therefore, $A_{Active} = 0.61556 m^2$

– **Total Area of holes:**

$$A_{Holes} = A_{Active} \times \beta$$

The tray selected is a standard 14 gauge tray with thickness $(T_{tray}) = 0.078$ in with a common hole diameter $d_o = \frac{3}{16}$ inch for normal operation and clean service. Pitch standard spacing between the holes is of $3.8d_o = 0.7125$ inches. A 2.5 inch space between the edge holes and the column wall is chosen, and a space of 4 inch between the edge hole and the tray weir. All these values are standard values used in the industry.

Generally $\beta = 0.1$

Therefore, $A_{Holes} = 0.1 \times 0.61556 = 0.061556 m^2$

– **Vapor velocity through Tray Hole:**

$$v = \frac{V \times M_{avg}}{3600 \times \rho_V \times A_{Holes}}$$

$$v = \frac{131.5256 \times 30.644}{3600 \times 1.0847 \times 0.061556}$$

$$v = 16.76 m/s$$

– **Orifice Coefficient:**

It is used in the calculation of dry pressure drop through the tray. Orifice Coefficient C_o is found out using the $\frac{T_{Tray}}{HoleDiameter}$ and $\frac{A_{Holes}}{A_{Active}}$. Following figure gives the correlation used to find C_o:

Now,

$$\frac{A_{Holes}}{A_{Active}} = \frac{0.061556}{0.61556} = 0.1$$

$$\frac{T_{Tray}}{HoleDiameter} = \frac{0.078}{0.1875} = 0.416$$

Using Figure 22,

$$C_o = 0.73$$

Using a much accurate correlation given by Hughmark and O'Conell:

$$C_o = 0.85032 - 0.04231 \frac{d_o}{T_{Tray}} + 0.00017954 \frac{d_o}{T_{Tray}}^2$$

Calculating, $C_o = 0.759$

– **Total Head Loss of Liquid through dry Tray:**

The following equation was given by Ludwig:

$$h_{\triangle P,dry} = \frac{0.003 \times v^2 \times \rho_V (\frac{\rho_{Water}}{\rho_L})(1 - \beta^2)}{C_o^2}$$
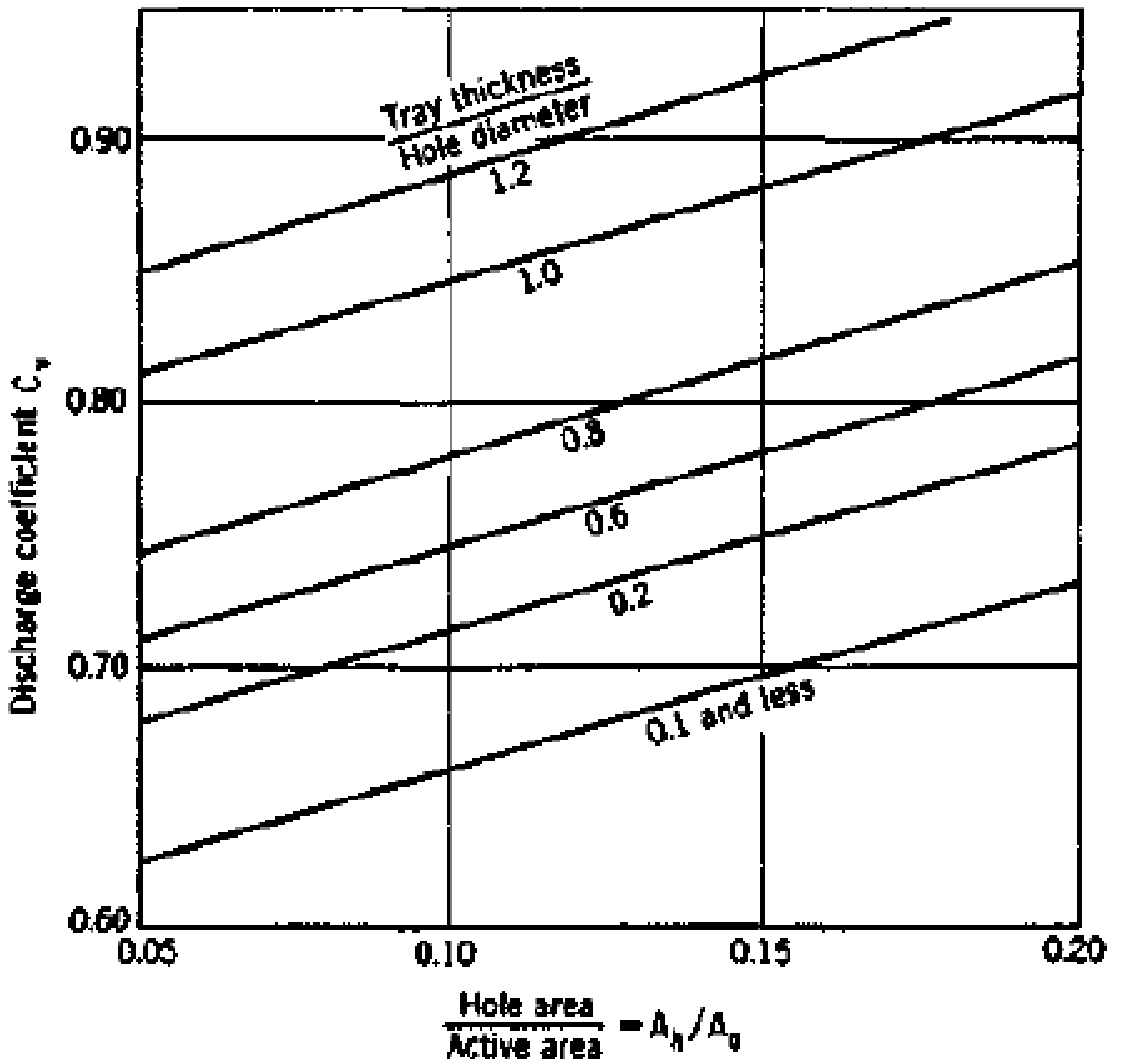
Now, $v = 16.76 m/s = 54.48 ft/sec$

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Figure 22: Orifice Coefficient Correlation for Sieve Tray [6]

$$\rho_V = 1.0847 kg/m^3 = 0.0677 lb/ft^3$$

Therefore, on calculating

$$h_{\triangle P,dry} = 1.377 inch = 0.03497 m$$

– **Liquid Height over the weir:**
Now, Liquid height over the weir$=h_{crest} = h_{ow} = 0.48F_w(Q_L)^{\frac{2}{3}}$

where $F_w = WeirCorrectionFactor$

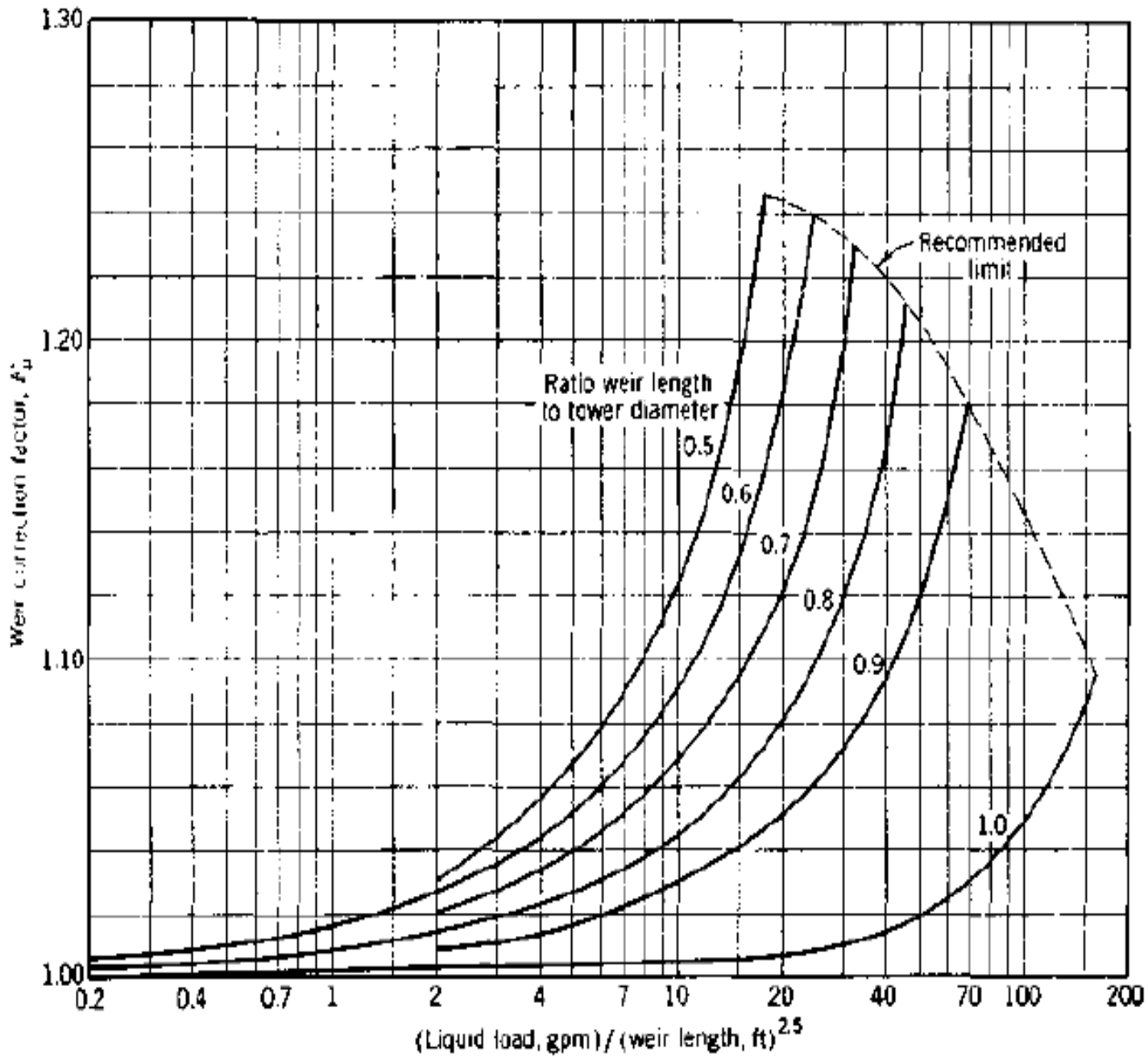$$Q_L = \frac{\text{Liquid Load in gallons per minute}}{\text{Length of the Weir}}$$

Figure 23: Weir Correction Factor for Segmental Downcomers in calculation of liquid head over the weir [6]

Now Liquid Load in gallons per minute, L_g:

$$L_g = \frac{(L + L_e)M_{avg}}{\rho_L}$$

$$L_g = \frac{135.724 \times 30.644}{752.303} = 5.5285 m^3/hr = 1460 gallons/hr = 24.33 gallons/minute = 0.0542076 ft^3$$

Abscissa= $\frac{L_g}{WeirLength,ft^{2.5}}$

$$Abscissa = \frac{24.33}{2.361^{2.5}} = 2.8405$$

Ratio weir length to tower diameter= $\frac{0.7196}{0.9898} = 0.727$

Using Figure 23, $F_w = 1.026$

Now, $h_{ow} = h_{crest} = 0.48 F_w (Q_L)^{\frac{2}{3}}$

$$h_{crest} = 0.48 \times 1.026 \times (\frac{24.33}{28.331})^{\frac{2}{3}}$$

$$h_{crest} = 0.4449 in = 0.0113 m$$

– **Liquid Fractional Head Loss:**

The flow area under the downcomer is calcualted as follows:

$$A_{du} = gap \times WeirLength$$

The gap is usually 1 inch=0.083 ft in all the distillation columns working above the atmospheric pressure.

Therefore,

$$A_{du} = 0.083 ft \times 2.361 ft = 0.195963 ft^2$$

Fractional loss in liquid head, $h_{du} = 0.56 \times \frac{L_g, ft^3/sec^2}{A_{du}}$

$$h_{du} = 0.56 \times \frac{0.0542076^2}{0.195963} = 0.04285 inch = 0.00108839 m$$

– **Residence Time:**

The total pressure head on the downcomer is the summation of all the hydrodynamic effects determined previously as follows:

$$h_{dc} = h_{\triangle P,dry} + h_{weir} + h_{crest} + h_{grad} + h_{du}$$

h_{grad} is generally zero for sieve tray tower.

$$h_{dc} = 0.03497 + 0.045 + 0.0113 + 0.00108839 = 0.092 m$$

The residence time is calculated as follows:

$$\theta_{res} = \frac{A_d h_{dc} \rho_L}{(L + L_e) M_{avg}}$$

$$\theta_{res} = \frac{0.076945 \times 0.092 \times 752.303}{135.724 \times 30.644} = 1.2804 \times 10^{-03} = 4.609 sec$$

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– **Flooding Check:**

As calculated above, total pressure head on the downcomer, $h_{dc} = 0.092m$

The actual aerated head, $h_{dc,aerated} = \frac{h_{dc}}{\phi}$

where ϕ = froth density at normal operation

Now,

$$\phi = \frac{1}{\eta + 1}$$

$$\eta = 0.9$$

Therefore, $\phi = \frac{1}{1+0.9} = 0.52$

$$h_{dc,aerated} = \frac{0.092}{0.52} = 0.1769$$

Since, $h_{dc,aerated} = 0.1769m$ is less than tray spacing which is $12 inch \approx 0.3m$.

The aerated liquid head is much less than the tray spacing which is 12 inch, there would be no operational problem and the liquid flooding is regulated.

– **Weeping Check:**

An analysis is done to check for the operation to be above the weeping and dumping points and avoid excessive weeping. Surface Tension head h_σ is calculated and follwing condition is checked for weeping:

Condition for weeping:

$$h_{\triangle P,dry} + h_\sigma \geq 0.10392 + 0.21599X - 0.021675X^2$$

where $X = h_{weir} + h_{crest} + h_{grad}$

Now, $\sigma = 0.02266N/m$

$$\rho_L = 752.303kg/m^3 = 46.964lb/ft^3$$
$$d_o = \frac{3}{16}inch = 0.0047625m$$
$$h_{\triangle P,dry} = 0.032197m = 1.377inch$$
$$h_{weir} = 45mm = 0.045m = 1.771in$$
$$h_{crest} = 1.60875inch$$

Calculation for X:

$$X = 1.771 + 1.60875 + 0 = 3.3804inch$$

Now, $0.10392 + 0.25199 \times 3.3804 - 0.021675 \times 3.3804^2 = 0.70806$

Now,

$$h_\sigma = \frac{0.04 \times \sigma}{\rho_L \times d_o} = \frac{0.04 \times 22.66}{46.94 \times \frac{3}{16}} = 0.102932inch$$

Therefore $h_{\triangle P,dry} + h_\sigma = 1.3770 + 0.102932 = 1.479932inch$

Condition:

$$h_{\triangle P,dry} + h_\sigma \geq 0.10392 + 0.21599X - 0.021675X^2$$

$$1.479932 \geq 0.70806$$

Condition is satisfied. Hence, no weeping or dumping occurs.

$$h_{\triangle P,dry} + h_\sigma \geq 0.10392 + 0.21599X - 0.021675X^2$$

- **Stripping Section**

  Diameter of the column: $0.8131m$

  - **Entrainment at a Flooding Point of 85% :**

    Fractional Entrainment ψ is found out using the Fair's Entrainment Correlation (Figure 21). Abscissa ($F_{LV} = \frac{L}{G} \left(\frac{\rho_G}{\rho_L} \right)^{0.5}$) is found out and for the corresponding percent of flood, fractional entrainment is found out.

    $$F_{LV} = 0.8972 \times \frac{1.012}{841.75}^{0.5} = 0.031$$

    Now, $F_{LV} = 0.031$

    Percent Flood$= \frac{\text{Actual Vapour Velocity based on net area}}{\text{Flooding Velocity}} \times 100$

    We have taken the actual velocity to be 85% of the flooding velocity. Its a standard value to assume.

    Therefore, for a percent flood of 85%-

    $$\psi = 0.15 = 15\%$$

  - **Entrained Liquid:**

    $$\overline{L}_{entrained} = L_e = \frac{\psi}{1 - \psi} \times \overline{L}$$

    $$\overline{L}_e = \frac{0.15}{1 - 0.15} \times 346.582 = 61.16 kmol/hr$$

  - **Amount Entrained on Top:**

    $$\overline{L} + \overline{L}_{entrained} = \overline{L} + \overline{L}_e = 346.582 + 61.16 = 407.74 kmol/hr$$

  - **Column Cross Sectional Area:**

    $$A_{Total} = \frac{\pi \times Dia^2}{4} = \frac{\pi \times 0.8131^2}{4}$$

    $$A_{Total} = 0.51925 m^2$$

  - **Downcomer Area:**

    $$A_d = (1 - \eta) \times A_{Total}$$

    η denotes the fraction of tray area excluding the downcomer portion. Generally its value lies between 0.8 to 0.95. Assuming the value to be 0.9.

    $$A_d = (1 - 0.9) \times 0.51925 = 0.051925 m^2$$

– **Weir Length and Height**

For a distillation column operating above atmospheric pressure, weir height is keptin between 40 mm to 90 mm; while 40mm to 50 mm is more common.

Therefore, Weir height in the bottom section is also kept as $h_w = 45mm$.

Weir Length is normally kept in between 0.6 to 0.85 times the column diameter. For $\frac{A_d}{A_{Total}} = 0.1$, 0.727 can be used as the multiplying factor.

Therefore, Weir Length= $0.727 \times 0.8131 = 0.5911m = 23.2716inch = 1.9393ft$

– **Active Area of the Tray:**

$$\eta = 0.9$$

$$A_{Active} = A_{Total} \times (1 - 2(1 - \eta))$$

Therefore, $A_{Active} = 0.51925 \times (1 - 2(1 - 0.9)) = 0.4154m^2$

– **Total Area of holes:**

$$A_{Holes} = A_{Active} \times \beta$$

The tray selected for the bottom section is a standard 14 gauge tray with thickness (T_{tray}) = 0.078 in with a common hole diameter $d_o = \frac{3}{16}$ inch for normal operation and clean service. Pitch standard spacing between the holes is of $3.8d_o = 0.7125$ inches. A 2.5 inch space between the edge holes and the column wall is chosen, and a space of 4 inch between the edge hole and the tray weir. All these values are standard values used in the industry.

Generally $\beta = 0.1$

Therefore, $A_{Holes} = 0.1 \times 0.4154 = 0.04154m^2$

– **Vapor velocity through Tray Hole:**

$$v = \frac{\overline{V} \times M_{avg}}{3600 \times \rho_V \times A_{Holes}}$$

$$v = \frac{131.5256 \times 22.7239}{3600 \times 1.012 \times 0.04154}$$

$$v = 19.74m/s$$

– **Orifice Coefficient:**

It is used in the calculation of dry pressure drop through the tray. Orifice Coefficient C_o is found out using the $\frac{T_{Tray}}{HoleDiameter}$ and $\frac{A_{Holes}}{A_{Active}}$. Refer to Figure 22 which gives the correlation used to find C_o

Now,

$$\frac{A_{Holes}}{A_{Active}} = \frac{0.04154}{0.4154} = 0.1$$

$$\frac{T_{Tray}}{HoleDiameter} = \frac{0.078}{0.1875} = 0.416$$

Using Figure 22,

$$C_o = 0.73$$

Using a much accurate correlation given by Hughmark and O'Conell:

$$C_o = 0.85032 - 0.04231\frac{d_o}{T_{Tray}} + 0.00017954\frac{d_o}{T_{Tray}}^2$$

Calculating, $C_o = 0.759$

– **Total Head Loss of Liquid through dry Tray:**
  The following equation was given by Ludwig:

$$h_{\triangle P, dry} = \frac{0.003 \times v^2 \times \rho_V(\frac{\rho_{Water}}{\rho_L})(1 - \beta^2)}{C_o^2}$$

Now, $v = 19.74 m/s = 64.763 ft/sec$

$$\rho_V = 1.012 kg/m^3 = 0.063177 lb/ft^3$$

Therefore, on calculating

$$h_{\triangle P, dry} = 1.6229 inch = 0.04122m$$

– **Liquid Height over the weir:**
  Now, Liquid height over the weir=$h_{crest} = h_{ow} = 0.48F_w(Q_L)^{\frac{2}{3}}$
  where $F_w = Weir Correction Factor$

$$Q_L = \frac{\text{Liquid Load in gallons per minute}}{\text{Length of the Weir}}$$

For F_w, referring to Figure 23 :
Now Liquid Load in gallons per minute, L_g:

$$\overline{L}_g = \frac{(\overline{L} + \overline{L}_e)M_{avg}}{\rho_L}$$

$$\overline{L}_g = \frac{346.582 \times 22.7239}{841.75} = 9.3563 m^3/hr = 2471.672 gallons/hr = 41.19 gallons/minute = 0.09177$$

Abscissa= $\frac{\overline{L}_g}{WeirLength, ft^{2.5}}$

$$Abscissa = \frac{41.19}{1.9393^{2.5}} = 7.8646$$

Ratio weir length to tower diameter= $\frac{0.5911}{0.8131} = 0.727$
Using Figure 23, $F_w = 1.053$
Now, $h_{ow} = h_{crest} = 0.48F_w(Q_L)^{\frac{2}{3}}$

$$h_{crest} = 0.48 \times 1.053 \times (\frac{41.19}{23.2716})^{\frac{2}{3}}$$

$$h_{crest} = 0.7395 in = 0.018783m$$

- **Liquid Fractional Head Loss:**

  The flow area under the downcomer is calcualted as follows:

  $$A_{du} = gap \times Weir Length$$

  The gap is usually 1 inch=0.083 ft in all the distillation columns working above the atmospheric pressure.

  Therefore,

  $$A_{du} = 0.083 ft \times 1.9393 ft = 0.16096 ft^2$$

  Fractional loss in liquid head, $h_{du} = 0.56 \times \frac{\overline{L}_g, ft^3/sec}{A_{du}}^2$

  $$h_{du} = 0.56 \times \frac{0.09177^2}{0.16096} = 0.18203 inch = 0.0046235 m$$

- **Residence Time:**

  The total pressure head on the downcomer is the summation of all the hydrodynamic effects determined previously as follows:

  $$h_{dc} = h_{\triangle P,dry} + h_{weir} + h_{crest} + h_{grad} + h_{du}$$

  h_{grad} is generally zero for sieve tray tower.

  $$h_{dc} = 0.04122 + 0.045 + 0.018783 + 0.0046235 = 0.1096 m$$

  The residence time is calculated as follows:

  $$\theta_{res} = \frac{A_d h_{dc} \rho_L}{(\overline{L} + \overline{L}_e) M_{avg}}$$

  $$\theta_{res} = \frac{0.051925 \times 0.1096 \times 841.75}{407.74 \times 22.7239} = 5.17 \times 10^{-04} = 1.8609 sec$$

- **Flooding Check:**

  As calculated above, total pressure head on the downcomer, $h_{dc} = 0.1096 m$

  The actual aerated head, $h_{dc,aerated} = \frac{h_{dc}}{\phi}$

  where ϕ = froth density at normal operation

  Now,

  $$\phi = \frac{1}{\eta + 1}$$

  $$\eta = 0.9$$

  Therefore, $\phi = \frac{1}{1+0.9} = 0.52$

  $$h_{dc,aerated} = \frac{0.1096}{0.52} = 0.2107$$

  Since, $h_{dc,aerated} = 0.2107 m$ is less than tray spacing which is $12 inch \approx 0.3m$.

  The aerated liquid head is much less than the tray spacing which is 12 inch, there would be no operational problem and the liquid flooding is regulated.

– **Weeping Check:**

An analysis is done to check for the operation to be above the weeping and dumping points and avoid excessive weeping.Surface Tension head h_σ is calculated and follwing condition is checked for weeping:

Condition for weeping:

$$h_{\triangle P,dry} + h_\sigma \geq 0.10392 + 0.21599X - 0.021675X^2$$

where $X = h_{weir} + h_{crest} + h_{grad}$
Now, $\sigma = 0.03518N/m$

$$\rho_L = 841.75kg/m^3 = 52.548lb/ft^3$$

$$d_o = \frac{3}{16}inch = 0.0047625m$$

$$h_{\triangle P,dry} = 1.6229inch$$

$$h_{weir} = 45mm = 0.045m = 1.771in$$

$$h_{crest} = 0.7395inch$$

Calculation for X:

$$X = 1.771 + 0.7395 + 0 = 2.5105inch$$

Now, $0.10392 + 0.25199 \times 2.5105 - 0.021675 \times 2.5105^2 = 0.6011$
Now,

$$h_\sigma = \frac{0.04 \times \sigma}{\rho_L \times d_o} = \frac{0.04 \times 35.18}{52.548 \times \frac{3}{16}} = 0.1428inch$$

Therefore $h_{\triangle P,dry} + h_\sigma = 1.6229 + 0.1428 = 1.7657inch$
Condition:

$$h_{\triangle P,dry} + h_\sigma \geq 0.10392 + 0.21599X - 0.021675X^2$$

$$1.7657 \geq 0.6011$$

Condition is satisfied. Hence, no weeping or dumping occurs.

7.9 Design Schematics for Rectifying Section

Following figures depict the tray layout and design schematics for the tray in the top of the column. Similar sketches can be drawn for the stripping section as well.
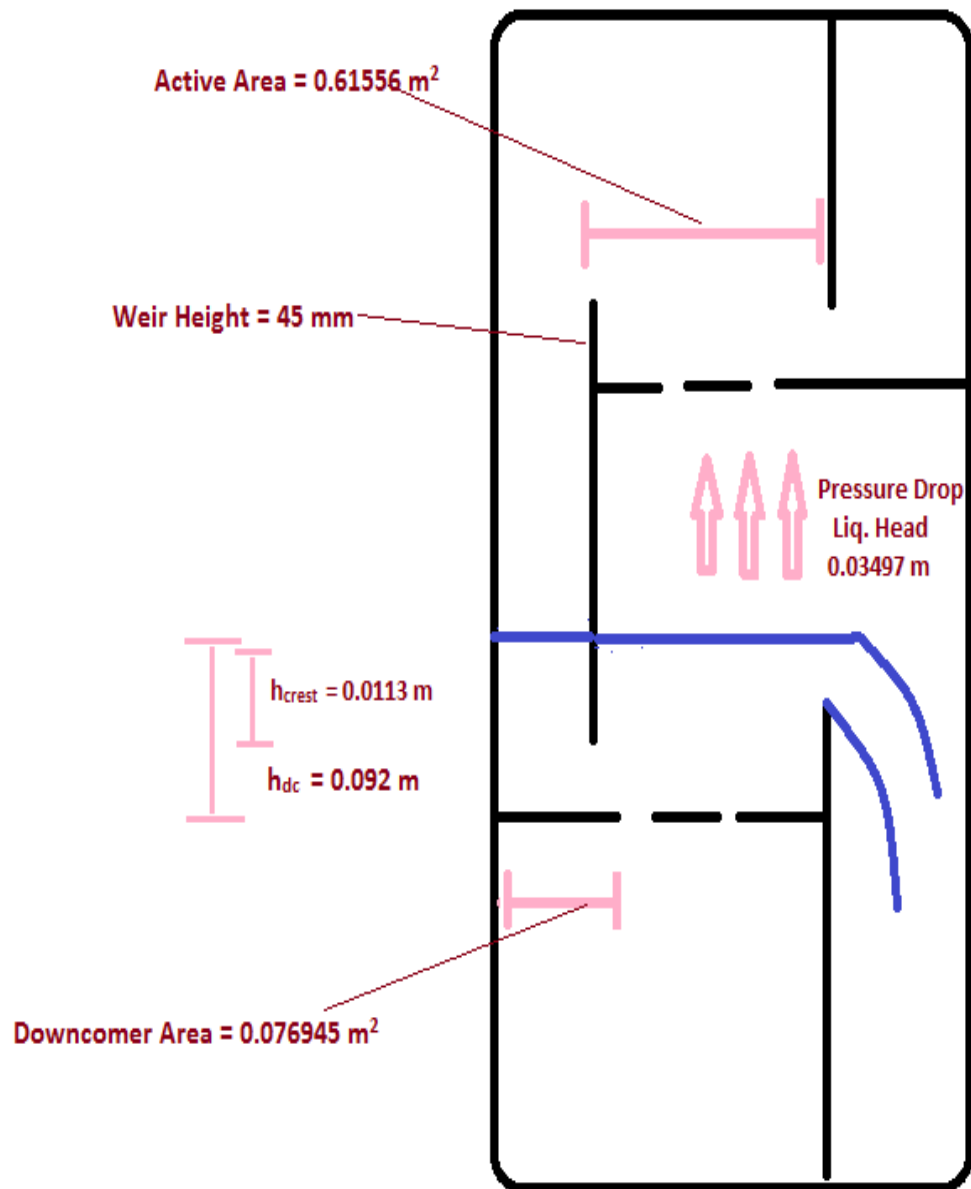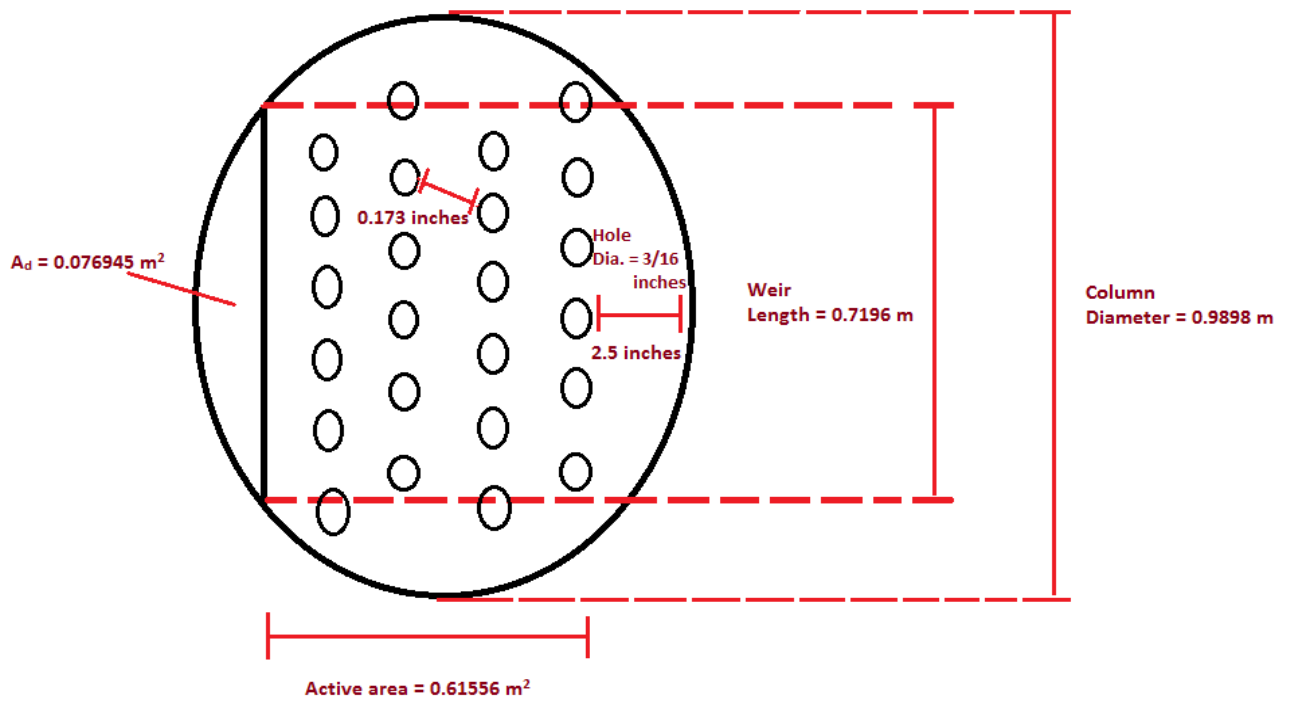

Figure 24: Column Design Schematics

Figure 25: Tray Design Schematics

8 Design of Auxilliary Equipment

Under this section, we will refer to the designing of the reboiler, condenser and the heat exchanger which is being used to cool down the formalin.

8.1 Reboiler Design

Duty of Reboiler (Refer to Energy Balance of Distillation Column)=1353.295 kW

Tube Side Fluid: Medium Pressure Steam ; Shell Side Fluid: Bottom Stream Coming out of distillation column

We have decided to use a Kettle typeof Reboiler because it is economically more compatible with our project. For a thermosyphon reboiler, we need to put a reboiler at a certain height so that gravitational force can work. To avoid such a situation, we have selected kettle reboiler. Also, the pressure drop in the shell side is negligible for a kettle type of reboiler.

Now, Medium pressure Steam is used to vaporize the bottom stream coming from distillation column.

Temperature of the Steam=180°C Pressure of the Steam= 45 psi=3.102 Bar

Steam Required= $\frac{\phi_b}{\lambda_s}$

From Steam Table:

$$\lambda_s = 2159.982 kJ/kg$$

Hence, Steam Required: $\frac{1353.295}{2159.982} = 0.6265 kg/s = 2255.51 kg/hr$

For the first trial Calculations, Assume $U_o = 1500 \frac{W}{m^{2}°C}$

For a reboiler, generally $\triangle T_m = 35°C$

Therefore, $A_o = \frac{\phi_b}{U_o \times \triangle T_m} = \frac{1353.295 \times 10^3}{1500 \times 35} = 25.77 m^2 = A_{o,provided}$

$$A_{o,provided} = 25.77 = N_t \pi d_o L$$

Nominal U-Tube length used for a reboiler= 12 ft=3.6576 m

Using 14 Birmingham Wire gauge (BWG) specifications for tube:

$$d_o = 25.4mm(1 inch); d_i = 21.1836mm(0.834 inch)$$

To calculate N_t :

$$A_{o,provided} = N_t \pi d_o L$$

$$N_t = 88.29 \approx 90(U - tubes)$$

Number of holes in tube sheet = $90 \times 2 = 180$

Use square pitch arrangement:

$$P_t = 1.5 d_o = 38.1mm$$

$$Maximum U - Bend Radius = 1.5 d_o = 38.1mm$$

- **Calculation of Boiling Heat Transfer Coefficient**

Now, $h_o = h_{nbo}$ = Nucleate Boiling Heat Transfer Coefficient for mixture = $f_m \times h_{nb}$

h_{nb} is calcultated using Mostinski's Equation.

$$h_{nb} = 0.104 \times (p_{cm})^{0.69}(\frac{\phi_b}{A})^{0.7}\left[1.8(\frac{P}{P_{cm}})^{0.17} + 4(\frac{P}{P_{cm}})^{1.2} + 10(\frac{P}{P_{cm}})^{10}\right]$$

$$\frac{\phi_b}{A} = \frac{1353.295 \times 10^3}{25.77} = 52514.35 W/m^2$$

P=1.3973 atm=1.4158 Bar

To calculate P_{cm}, use Kay's Law:

$$P_{cm} = \sum y_i P_{ci}$$

Table 19: Calculation of Critical Pressure for a mixture

Component	y_i	P_{ci} in Bar	$y_i \times P_{ci}$
Methanol	0.0001724	81	0.013932
Water	0.6060	220.584	133.673
Formaldehyde	0.39372	75.6	

Now, $P_{cm} = \sum y_i P_{ci} = 163.452 Bar$

Therefore, $\frac{P}{P_{cm}} = 8.661 \times 10^{-3}$

f_m=Correction Factor

$$f_m = exp(-0.0083(T_{bo} - T_{bi})$$

where T_{bo} = Temperature of the vapour leaving the reboiler = $110°C$

T_{bi} = Temperature of the liquid entering the reboiler = $108.7°ircCr$

Therefore, $f_m = 0.989$

Now, Substituting the value of P, P_{cm} in the expression for h_{nb}:

On calculation:
$$h_{nb} = 5785.578 W/m^{2°}C$$

Now, $h_b = f_m \times h_{nb}$

$$h_b = 5695.23 W/m^{2°}C$$

For condensation of steam, the generally taken value for $h_{i'}$. This term includes fouling also.

$$h_{i'} = 6000 W/m^{2°}C$$

Also, the general Shell Side fouling Coefficient, h_{od} =3000$W/m^{2°}C$

Now,

$$\frac{1}{U_o} = \frac{1}{h_b} + \frac{1}{h_{od}} + \frac{d_o \ln \frac{d_o}{d_i}}{2 \times k} + \frac{d_o}{d_i}\frac{1}{h_{i'}}$$

Substituting the values , U_o is found out to be $1335.41 W/m^{2\circ}C$.

Referring to literature, the value of U_o calculated lies within the normal range.

- **Calculation of Heat Transfer Area Required** Heat Transfer Area Required:

$$A_{Req} = \frac{\phi_b}{U_o \times \triangle T_m} = \frac{1353.295 \times 10^3}{1335.41 \times 35} = 28.95 m^2$$

To decrease the area required, we have to increase the pressure and temperature of the steam.

Therefore, taking the steam at $4 Bar and 200^{\circ}C$, $\lambda_s = 2132.95 kJ/kg$

Amount of steam required $= \frac{\phi_b}{\lambda_s} = \frac{1353.295}{2132.95} = 0.6347 kg/s = 2285.16 kg/hr$

Now, temperature drivng force for steam at 4 bar, $\triangle T_m = 40^{\circ}C$ Therefore, Area Required$= \frac{1353.295 \times 10^3}{1335.41 times 40} = 25.33 m^2$

Now, Percent Excess Heat Transfer Area$= \frac{25.77}{25.33} - 1 \times 100 = 1.73\%$

- **Calculation of Heat Flux**

Actual heat Flux$= \frac{\phi_b}{A_{o,provided}} = \frac{1353.295 \times 10^3}{25.77} = 52514.35 W/m^2$

Actual heat flux should be less than 0.7 times the critical heat flux to function properly.

Critical Heat Flux:

$$(\frac{\phi}{A})_c = K_b(\frac{P_t}{d_o})(\frac{\lambda}{\sqrt{N_t}})(\sigma(\rho_L - \rho_V)(\rho_v)^2)^{0.25}$$

Now, $K_b = Constant, 0.44 for square pitch arrangement$

$$P_t =$$

Tube Pitch, mm=38.1mm

$$d_o =$$

Tube Outside Diameter, mm=25.4 mm

$$N_t = 180$$
$$g = 9.81 m/S^2$$

Liquid Density=
$$\rho_L = 841.75 kg/m^3$$

Vapour Density$=\rho_V = 1.012 kg/m^3$

$$\sigma = 0.03518 N/m$$

Heat of vaporization Calculation:

Table 20: Heat of Vaporization calculation for Bottom Stream

Component	x_i	T_c in K	T_1 in K	$l\lambda$ at T_1
Methanol	0.0001724	512.6	338.15	35278
Formaldehyde	0.39372	410.150	253.7	23300
Water	0.6060	647.15	373.15	

Using Watson's Equation:

$$\lambda_{methanol} = 31556.90 kJ/kmol$$

$$\lambda_{formaldehyde} = 12059.72 kJ/kmol$$

$$\lambda_{water} = 40117.2 kJ/kmol$$

Now, $\lambda_{avg} = 29064.60 kJ/kmol = \frac{29064.60}{22.7239} = 1279032.208 J/kg$ Substituting the values for the calculation of critical heat flux:

Calculating $(\frac{\phi}{A})_c = 261237.669 W/m^2$

Now, 0.7 times the critical flux value= $182866.36 W/m^2$

Actual heat flux is less than 0.7 times the critical flux value.

Therefore, the reboiler design is acceptable as the actual flux value is very less than the critical flux value.

- **Calculation of Tube bundle Diamter**

$$n_b = \frac{P_t + \sqrt{P_t^2 + \pi N_t P_t^2}}{\frac{\pi}{2}} + d_o$$

Now, $N_t = 180, P_t = 38.1mm, d_o = 25.4mm$

On calculating:

$$n_b = 627mm$$

Now, if the actual heat flux is greater than 40000 W/m^2, the ratio of shell inner diameter to Tube bundle Diameter is taken to be around 1.7 to 2. Taking 1.8 as the ratio:

Shell Dimater= $1.8 \times n_b = 1.8 \times 627 = 1130mm$

- **Checking of Liquid Entrainment**

For no entrainment : Actual velocity should be less than Maximum velocity (v_{max})

Actual Velocity, v=$\frac{\dot{m}_v}{\rho_V \times L' \times x}$

Tube length, $L' = \frac{3.6576}{2} = 1.8288m$

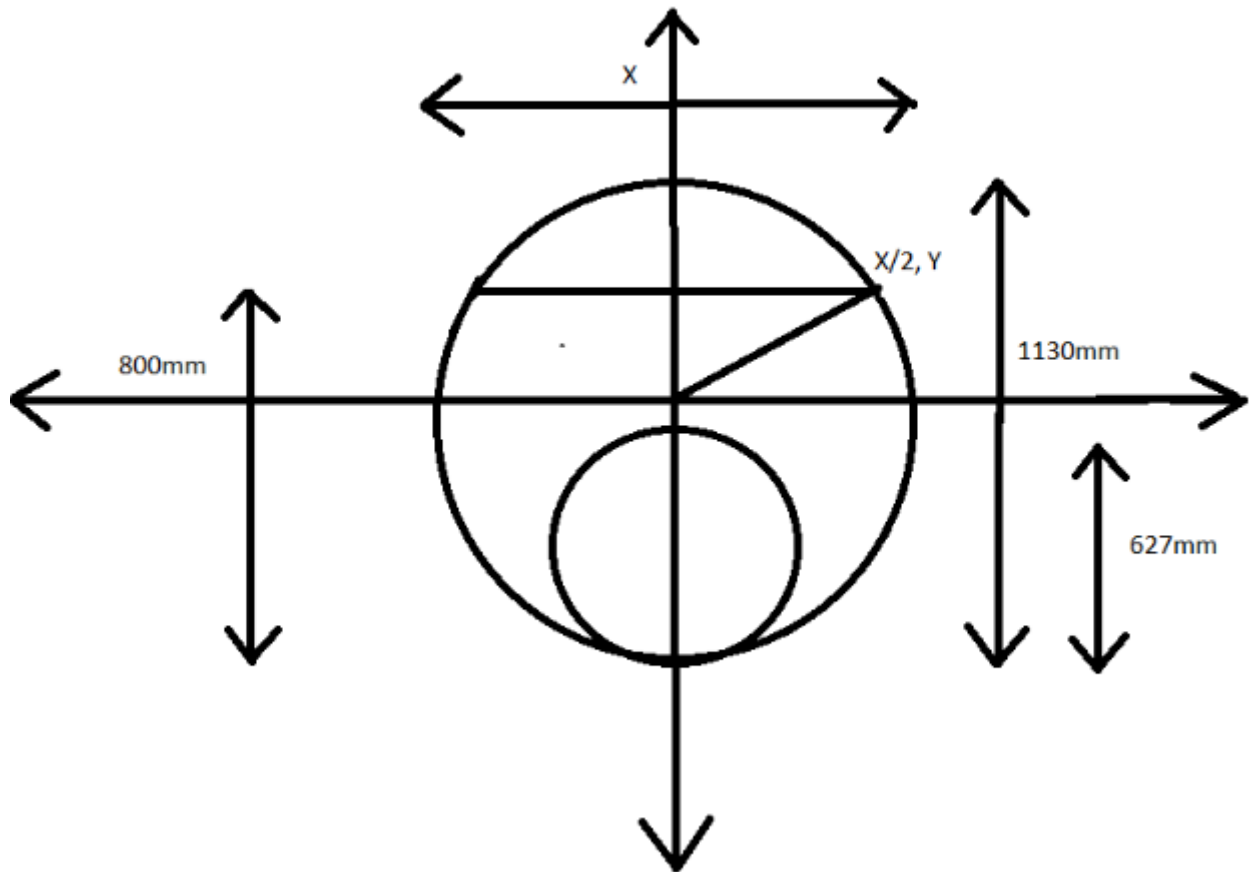Let the liquid level be 800 mm.

$$y = 800 - \frac{1130}{2} = 235mm$$

Figure 26: Reboiler Geometry

From the geometry of the figure:

$$(\frac{x}{2})^2 + 235^2 = (\frac{1130}{2})^2$$

On solving,

$$x = 1027.61mm = 1.0276m$$

Therefore, Liquid Surface Area$= L' \times x = 1.8793m^2$

$$v_{max} = 0.2 \times \sqrt{(\frac{\rho_L - \rho_V}{\rho_V})} = 5.764m/s$$

$$v = \frac{\overline{V} \times M_{avg}}{\rho_V \times \text{Liquid Surface Area}}$$

$$v = \frac{131.5256 \times 22.7239}{1.012 \times 1.8793} = 1571.50m/hr = 0.436m/s$$

Since v is less than v_{max} , no liquid entrainment occurs.

- **Shell Side Pressure Drop**

  Shell Side pressure drop is negligible in case of kettle reboiler.

- **Tube Side pressure Drop**

  Tube side pressure drop is calculated using the equation:

  $$\triangle P_t = 0.5 \triangle P_t' = 0.5 N_p (8 J_f \frac{L}{d_i} (\frac{\mu}{\mu_w})^{-m} + 2.5) \frac{\rho u_t^2}{2}$$

  $$N_p = 2; L = 1.8288m; d_i = 0.0211836m$$

At 4 Bar, $\rho = \frac{1}{\text{Specific Volume}} = \frac{1}{0.462} = 2.164 kg/m^3$

Tube Side Mass velocity,

$$G_t = \frac{\dot{m}_s}{\frac{N_t}{N_p} \times \frac{\pi}{4} d_i^2}$$

Substitute the values in G_t, we get$20.009 kg/m^2.s$

Tube Side Reynold's Number:

$$\frac{d_i \times G_t}{\mu}$$

$\mu_{water} = 2.5091 \times 10^{-4} kg/m.s; \mu_{methanol} = 0.00021412 kg/m.s; \mu_{formaldehyde} = 0.00258 kg/m.s$

Therefore $\mu_{mix} = 1.167 \times 10^{-4} kg/m.s$

Therefore, Reynold's Number=364.400

Referring to Figure 27, we get the value of $j_f = 0.024$

Now, $\mu_{vapour} = 1.608 \times 10^{-5} Pa.s$

$$u_t = \frac{G_t}{\rho} = \frac{20.009}{2.164} = 9.246 m/s$$

Substituting the corresponding values to get $\triangle P$, we get

$$\triangle P = 360.13 Pa = 0.360 kPa$$

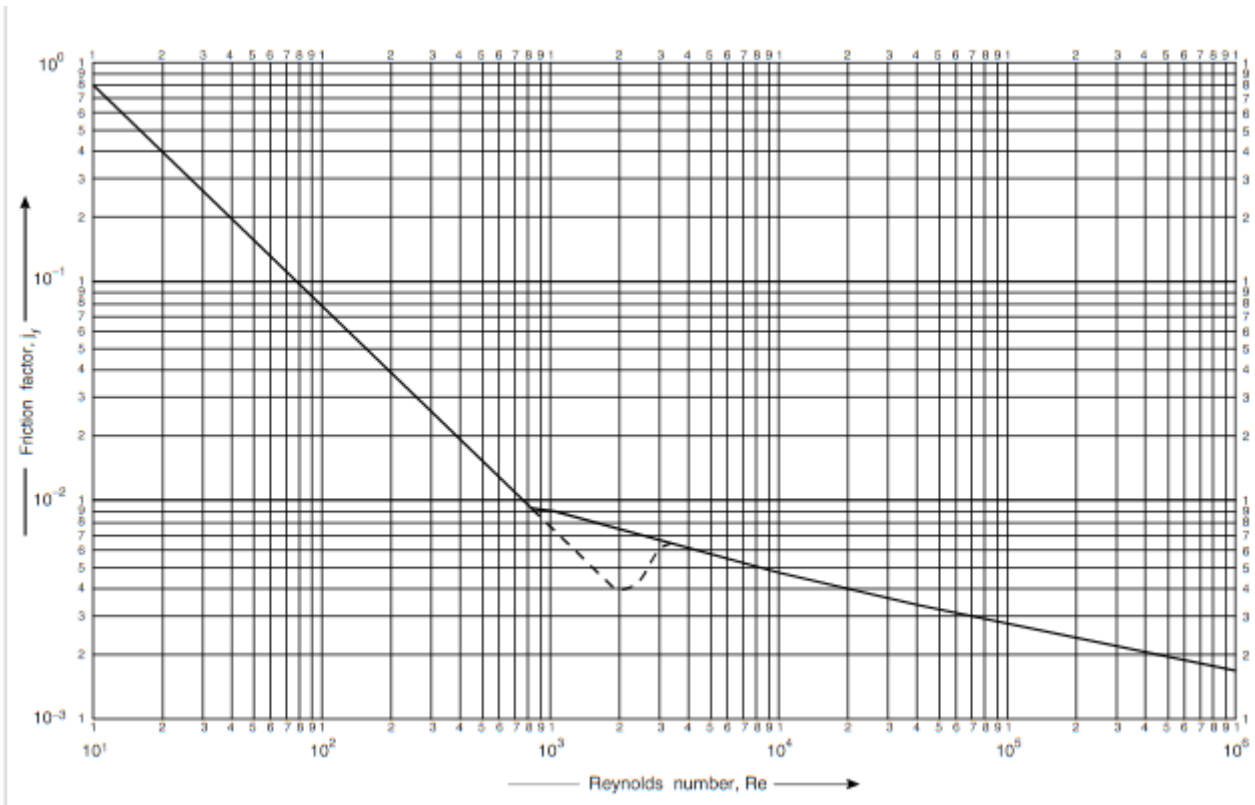This pressure drop value is acceptable.

Figure 27: Tube Side Friction Factor [8]

8.2   Condenser Design

Heat Duty of the Condenser (Refer to the Energy Balance Calculation)=1296.331 kW
Water at $32°C$ to be reduced to $40°C$.

$$\dot{m}_w C_{pw}(t_0 - 32) = 1296.331 kW$$

$$C_{pw} = 4.18 kJ/kg.K$$

$$\dot{m}_w = \frac{1296.331}{4.188 \times 8} = 38.6 kg/s$$

Referring to the data listing the general overall heat transfer coefficients, we take $U_o = 800 W/m^{2°}C$
LMTD for condensation= $\frac{30.94-38.94}{\ln(30.94/38.94)} = 34.78°C$

- **Calculation of Condenser Area**

  Area based on assumed value of U_o:

$$A_c = \frac{\text{Heat Duty}}{U_o \times LMTD} = \frac{1296.331 \times 1000}{800 \times 3478} = 46.59 m^2$$

Its a horizontal condenser. Therefore

$$A = 46.59 = N_t \pi d_o L$$

Outer Diameter, $d_o = 19.05 mm = \frac{3}{4} inch$

79

Length, L=6 ft= 1.83 m

Therefore, $N_t = \frac{46.59}{\pi \times 0.019505 \times 1.83} = 425.7 \approx 426 Tubes$

We selected triangular pitch (Clean Shell Fluid, hence, easier cleaning).

$$k_1 = 0.249$$
$$n_1 = 2.207$$

Therefore, $P_t = 1.25d_o = 23.8125mm$

Number of Tube Passes=2 (first trial calculation)

- **Calculation of Tube Bundle Diameter**
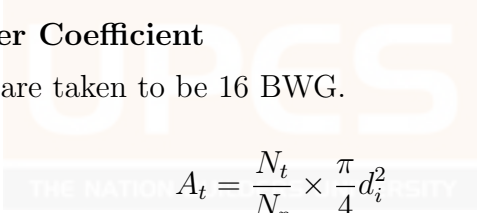
$$D_b = d_o(\frac{N_t}{K_1})^{(\frac{1}{n_1})}$$

$$D_b = 19.05(\frac{426}{0.249})^{(\frac{1}{2.207})} = 556.327mm$$

Clearance between shell and tube Bundle=13 mm

Therefore, Shell Inside Diameter= 570mm

- **Tube Side Heat Transfer Coefficient**

Inner and outer Diameter are taken to be 16 BWG.

$$A_t = \frac{N_t}{N_p} \times \frac{\pi}{4}d_i^2$$

$$A_t = \frac{426}{2} \times \frac{\pi}{4} \times (0.015748)^2 = 0.04148m^2$$

Tube Mass velocity=
$$G_f = \frac{\dot{m}}{A_t} = \frac{38.6}{0.04148} = 930.39kg/m^2.s$$

$$u_t = \frac{G_t}{\rho} = \frac{930.39}{992.6} = 0.937m/s$$

which is less than 1 m/s.

To avoid fouling velocity should be greater than 1 m/s.

Therefore, to increase the number of tube side passes from 2 to 4.

Now, $N_p = 4, A_t = 0.02074m^2$

Therefore new bundle diameter, $d_b = 19.05 \times (\frac{426}{0.175})^{(\frac{1}{2.285})} = 577.9 \approx 578mm$

Clearance=15 mm

Shell Inside Diameter=578+15=593 mm (revised)

$$G_t = \frac{38.6}{0.02074} = 1861.13kg/m^2.s$$

$$u_t = \frac{1861.13}{992.9} = 1.875 m/s$$

Reynolds Number= $\frac{d_i \times G_t}{\mu}$

$$\mu = 0.72 \times 10^{-3} kg/m.s$$

$$Re = \frac{0.015748 \times 1861.13}{0.72 \times 10^{(-3)}} = 40707.04$$

Prandtl Number:

Thermal Conductivity of water, $k = 0.6228 W/m.°C$

$$Pr = \frac{C_p \times \mu}{k} = \frac{4.1868 \times 0.72 \times 10^{-3} \times 10^3}{0.6228} = 4.84$$

Using Dittus-Boelter Equation,

$$\frac{h_i d_i}{k} = 0.023 Re^0.8 Pr^0.3 (\frac{\mu}{\mu_w})^0.14$$

Neglecting, $\frac{\mu}{\mu_w}$, since it will be close to unity.

Calculating for h_i, we get $7457.36 W/m.°C$

- **Shell Side Condensation Coefficient**

For condensation, calculation of mean temperature of condensate film: Let t_w=tube wall temperature,°C

Assumed for first trial calculation:

$$h_{co} = 1500 W/m.°C$$

Heat Transfer rate through condensate film=Overall Rate of heat transfer

$$h_{co} A_c (t_c - t_w) = U_o A_c (t_c - t_{avg})$$

where, t_c=Condensation temperature of vapour on shell in ° C

t_{avg} = average tube side fluid temperature for condensation zone in°C

$$1500 \times 46.59 \times (70.94 - t_w) = 800 \times 46.59 \times (70.94 - \frac{32+40}{2})$$

Solving, $t_w = 52.305°C$

Now, Mean temperature for condensate film:

$$\frac{t_c + t_w}{2} = \frac{70.94 + 52.305}{2} = 61.6225°C$$

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Properties of Methanol- water at $61.6225°C$:

$$\mu_L = 0.308459 mPa$$
$$\rho_L = 751.986 kg.m^3$$
$$k_L = 0.203 W/m.°C$$

Shell Side Condensation Coefficient:

$$h_{co} = 0.95 k_L \left(\frac{\rho_L(\rho_L - \rho_V)g}{\mu_L \tau_h}\right)^{\left(\frac{1}{3}\right)} N_r^{\left(\frac{-1}{6}\right)}$$

Now,

$$\tau_h = \frac{W_c}{L N_t} = \frac{131.5256 \times 30.644}{3600 \times 1.83 \times 426} = 1.43612 \times 10^{-3} kg/m.s$$

Average Number of Tubes in Vertical Rows:

$$N_r = \frac{2}{3} N_r' = \frac{2}{3} \times \frac{D_o}{P_t} = \frac{2}{3} \times \frac{557}{23.8125} = 15.59 \approx 15.6$$

Substituting all the parameters in the h_{co} equation, on calculating:

$$h_{co} = 2831.1 W/m^2.°C$$

The calculated value is near to assumed value.

Therefore, we can use this value in our subsequent calculations.

Overall heat transfer coefficient:

$$\frac{1}{U_o} = \frac{1}{h_{oc}} + \frac{1}{h_{od}} + \frac{d_o \ln \frac{d_o}{d_i}}{2 \times k} + \frac{d_o}{d_i}\frac{1}{h_i} + \frac{d_o}{d_i}\frac{1}{h_{id}}$$

Taking the values of $h_{od} = 10000 W/m^2.°C$ and $h_{id} = 4000 W/m^2.°C$

Substituting all the values, we get $U_o = 976.08 W/m^2.°C$.

Area required heat transfer= $\frac{1296.331 \times 1000}{976.08 \times 34.78} = 38.175 m^2$

Percent Excess Heat transfer area should be atleast 15 %.

Therefore, we can increase tube length.

$$\frac{Area_{before}}{Area_{before}} = \frac{46.59}{38.175} = 1.2204$$

$$\frac{L_{new}}{L_{old}} = 1.15 \times \frac{A_{new}}{A_{before}} = 1.724 m$$

Now, L is kept 1.83 m.

- **Shell Side Pressure Drop**

Shell Side flow area:

Inner shell diameter $=0.593$m

$$A_s = \frac{(P_t - d_o) \times B_s \times D_s}{P_t} = \frac{(23.8125 - 19.05) \times 0.593 \times 0.593}{23.8125} = 0.0703298 m^2$$

Shell Side Velocity:

$$G_s = \frac{\dot{m}}{A_s} = \frac{131.5256 \times 30.644}{3600 \times 0.0703298} = 15.9189 kg/m^2.s$$

$$u_s = \frac{G_s}{\rho_V} = \frac{15.9189}{1.47} = 10.829 m/s$$

Equivalent Diameter of Triangular Pitch Arrangement:

$$d_e = \frac{1.1}{d_o} \times (P_t^2 - 0.907 d_o^2)$$

$$d_e = \frac{1.1}{19.05} \times (23.8125^2 - 0.907 \times 19.05^2)$$

$$d_e = 13.73 mm$$

$$Reynolds Number = \frac{d_e \times G_s}{\mu}$$

$$\mu = 1.1 \times 10^{-2} cP$$

$$Re = \frac{13.73 \times 10^{-3} \times 15.9189}{1.1 \times 10^{-5}} = 19869.68$$

Referring to Figure 28, for 25% cut segmental baffle

$$j_f = 0.042$$

$$\triangle P_s = 8 j_f \times \frac{D_s}{d_e} \frac{L}{B_s} \frac{\rho_V u_s^2}{2} (\frac{\mu}{\mu_w})^{-0.14}$$

Neglecting $\frac{\mu}{\mu_w}$, as its value is close to unity.

Calculating,

$$\triangle P_s = 8 \times 0.042 \times \frac{593}{13.736} \times \frac{1.83}{0.593} \times \frac{10.829^2 \times 1.47}{2}$$
$$\triangle P_s = 918.638 Pa = 0.0918 kPa$$

This is an acceptable value for shell side pressure drop.
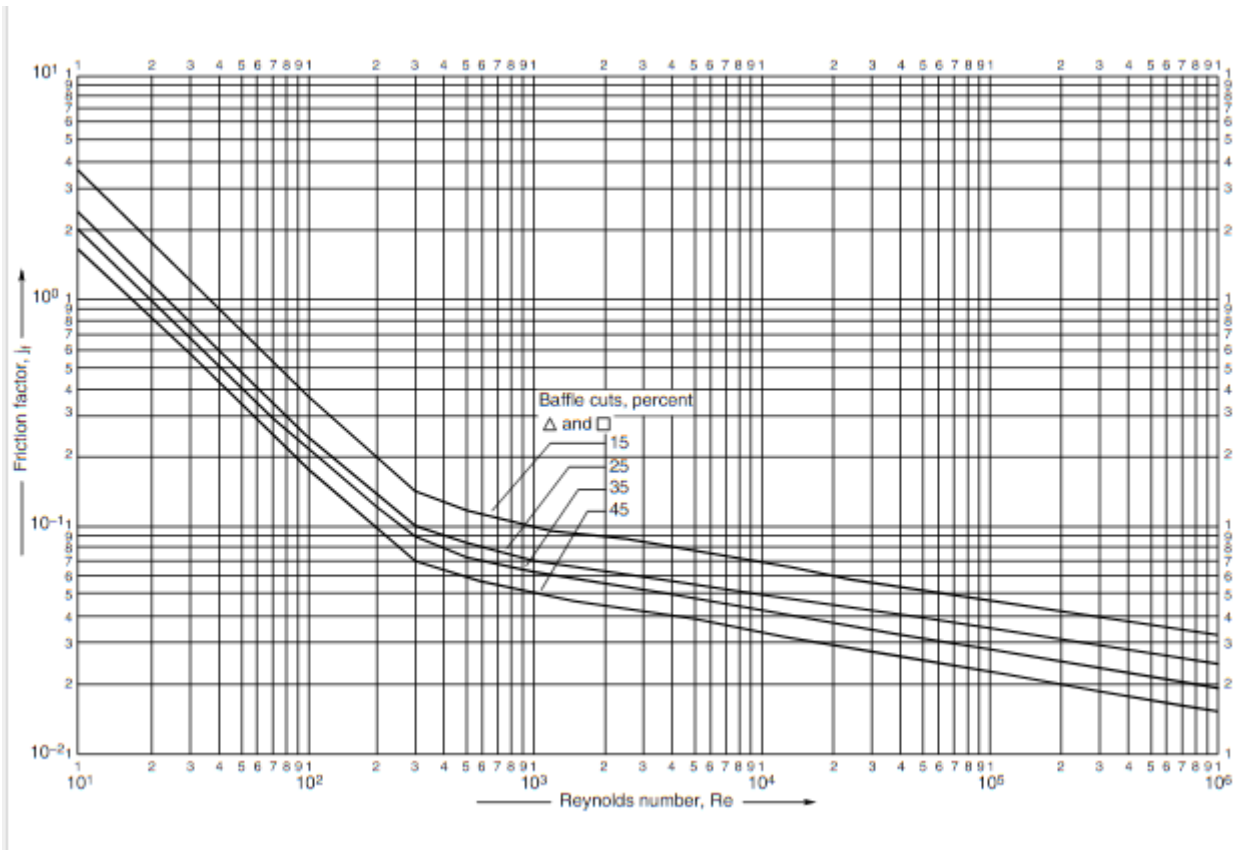
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Figure 28: Shell-Side friction factors, segmental baffles [8]

- **Tube Side pressure Drop**

  Tube side pressure drop is calculated using the following equation:

$$\triangle P_t = N_p(8J_f\frac{L}{d_i}(\frac{\mu}{\mu_w})^{-m} + 2.5)\frac{\rho u_t^2}{2}$$

$$N_p = 4$$
$$Re = 40707.4$$

Neglecting $\frac{\mu}{\mu_w}$, as its value is close to unity. Referring to Figure 27:

$$J_f = 3.5 \times 10^-3$$

$$L = 1.83m$$

$$d_i = 15.748mm$$
$$\rho_{water} = 993.684kg/m^3$$
$$u_t = 1.875m/s$$

Substituting the values in the $\triangle P_t$ expression, we get

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$$\triangle P_t = 40200.5 Pa = 40.2 kPa$$

This pressure drop is acceptable.

- **Condenser Specifications:**

  - 1-4 fixed tube shell Heat Exchanger
  - Shell Inner Diameter=593 mm
  - Baffles=25 % cut segmental
  - Baffle Spacing=593 mm
  - Tube O.D=19.05 mm-16 BWG
  - Tube Length=1.83 m=6ft
  - Triangular Pitch Arrangement, Pitch= $1.25 \times d_o = 23.8125 mm$
  - Number of Tubes: 426

8.3 Heat Exchanger Design

We will be designing the heat exchanger which is being used to cool the temperature of the formaldehyde dissolved in water stream after the distillation column.

Fluid Allocation: Shell=Cooling Water; Tube=Formalin

Duty of the heat exchanger: 905400 kJ/hr=251.1 kW

We are taking the cooling water at 15 °C and leaves the heat exchanger at 44 °C . The process fluid enters at 48 °C and leaves at 30 °C.

Stream 19

T=48°C

$N_{Methanol}$ = 0.037174 kmol/hr

$N_{Formaldehyde}$ =84.64511kmol/hr

N_{water}=239.5409 kmol/hr

N_{Total}=324.1244 kmol/hr

Cold Utility, Water
(15°C)

Stream 20

T=30°C

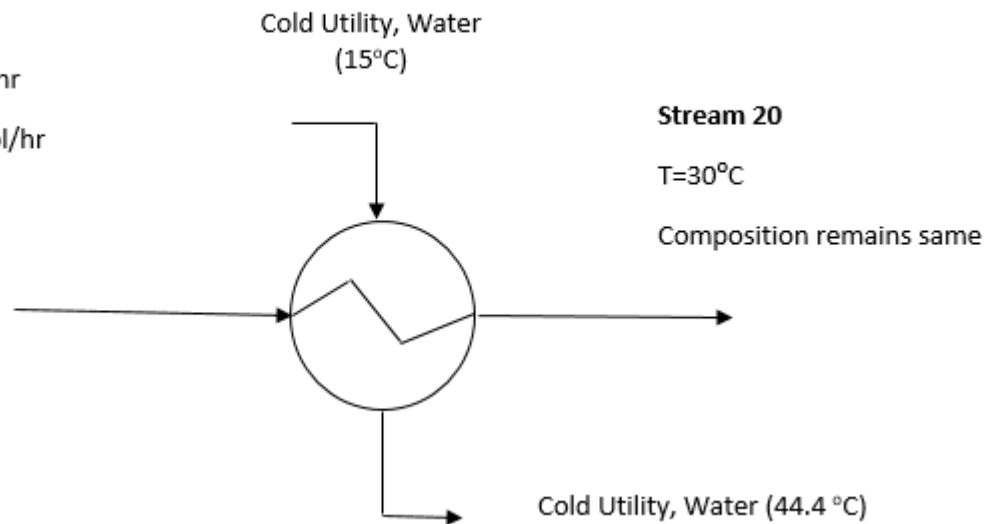Composition remains same

Cold Utility, Water (44.4 °C)

Figure 29: Heat Exchanger cooling the Formalin Stream from 48°C to 30 °C

Calculating LMTD:

$$LMTD = \frac{(48 - 44.4) - (30 - 15)}{\ln \frac{(48-44.4)}{(30-15)}} = 8.1°C$$

Correction Factor:

$$R = \frac{T_1 - T_2}{t_2 - t_1} = \frac{15 - 44.4}{30 - 48} = 1.633$$

$$S = \frac{t_2 - t_1}{T_1 - t_1} = \frac{30 - 48}{15 - 48} = 0.5454$$

Referring to Figure 30 , F=0.98

Therefore corrected LMTD= $\triangle T = 0.9 \times 8.1 = 7.84$

- **Calcualtion of Mass flow rate of water required**

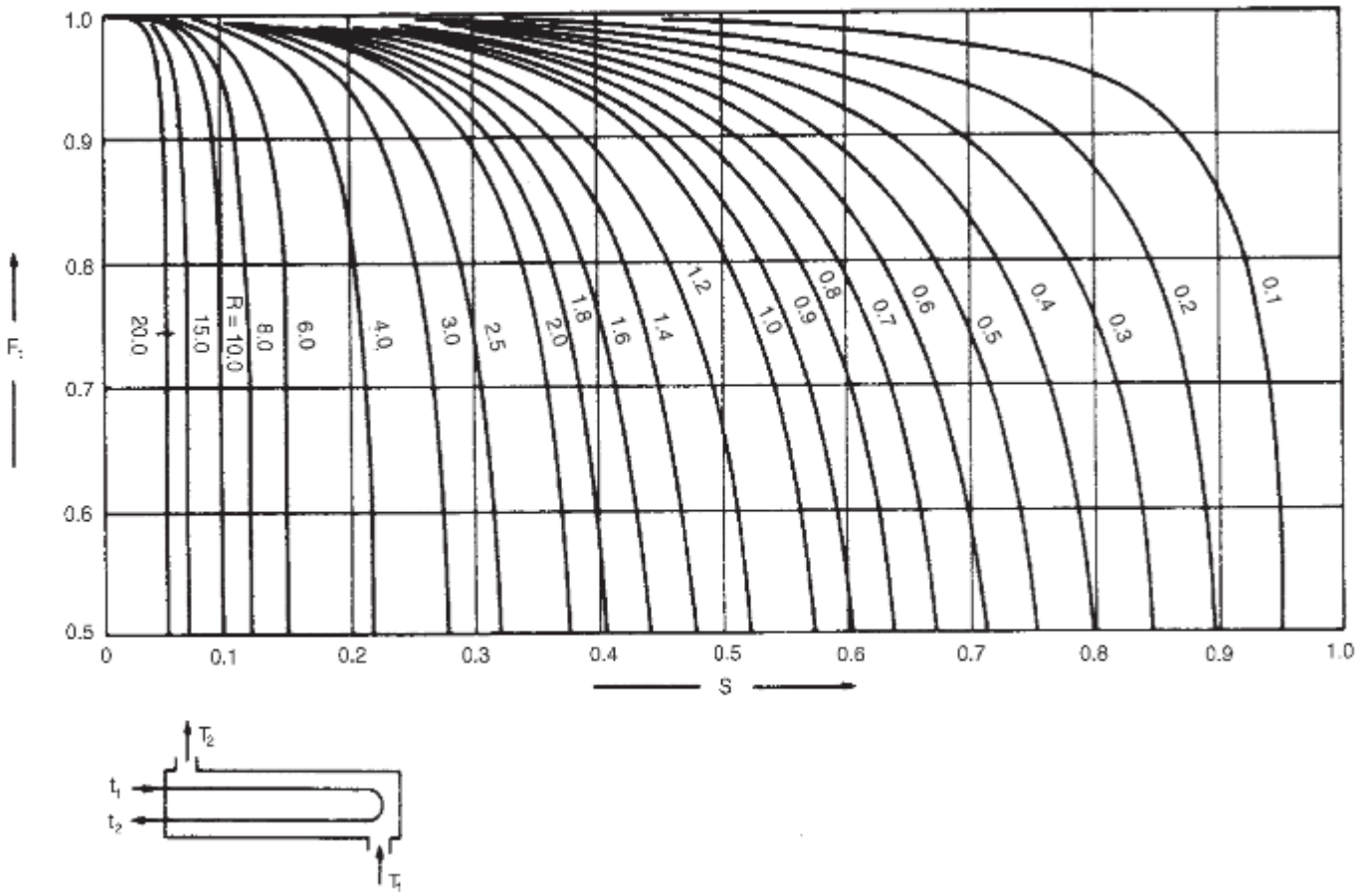 We know that

$$Q = \dot{m}C_p \triangle T$$

$$Q = 251.1 kW$$

Figure 30: Temperature Correction Factor-one shell pass,two or more even tube passes [12]

Water: $C_p = 4.18 kJ/kg.°C$

Therefore, Substituting all the values in the equation above:

$$\dot{m} = 6.012 kg/s$$

- **Calculation of Heat Transfer Area Required**

  Assuming overall heat transfer coefficient to be 800 $W/m^{2°}C$

$$A = \frac{Q}{U \times \triangle T}$$

$$A = \frac{251.1 \times 1000}{800 \times 7.84} = 43.59 m^2 \approx 44 m^2$$

Now, taking the Outer Diameterd_o=20 mm

Inner Diameterd_i=16mm

L=1.83m=12 ft

Area of one tube= $\pi \times d_o \times L = 0.114 m^2$

Number of tubes=$\frac{44}{0.114} = 385.96 \approx 386$ Tubes

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- **Calcualtion of Tube Bundle Diameter**

  Taking triangular pitch,
  $$n_1 = 2.207, k_1 = 0.249$$

  Therefore, $D_b = d_o(\frac{N_t}{k_1})^{(\frac{1}{n_1})}$

  On soving, $D_b = 557.96mm$

  Shell Diameter, taking a clearance of 40 mm= $557.96 + 40 = 597.96mm \approx 600mm$

- **Tube Side Coefficient**

  Mean Formalin Temperature= $\frac{48+20}{2} = 39°C$

  Tube Cross Sectional Area= $\frac{\pi}{4} \times d_i^2 = \frac{\pi}{4} \times 16^2 = 201mm^2$

  Tube per pass= $\frac{386}{2} = 193$

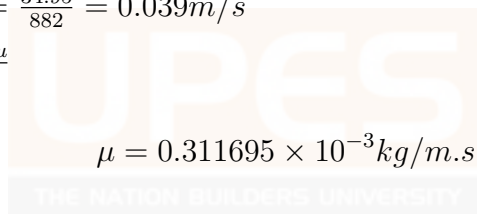 Total Flow Area=Tube per pass $\times$ Tube cross sectional area

  $$= 193 \times 201 \times 10^{-6} = 0.03879m^2$$

  Formalin Mass velocity= $\frac{\text{Mass Flow Rate}}{\text{Flow Area}} = \frac{4881.8}{0.03879} \times 3600 = 34.95kg/sm^2$

  Density= $\frac{1}{\sum \frac{w_i}{\rho_i}} = \frac{1}{\frac{0.5195}{798} + \frac{0.48}{995}} = 882kg/m^3$

  Formalin Linear Velocity= $\frac{34.95}{882} = 0.039m/s$

  Reynolds Number= $\frac{d_i \times \rho \times u}{\mu}$

  $$\mu = 0.311695 \times 10^{-3}kg/m.s$$

  $$Re = \frac{882 \times 0.039 \times 16 \times 10-3}{0.311695 \times 10^{(-3)}} = 1765.72$$

  Prandtl Number:

  Thermal Conductivity of water, $k = 0.609W/m.°C$

  $$Pr = \frac{C_p \times \mu}{k} = \frac{5.2 \times 1000 \times 0.311695 \times 10^{-3}}{0.609} = 2.66$$

  Using Dittus-Boelter Equation,

  $$\frac{h_i d_i}{k} = j_h Re Pr^0.3(\frac{\mu}{\mu_w})^{0.14}$$

  Neglecting, $\frac{\mu}{\mu_w}$, since it will be close to unity.

  Referring to Figure 31,

  $$\frac{L}{d_i} = \frac{1.83 \times 1000}{16} = 114.3$$
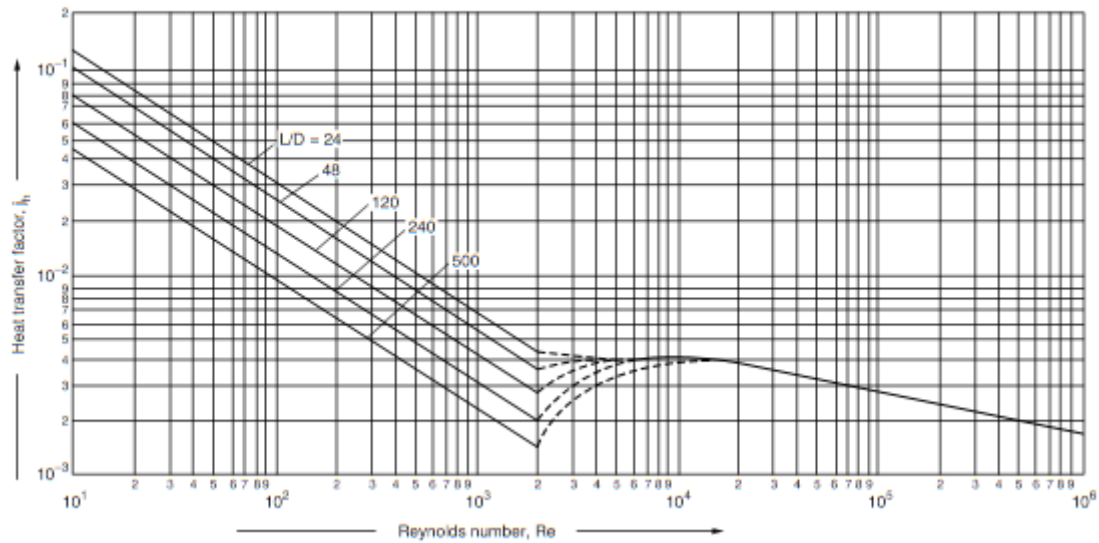
  $$j_h = 3 \times 10^{-3}$$

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Figure 31: Tube Side Heat Transfer Factor [12]

Substituting in the expression above for h_i, we get:

$$h_i = 278.4 W/m.°C$$

- **Shell Side Coefficient**

  Baffle Spacing $B_s = \frac{D_s}{3} = \frac{0.600}{3} = 0.2m$

  Tube Pitch $= 1.25 \times d_o = 25mm$

  Cross Flow Rate, $A_s = \frac{d_o - d_i}{d_i} \times D_s B_s$

  $$A_s = \frac{25 - 20}{20} \times 0.6 \times 02 = 0.03m^2$$

  Mass Velocity $G_s = \frac{6.012}{0.03} = 200 kg/sm^2$

  Equivalent Diameter of Triangular Pitch Arrangement:

  $$d_e = \frac{1.1}{d_o} \times (P_t^2 - 0.907 d_o^2)$$

  $$d_e = \frac{1.1}{20} \times (25^2 - 0.907 \times 20^2)$$

  $$d_e = 14.4mm$$

  Mean Shell-Side temperature $= \frac{15 + 44.4}{2} = 29.7°C$ Water Density $= 992 kg/m^3$

  Therefore $u = \frac{200}{992} = 0.189 m/s$ Viscosity $= 5.3 \times 10^{-4} kg/ms$

  Heat Capacity $= 4.18$ kJ/kg.K

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Thermal Conductivity$= 0.609 W/m°C$

$$Re = \frac{\rho v d}{\mu} = \frac{992 \times 0.189 \times 14.4 \times 10^{-3}}{5.3 \times 10^{-4}} = 5433.9$$

$$Pr = \frac{C_p \mu}{k} = \frac{4.18 \times 1000 \times 5.3 \times 10^{-4}}{0.609} = 3.36377$$

For 25 % Segmental Baffle Cut, referring to Figure 28,

$$j_h = 8 \times 10{-3}$$

$$\frac{h_i d_i}{k} = j_h Re Pr^0.3 (\frac{\mu}{\mu_w})^{0.14}$$

Neglecting, $\frac{\mu}{\mu_w}$, since it will be close to unity.

Solving for $h_i = 2813.36 W/m^{2°}C$

Overall heat transfer coefficient:

$$\frac{1}{U_o} = \frac{1}{h_o} + + \frac{d_o \ln \frac{d_o}{d_i}}{2 \times k} + \frac{d_o}{d_i} \frac{1}{h_i}$$

Now $h_o = 2813.3 W/m^{2°}C$

$$h_i = 278.4 W/m^{2°}C$$

Solving fro U_o, we get $204.72 W/m^{2°}C$

- **Tube Side Pressure Drop**

  Tube side pressure drop is calculated using the following equation:

  $$\triangle P_t = N_p (8 J_f \frac{L}{d_i} (\frac{\mu}{\mu_w})^{-m} + 2.5) \frac{\rho u_t^2}{2}$$

  $$N_p = 2$$
  $$Re = 1765.7$$

Neglecting $\frac{\mu}{\mu_w}$, as its value is close to unity. Referring to Figure 27:

$$J_f = 7 \times 10^-3$$

$$L = 1.83 m$$

$$d_i = 16 mm$$
$$\rho_{water} = 882 kg/m^3$$

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$$u_t = 0.039 m/s$$

Substituting the values in the $\triangle P_t$ expression, we get

$$\triangle P_t = 11940 Pa = 11.94 kPa$$

This pressure drop is acceptable.

- **Shell Side Pressure Drop**

  Referring to Figure 28, for 25% cut segmental baffle

  $$j_f = 0.008$$

  $$\triangle P_s = 8 j_f \times \frac{D_s}{l_b} \frac{\rho_V u_s^2}{2} \left(\frac{\mu}{\mu_w}\right)^{-0.14}$$

  Neglecting $\frac{\mu}{\mu_w}$, as its value is close to unity.
  Calculating,

  $$\triangle P_s = 8 \times 0.008 \times \frac{0.6}{0.2} \times \frac{0.2^2 \times 992}{2}$$
  $$\triangle P_s = 3800 Pa = 3.80 kPa$$

  This is an acceptable value for shell side pressure drop.

- **Heat Exchanger Specifications**

  Load=251.5 kW; 1-2 Fixed Tube Shell Heat Exchanger
  Tube Side Specifications

  – Number of Tubes=386
  – tube Pitch =25 mm
  – Bundle Diameter=0.557m
  – Bundle Diameter Clearance=0.040m
  – Number of tubes per pass=193
  – Tube Side heat transfer coefficient = $278.4 W/m^{2\circ}C$
  – Tube Side Pressure Drop=11.94 kPa

  Shell Side Specifications

  – Shell Diameter=0.6m
  – Baffle Spacing=0.2 m
  – Cross Flow Area=0.03 m^2
  – Shell Equivalent Diameter=0.0144m
  – Shell Side heat transfer coefficient=2813.36 $W/m^{2\circ}C$
  – Shell Side Pressure Drop=3.8 kPa

9 Cost Estimation

We have estimated the cost of formalin produced per ton, operating labour cost and general manufacturing expenses. Once a plant is designed, analysing its economics is a very crucial step[18]. The plant should be able to return all the capital invested and also generate profit after the breakeven point. Order-of-the-Magnitude Study Estimate is done. Thus method gives an error value of around ±30%.

Assumptions:

- Average Total working period of single operator is 49 weeks/year.

- 3 weeks of Holidays, Sick leave, Emergency Leave and Casual Leave are given.

9.1 Operating Cost labour

Generally an operator works 5 shifts in a week.

Therefore in a year = $49 \times 5 = 245$ shifts/year

The plant is operated all year (3 eight hour shifts 365 days). Therefore, Total shifts per year = 1095 shifts/year.

Number of Operators required to fill 1095 shifts= $\frac{1095}{245} = 4.5$ operators

Now, the number of non-particulate steps in the formaldehyde plant:

$N_{np} = \sum(equipment)$

N_{np} =1 reactor+ 2 Towers +1 compressor+6 exchangers=10

Therefore, Number of operators per shift N_{OL}

$$N_{OL} = 6.29 + 3.17P^2 + 0.23N_{np}$$

Generally, P is zero, as it refers to the handling of particulate solids, but our plant is a non-particulate one.

On solving, $N_{OL} = 2.9308$

Therefore, Operating Labour= Number of operators required to fill 1095 shifts$\times N_{OL} = 4.5 \times 2.9308 = 13.188 \approx 14$ operators

Assuming, for single operator, wage is Rs. 797 /hr.

Hence yearly payment for single operator:

$$797 Rs./hr \times 8hr/shift \times 245 shifts/year = Rs.156229/year$$

Total Operating Labour Cost= $14 \times 156229 = Rs.21871216/year$

From the preliminary cost analysis, we found out that for Stainless Steel as a material of construction, our Fixed Capital Investment (FCI) will be Rs. 1012510200.

Raw Materials Cost (C_{RM})=Rs. 418587906 /year Waste Treatment Cost(C_{WT}) = Rs. 0 Utilities(C_{UT})=Rs. 19324186 /year

9.2 Cost of Manufacturing

Cost of Manufacturing tells us about the unit cost price of the product formed. It helps us in setting the market price so that enough profit can be made and a return higher than the minimum acceptable return is achieved.

To calculate C_{OM}:

$$C_{OM} = 0.280FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + CRM)$$

Substituting these values, we get $C_{OM} = Rs.881843152.7/year$
Now, Cost of Manufacturing without depreciation;

$$C_{OM,d} = 0.18FCI + 2.72C_{OL} + 1.23(C_{UT} + C_{WT} + CRM)$$

Substituting the values, we get $C_{OM,d} = Rs.780592185/year$
Capacity of our plant is 60000 TPA.
Therefore, $C_{OM,d} = \frac{Rs.780592185/year}{60000TPA}$ =Rs. 13009 /Ton
Therefore, Manufacturing Price of per kg Formalin= Rs.13
If Formalin is sold at a price higher than this, it will yield profit.

9.3 Direct Manufacturing Cost

All the precentages are taken and reviewed from the literature. All these percentages give a general idea about the kind of costs a company would be bearing.

- Operating Labour Cost (C_{OL})=Rs.21871216/year

- Direct Supervisory Labour $= 0.18(C_{OL})$=Rs.3936833/year

- Maintenance and Repairs= $0.06FCI$ =Rs.60750612/year

- Operating Supplies $= 0.009FCI$ =Rs. 9112591.8/year

- Laboratory Charges $= 0.15C_{OL}$ =Rs.3280682 /year

- Patents and Royalties $= 0.03C_{OM}$=Rs.26455284/year

Total Direct Manufacturing Cost : Rs. 563319315/year

9.4 Fixed Manufacturing Cost

- Depreciation= $0.1FCI$ =Rs. 101251020/year

- Local Taxes and Insurance= $0.032FCI$ =Rs.32400326/year

- Plant Overhead Cost= $0.708C_{OL} + 0.036FCI$ =Rs. 51935188/year

Total Fixed Manufacturing Cost : Rs.185586535 /year

9.5 General Manufacturing Expenses

These expenses gives us a rough idea about the cash outflow occuring in the plan once the plant becomes operational. These expenses should be looked after by the revenue generated.

- Administration Costs= $0.177C_{OL} + 0.009FCI$ =Rs.12983797 /year

- Distribution and Selling Costs= $0.11C_{OM}$ =Rs. 97002746/year

- Research and Development= $0.05C_{OM}$ =Rs. 44092157 /year

Total General Manufacturing Expenses : Rs.154078701 /year

10   Conclusion

The preliminary work (Mass Balances & Energy Balances) were done around the columns and other equipment. We got the composition of all the streams and the flow rates per hour for the production of 37% Formaldehyde and also computed the heat duties and power required for the columns and equipment installed. Designing of Distillation column and other equipment design(Reboiler, Condenser, Heat Exchanger) was carried out. The cost estimation using Order of the Magnitude Approach was carried out. The cost of manufacturing one ton of Formalin was found out.

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