CH2180 Separation Processes

Distillation Column Design

Sieve Tray

Group 5

Name	Index no
ARUSHAN R.	220045M
DE SILVA M.B.M.R	220101H
PERERA W.S.D	220477M
DISSANAYAKE R.T	220141E
OMATHTHAGE S.Y	220438V
VARUNAN S.	220660T

Department of Chemical and Process Engineering

University of Moratuwa

Table of Contents

Section 1 – Introducing the design problem

Distillation is a separation process that plays an integral role in many chemical engineering applications and industrial processes. It involves the separation of components in mixtures based on their relative volatilities. This paper discusses the implementation of a sieve tray distillation column for the separation of an n-hexane and n-heptane mixture at atmospheric pressure (1 bar).

It is desired to keep the necessary feed, distillate, and bottom product compositions constant and accomplish effective separation at a feed rate of 50 kmol/hr. Besides meeting technological specifications, column design endeavours to ensure steady performance under actual operating conditions. The column would be operated at 80% of its flooding capacity to maximize throughput without sacrificing product quality, thereby a compromise between operating effectiveness and safety.

The paper also stresses the significance of engineering considerations in practice in addition to the theoretical basis. This includes the choice of a suitable column diameter, careful tray design, and establishment of tray spacing. Specific parameters, such as the number and size of tray openings, are examined for the purpose of guaranteeing both the mechanical integrity of the column and operational reliability.

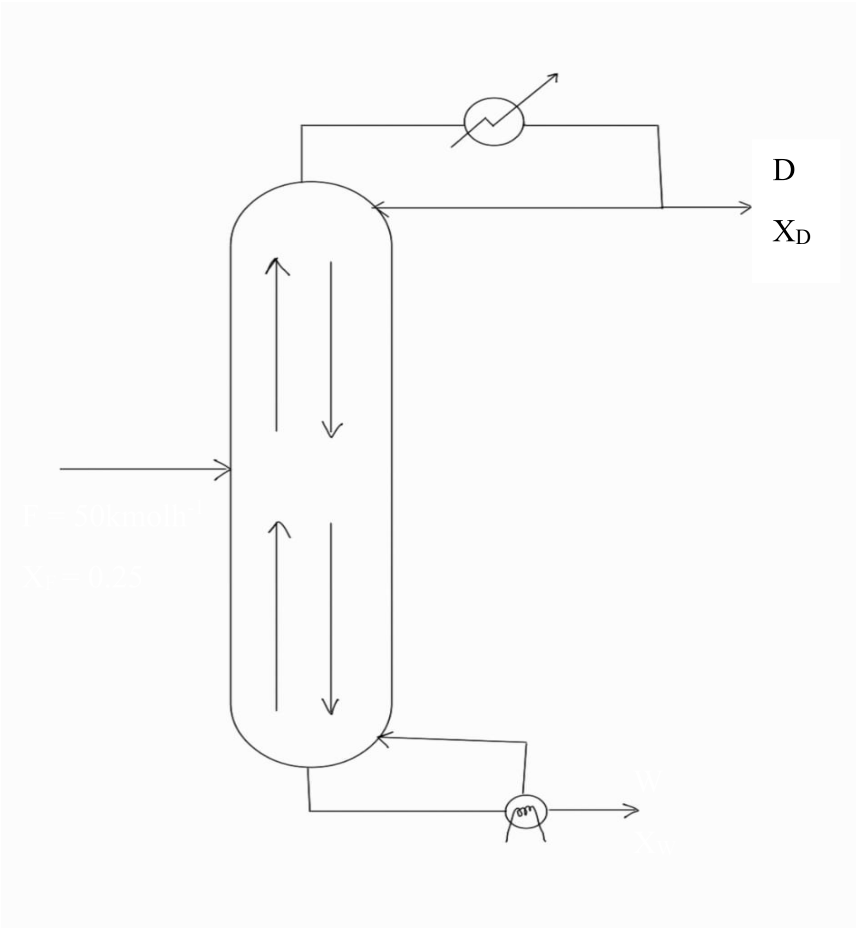
Taking everything into consideration, this study presents a comprehensive guide to designing and running a distillation column that is technically efficient and reliable, and effective under real industrial environments.

Specifications

Table 1: Specification table

Property	**Design Requirement**
System	n-hexane n-heptane
Operating pressure	1 bar
Feed rate	50 kmol/hr
Feed condition	Saturated liquid
Feed composition	25 mol% hexane
Distillate composition	95 mol% hexane
Column type	Sieve tray column
Operating condition	80 Percent of flooding
Condenser	Total
Reboiler	Partial

Calculations are as below.



Apply material balance for the whole column,

F = D + W -------- (1)

Apply material balance for MVC in the column,

$F.X_F = D.X_D + W.X_W$

$50 \times 0.25 = D \times 0.95 + W \times X_W$ ----------(2)

Since feed composition is 25 mol% hexane,

Hexane in feed = $\frac{25}{100} \times 50 = 12.5\ kmolh^{-1}$

85% of hexane recovery is required.

Hexane in distillate = $12.5 \times \frac{85}{100} = 10.625\ kmolh^{-1}$

Therefore, 10.625 $kmolh^{-1}$ of hexane should go to the distillate.

Since distillate contains 95% of hexane,

$D \times 0.95 = 10.625$

$D = 11.184\ kmolh^{-1}$

From (1),

$F = D+W$

$W = 50 - 11.184 = 38.816$ kmolh^{-1}

From (2),

$F.\ X_F = D.X_D + W.X_W$

$50\times 0.25 = 11.184 \times 0.95 + 38.816 \times X_W$

$X_W = 0.048\ kmolh^{-1}$

Section 2 – Suitability of using a tray column for this design problem

The feed enters the distillation column as a saturated liquid. Tray columns are particularly useful for such mixed-phase feeds since they can enable efficient separation of components and facilitate effective phase-to-phase interaction. The aim of this process is to generate a distillate that contains 95 mol% hexane and a bottom product that contains just 4.8 mol% hexane. Tray columns have an unmatched capability to handle such high compositional differentials efficiently.

The column is at 1 bar pressure level, which is in accordance with the conventional tray column equipment and materials of construction standard pressure requirements. Tray columns also have high design flexibility since they can accommodate changes in elements such as tray types, liquid flow arrangements, and tray separation. These elements could be optimized to accommodate specific performance objectives and processing conditions.

With the McCabe-Thiele method, it is possible to calculate important design parameters such as tray efficiency, the optimal number of theoretical stages, and the best location of the feed tray. These are very important calculations that help ensure that the column operates at the highest capacity. Thus, a tray distillation column is a suitable and effective option for the separation of n-hexane and n-heptane at the specified operating conditions.

Section 3 - Procedure used to obtain vapor pressure and other required physical property data for the system using Excel

The data provided for vapor pressures of the two components n-hexane and n-heptane at varying temperatures was used to obtain the compositions x and y pertaining to the more volatile component (n-hexane).

The temperature composition diagram was obtained by plotting 2 separate curves for temperature versus x and temperature versus y.

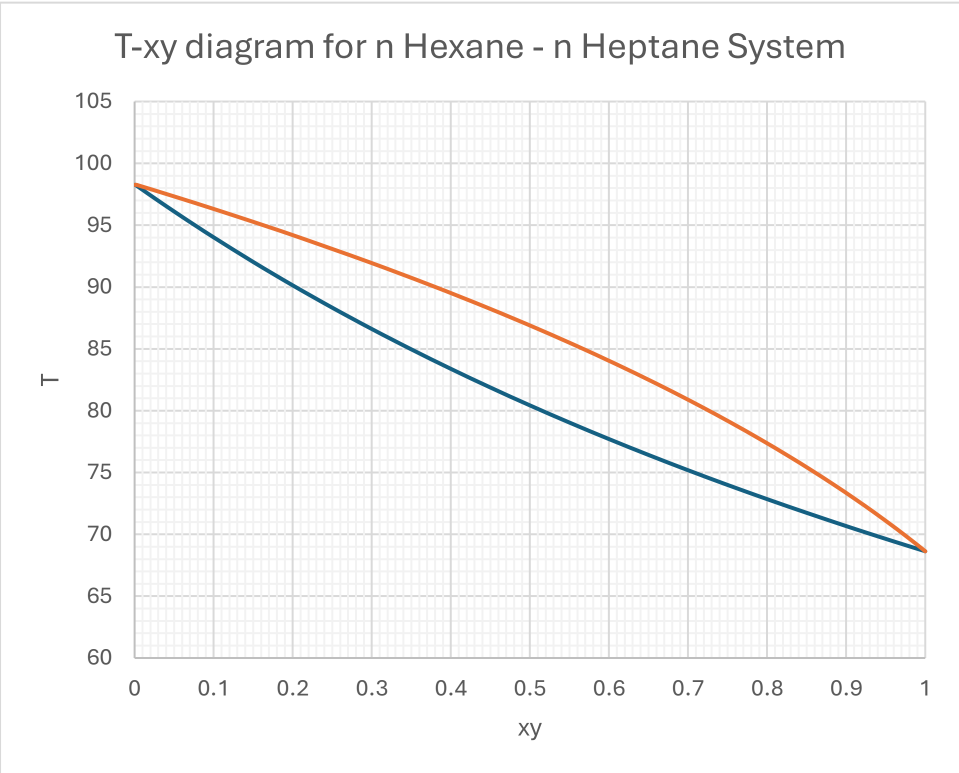


Figure 1:T-xy diagram for n Hexane - n Heptane System

The vapor liquid equilibrium (VLE) curve was obtained by plotting y values against the corresponding x values.

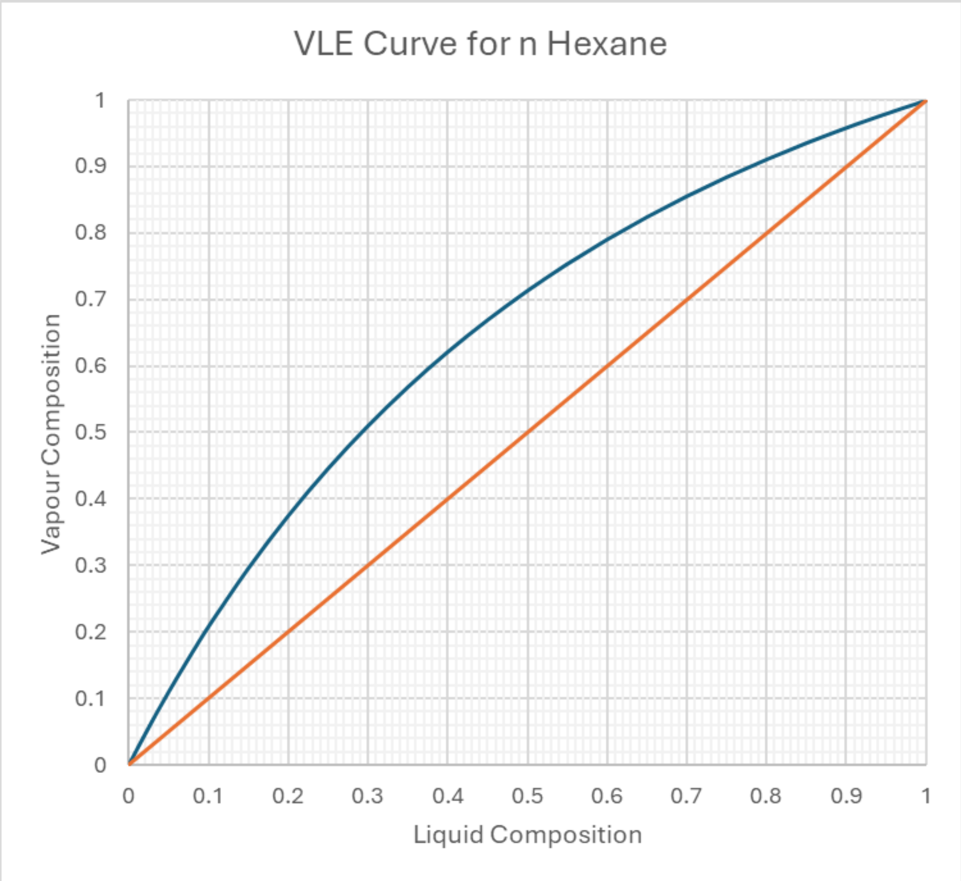


Figure 2: VLE curve

Section 4 – Determining the minimum reflux

The q line is drawn according to the given data

q = $\frac{Heat\ required\ to\ vapourize\ 1\ mole\ of\ feed}{Molar\ latent\ heat\ of\ the\ fee\square} = \frac{\lambda}{\lambda} = 1$

q = 1; feed is saturated liquid

q line equation is as below.

$$y = \frac{q}{q-1}x - \frac{1}{q-1}x_f$$

Gradient: $\frac{q}{q-1} = \frac{1}{1-1} \rightarrow \infty$

Therefore, the q line is a vertical line passing through (0.048,0.048)

The top operating line (TOL) equation is as follows:

$$TOL: \quad y = \frac{R}{R+1}x + \frac{1}{R+1}x_D$$

The TOL passes through point (x_D,x_D), which has coordinates of (0.95,0.95).

At the minimum reflux ratio, the point where the top operating line intersects the q-line falls on the equilibrium curve.

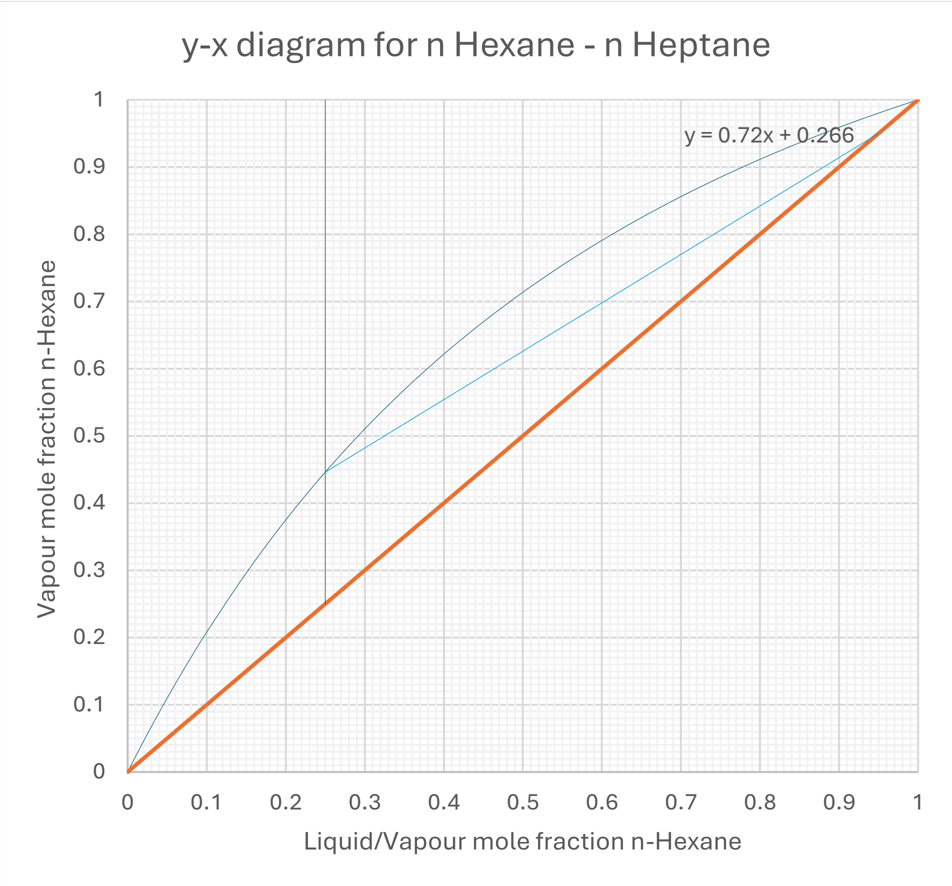


Figure 3: y-x diagram for the Rmin value

Accordingly, the TOL equation was obtained as y = 0.72x + 0.266 for R = R_{min}

Using the standard TOL equation, the gradient = $\frac{R}{R+1}$

Therefore, $\frac{Rmin}{Rmin+1}$ = 0.72

R_{min} = 2.5714

Section 5 – Determining the actual reflux ratio

- When R = 1.2 R_{min},

R = 1.2 × 2.5714

R = 3.08568

$$TOL: \quad y = \frac{R}{R+1}x + \frac{1}{R+1}x_D$$

$$y = \frac{3.08568}{3.08568+1}x + \frac{1}{3.08568}x_D$$

y = 0.7552x + 0.3241x_D

q-line goes through (x_f, x_f)

BOL: $y = \frac{L'}{L'-w}x + \frac{w}{L'-w}x_w$

BOL can be drawn by joining the intersection point of the TOL and q line and (x_w,x_w)

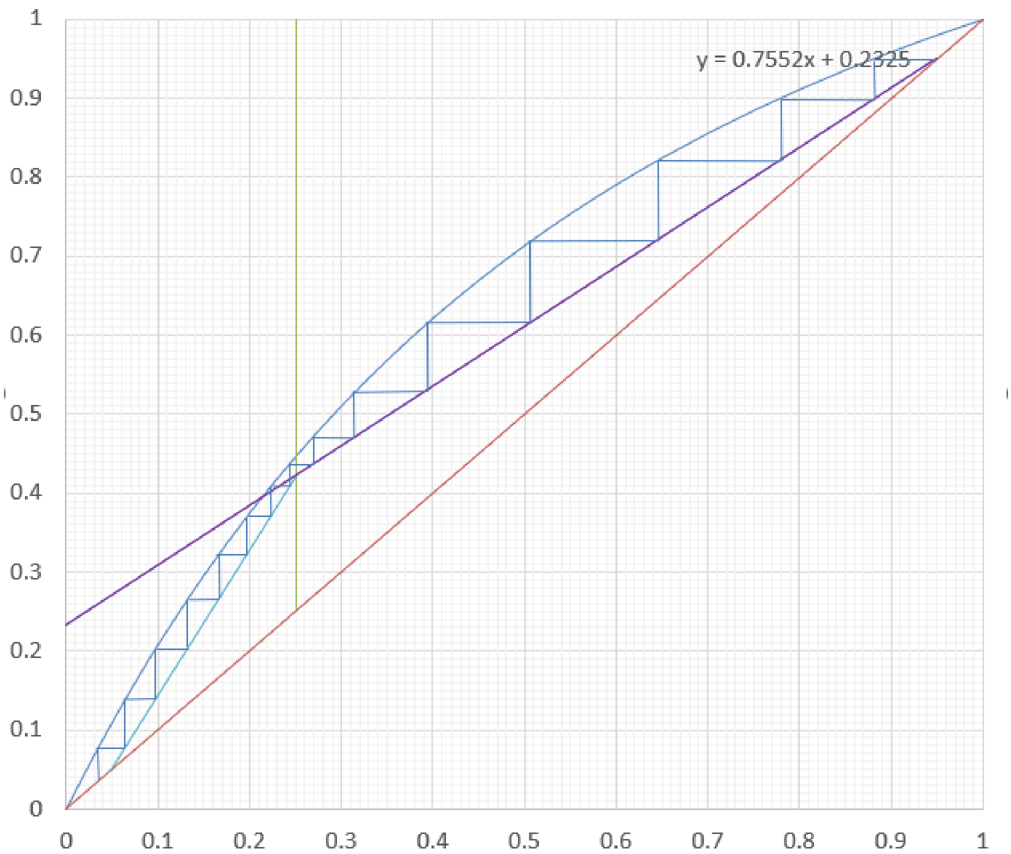


Figure 4: y-x diagram with stages drawn according to McCabe-Thiele method when R=1.2Rmin

No of stages = 15

As total condenser is used,

NTP = no of stages - 1

= 15-1

= 14

Feed plate from the top = 8^{th} plate

The same procedure can be followed to construct the graph and to find the NTP for each R value. Since a total condenser is used, NTP = no. of stages - 1

- When R= $1.3R_{min}$,

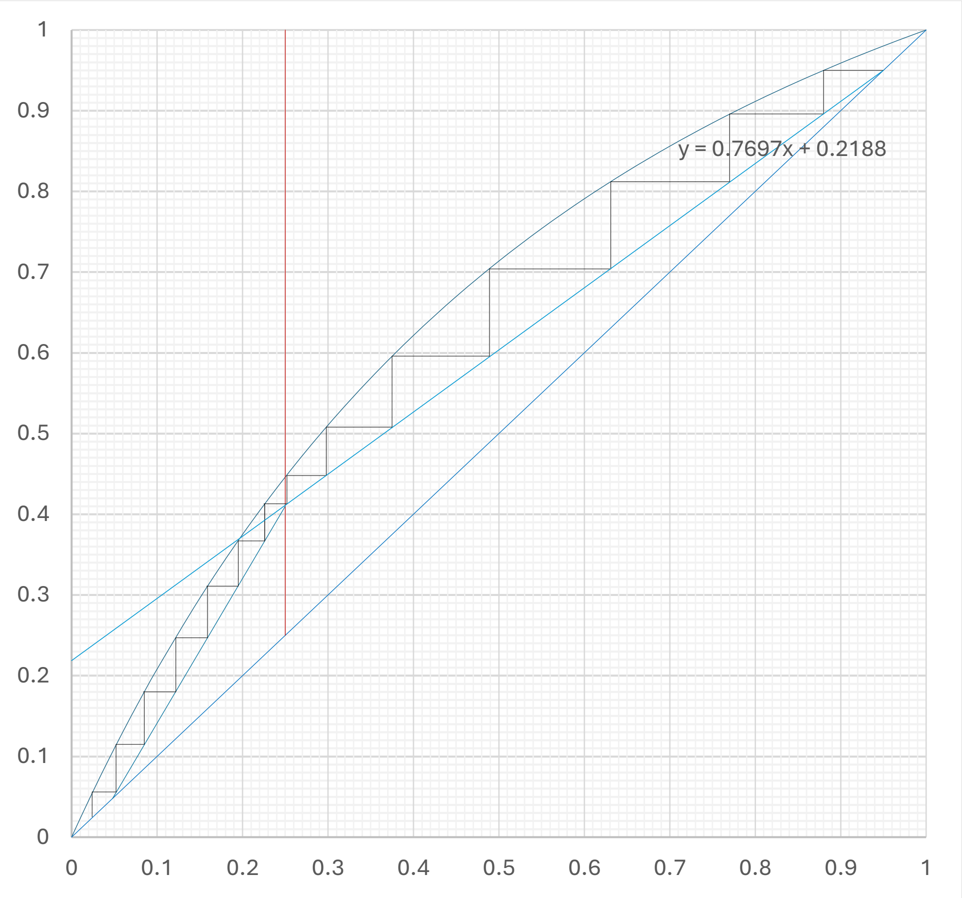


Figure 5: y-x diagram with stages draw using Mc-Cabe and Thiele method when R=1.3Rmin

No of stages = 14

As total condenser is used,

NTP = no of stages - 1

= 14-1

= 13

Feed plate from the top = 8th plate

- When R= 1.4R_{min} ,

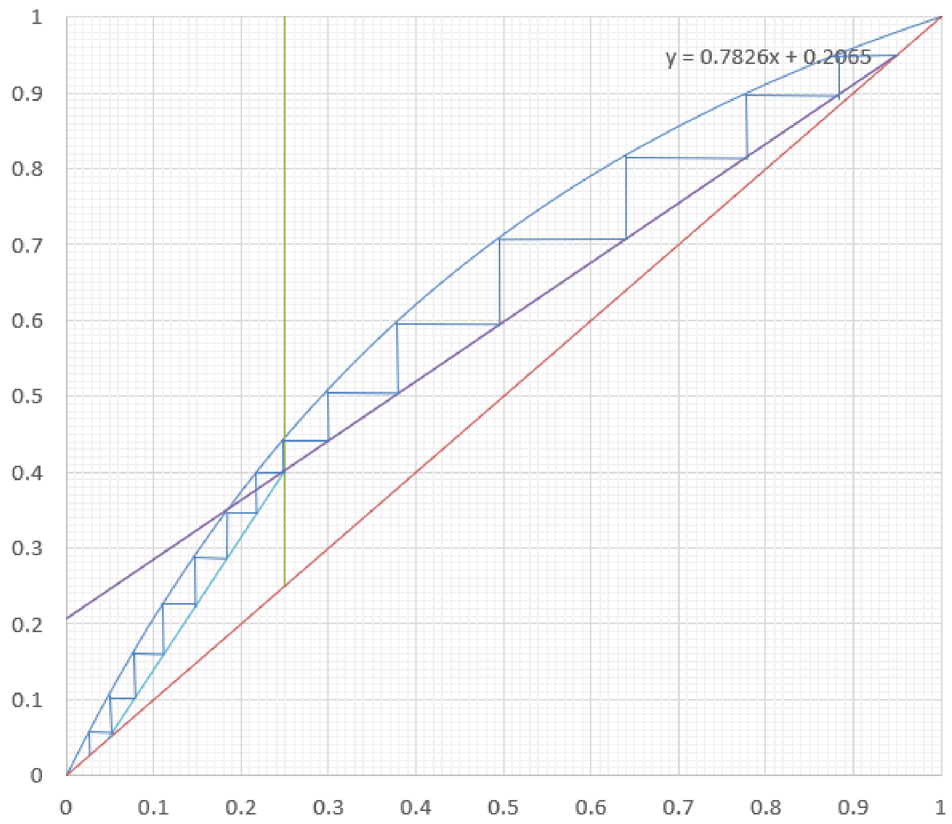


Figure 6: y-x diagram with stages drawn using Mc-Cabe and Thiele method when R=1.4Rmin

No of stages = 14

NTP = no of stages - 1

= 14-1

= 13

Feed plate from the top = 7th plate

Table 2: Reflux ratio and No of Plates for each reflux ratios

$R_{min} = 2.5714$	Reflux ratio (R)	NTP
1.2 R_{min}	3.08568	14
1.3 R_{min}	3.34282	13
1.4 R_{min}	3.59996	13

Reason to select $R_{operating}$ =1.3 x R_{min}:

Selecting a lower reflux ratio results in increased capital costs due to the requirement for larger equipment to achieve the desired separation. Conversely, a higher reflux ratio leads to higher energy consumption, as it demands more reboiling and condensing, thereby reducing the net present value (NPV) of the process.

The optimum reflux ratio is chosen to minimize the total cost, which includes both capital and operating expenses. Based on this trade-off, a value of $R_{operating}$ =1.3 ×R_{min} was selected as it provides a balanced and cost-effective solution.

Section 6 – Calculating temperatures & compositions of each tray

The temperature value corresponding to each x(n-hexane) was read from the temperature composition diagram. (Figure.1)

Table 3: Calculating temperatures & compositions of each tray

Plate	X(n-hexane)	Y(n-hexane)	Temperature
1	0.88	0.95	71
2	0.77	0.896	73.5
3	0.631	0.812	76.8
4	0.489	0.704	80.7
5	0.375	0.596	84.2
6	0.298	0.508	86.8
7	0.252	0.448	88.3
8	0.226	0.413	89.2
9	0.195	0.367	90.3
10	0.159	0.311	91.7
11	0.122	0.247	93.1
12	0.085	0.18	94.8
13	0.052	0.115	96.1
14	0.024	0.056	97.2

Section 7 - Finding the average liquid and vapor density values for each plate

Using Aspen Plus software, a set of density values were obtained for n-Hexane and n-Heptane in the temperature range 70-100 ^{0}C. The density values corresponding to the temperature of each plate were then found through interpolation.

Table 4:Average liquid density values for each plate

Plate	Temp (°C)	n-hexane composition in liquid	Density of liquid n-hexane (kg/m3)	n-heptane composition in liquid	Density of liquid n-heptane (kg/m3)	Average liquid density (kg/m3)
1	71	0.88	612.2	0.12	640.77	615.6284
2	73.5	0.77	609.76	0.23	648.5	618.6702
3	76.8	0.631	606.7	0.369	635.5	617.3272
4	80.7	0.489	602.4	0.511	631.7	617.3723
5	84.2	0.375	598.67	0.625	628.4	617.25125
6	86.8	0.298	596.13	0.702	625.959	617.069958
7	88.3	0.252	594.5	0.748	624.5	616.94
8	89.2	0.226	593.59	0.774	623.65	616.85644
9	90.3	0.195	592.41	0.805	622.5	616.63245
10	91.7	0.159	591	0.841	621.22	616.41502
11	93.1	0.122	589.3	0.878	619.7	615.9912
12	94.8	0.085	587.6	0.915	618.2	615.599
13	96.1	0.052	586	0.948	616.8	615.1984
14	97.2	0.024	585.1	0.976	616	615.2584

Table 5: Average vapor density values for each plate

Plate	Temp (°C)	n-hexane composition in vapour	Density of vapour n-hexane (kg/m3)	n-heptane composition in vapour	Density of vapour n-heptane (kg/m3)	Average vapour density (kg/m3)
1	71	0.95	3.11	0.05	3.49	3.129
2	73.5	0.896	2.98	0.104	3.475	3.03148
3	76.8	0.812	2.96	0.188	3.445	3.05118
4	80.7	0.704	2.92	0.296	3.4	3.06208
5	84.2	0.596	2.89	0.404	3.37	3.08392
6	86.8	0.508	2.8796	0.492	3.348	3.1100528
7	88.3	0.448	2.865	0.552	3.33	3.12168
8	89.2	0.413	2.86	0.587	3.3259	3.1334833
9	90.3	0.367	2.85	0.633	3.314	3.143712
10	91.7	0.311	2.84	0.689	3.3	3.15694
11	93.1	0.247	2.827	0.753	3.28	3.168109
12	94.8	0.18	2.818	0.82	3.277	3.19438
13	96.1	0.115	2.8	0.885	3.26	3.2071
14	97.2	0.056	2.8	0.944	3.25	3.2248

Table 6: Liquid Molar Volumes

Plate	Temp (°C)	n-hexane composition in liquid	Molar volume of liquid n-hexane (m3/kmol)	n-heptane composition in liquid	Molar volume of liquid n-heptane (m3/kmol)	Average liquid Molar Volume (m3/kmol
1	71	0.88	0.14077099	0.12	0.156374362	0.142643395
2	73.5	0.77	0.141334295	0.23	0.154510409	0.144364801
3	76.8	0.631	0.14204714	0.369	0.157671125	0.147812391
4	80.7	0.489	0.143061089	0.511	0.158619598	0.151011487
5	84.2	0.375	0.143952428	0.625	0.159452578	0.153640022
6	86.8	0.298	0.144565783	0.702	0.160074382	0.155452819
7	88.3	0.252	0.144962153	0.748	0.160448359	0.156545835
8	89.2	0.226	0.145184387	0.774	0.160667041	0.157167961
9	90.3	0.195	0.145473574	0.805	0.160963855	0.157943251
10	91.7	0.159	0.145820643	0.841	0.161295515	0.158835011
11	93.1	0.122	0.146241303	0.878	0.161691141	0.159806261
12	94.8	0.085	0.146664398	0.915	0.162083468	0.160772847
13	96.1	0.052	0.147064846	0.948	0.162451362	0.161651263
14	97.2	0.024	0.147291061	0.976	0.162662338	0.162293427

Table 7: Vapour molar volumes

Plate	Temp (°C)	n-hexane composition in vapour	Molar volume of vapour n-hexane (m3/kmol)	n-heptane composition in vapour	Molar Volume of vapour n-heptane (m3/kmol)	Average vapour Molar Volume (m3/kmol)
1	71	0.95	27.71061093	0.05	28.71060172	27.76061047
2	73.5	0.896	28.91946309	0.104	28.83453237	28.91063029
3	76.8	0.812	29.11486486	0.188	29.08563135	29.10936896
4	80.7	0.704	29.51369863	0.296	29.47058824	29.50093795
5	84.2	0.596	29.8200692	0.404	29.73293769	29.78486807
6	86.8	0.508	29.92776775	0.492	29.92831541	29.9280372
7	88.3	0.448	30.08027923	0.552	30.09009009	30.08569483
8	89.2	0.413	30.13286713	0.587	30.12718362	30.12953091
9	90.3	0.367	30.23859649	0.633	30.23536512	30.23655103
10	91.7	0.311	30.34507042	0.689	30.36363636	30.35786236
11	93.1	0.247	30.48461266	0.753	30.54878049	30.53293104
12	94.8	0.18	30.58197303	0.82	30.57674702	30.57768771
13	96.1	0.115	30.77857143	0.885	30.73619632	30.74106946
14	97.2	0.056	30.77857143	0.944	30.83076923	30.82784615

Section 8: Calculating the mass flow rates for each plate

<u>Rectifying section</u>

$R_{operating} = 1.3 \text{ x } R_{min} = \frac{L_0}{D}$

$1.3 \text{ x } R_{min} = \frac{L_0}{11.184\ kmol/h}$

$L_0 = L$ (Assume constant molal overflow)

Therefore, L = 37.38609888 kmol/hr

V=L+D (material balance)

V= 37.38609888 +11.184 = 48.57009888 kmol/hr

Stripping section

Since feed enters as saturated liquid, no additional heat is required to vaporise the feed. Therefore, the vapor leaving the stripping section need not condense.

V' = V = 48.57009888 kmol/hr

L' = L + qF

L' = 37.3860988 + 1 * 50

L' = 87.38609888 kmol/hr

Using the following equations, mass flow rates were calculated for each tray of the column.

For i^th tray of rectifying section

$$L_{w,Tray\ (i)} = L[M_{w,comp1}x_{comp1,Tray\ (i)} + M_{w,comp2}(1 - x_{comp1,Tray\ (i)})]$$

$$V_{w,Tray\ (i)} = V[M_{w,comp1}y_{comp1,Tray\ (i)} + M_{w,comp2}(1 - y_{comp1,Tray\ (i)})]$$

For j^th tray of stripping section

$$L'_{w,Tray\ (j)} = L'[M_{w,comp1}x_{comp1,Tray\ (j)} + M_{w,comp2}(1 - x_{comp1,Tray\ (j)})]$$

$$V'_{w,Tray\ (i)} = V'[M_{w,comp1}y_{comp1,Tray\ (j)} + M_{w,comp2}(1 - y_{comp1,Tray\ (j)})]$$

$M_{n\text{-}hexane}$ = 86.18 g/mol

$M_{n\text{-}heptane}$ = 100.2 g/mol

Table 8: Calculating the mass flow rates for each plate

	Plate	T(⁰C)	X(n-hexane)	X(n-heptane)	Lw,Tray(i) (kg/hr)	L'w, Tray(i)(kg/hr)	Y(n-hexane)	Y(n-heptane)	Vw, Tray(i)(kg/hr)	V'w, Tray(i)(kg/hr)
Rectifying section	1	71	0.88	0.12	3284.832374		0.95	0.05	4219.818761	
	2	73.5	0.77	0.23	3342.489216		0.896	0.104	4256.590211	
	3	76.8	0.631	0.369	3415.346498		0.812	0.188	4313.790245	
	4	80.7	0.489	0.511	3489.776239		0.704	0.296	4387.333146	
	5	84.2	0.375	0.625	3549.529693		0.596	0.404	4460.876047	
	6	86.8	0.298	0.702	3589.889482		0.508	0.492	4520.799892	
	7	88.3	0.252	0.748	3614.000525		0.448	0.552	4561.65706	
Feed	8	89.2	0.226	0.774	3627.628506		0.413	0.587	4585.490407	
Stripping Section	9	90.3	0.195	0.805		8517.182252	0.367	0.633		4616.814235
	10	91.7	0.159	0.841		8561.287764	0.311	0.689		4654.947591
	11	93.1	0.122	0.878		8606.618429	0.247	0.753		4698.52857
	12	94.8	0.085	0.915		8651.949094	0.18	0.82		4744.152406
	13	96.1	0.052	0.948		8692.379146	0.115	0.885		4788.414337
Reboiler	14	97.2	0.024	0.976		8726.683433	0.056	0.944		4828.590552

Section 9: Calculating F_{lv} values

The following equation was used to calculate F_{lv} values for each tray.

$$\frac{L_{wi}{}^{*}}{V_{wi}{}^{*}}\sqrt{\frac{\rho_{Vi}}{\rho_{Li}}}$$

Table 9: Calculating Flv values

	Plate	Temp (^0C)	$\rho_{L,i}$ (kg/m³)	$\rho_{V,I}$ (kg/m³)	Lw*,Tray(i) (kg/hr)	Vw*,Tray(i) (kg/hr)	F_{LV}	Avg.F_{LV}
Rectifying section	1	71	615.6284	3.129	3284.832374	4219.818761	0.055496152	0.055873839
	2	73.5	618.6702	3.03148	3342.489216	4256.590211	0.054967493	
	3	76.8	617.3272	3.05118	3415.346498	4313.790245	0.055661122	
	4	80.7	617.3723	3.06208	3489.776239	4387.333146	0.056018524	
	5	84.2	617.25125	3.08392	3549.529693	4460.876047	0.056243355	
	6	86.8	617.06996	3.110053	3589.889482	4520.799892	0.05637447	
	7	88.3	616.94	3.12168	3614.000525	4561.65706	0.056355759	
Feed	8	89.2	616.85644	3.133483	3627.628506	4585.490407	0.05638436	
Stripping Section	9	90.3	616.63245	3.143712	8517.182252	4616.814235	0.131723035	0.131429519
	10	91.7	616.41502	3.15694	8561.287764	4654.947591	0.131619688	
	11	93.1	615.9912	3.168109	8606.618429	4698.52857	0.131366153	
	12	94.8	615.599	3.19438	8651.949094	4744.152406	0.131371046	
	13	96.1	615.1984	3.2071	8692.379146	4788.414337	0.131067674	
Reboiler	14	97.2	615.2584	3.2248	8726.683433	4828.590552	0.130843291	

Section 10 – Calculating column diameter

K1 values were obtained corresponding to average F_{lv} values for rectifying and stripping sections separately, using the chart below. Three different values were considered as possible options for tray spacing.

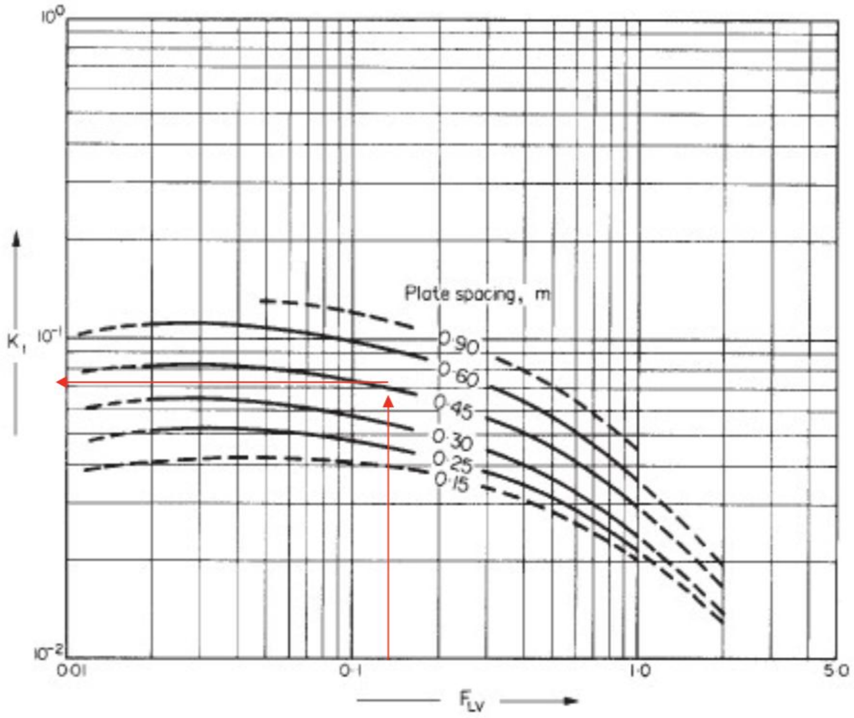


Figure 7:Flooding velocity curves

The following restrictions apply to the use of above Figure:

1. Hole size less than 6.5 mm. Entrainment may be greater with larger hole sizes.

2. Weir height less than 15 per cent of the plate spacing

3. Non-foaming systems.

4. Hole: active area ratio greater than 0.10; for other ratios apply the following corrections:

Hole:active area	Multiply K1 by
0.1	1.0
0.08	0.9
0.06	0.8

5. Liquid surface tension 0.02 N/m, for other surface tensions σ multiply the value of K1 by $[\sigma/0.02]^{0.2}$.

Table 10: K1 calculations for different tray spacing values

					Assume Tray space=0.6m			Assume Tray space=0.45m			Assume Tray space=0.30m		
	Plate	Average F_{LV}	$\rho_{L,i}$ (kg/m³)	$\rho_{V,i}$ (kg/m³)	K1 value	K1 *((ρL-ρV)/ρV)^0.5	K1*[((ρL-ρV)/ρV)^0.5] min	K1 value	K1 *((ρL-ρV)/ρV)^0.5	K1*[((ρL-ρV)/ρV)^0.5] min	K1 value	K1 *((ρL-ρV)/ρV)^0.5	K1*[((ρL-ρV)/ρV)^0.5] min
Rectifying section	1		615.6284	3.129		1.818835426			1.119283339			0.867444588	
	2		618.6702	3.03148		1.852588492			1.140054456			0.883542204	
	3		617.3272	3.05118		1.844553341			1.135109748			0.879710055	
	4	0.05587384	617.3723	3.06208	0.13	1.841318668	1.818835426	0.08	1.13311918	1.11928334	0.062	0.878167365	0.867444588
	5		617.25125	3.08392		1.834573669			1.128968412			0.874950519	
	6		617.069958	3.1100528		1.826541215			1.124025363			0.871119657	
	7		616.94	3.12168		1.82292619			1.121800732			0.869395567	
Feed	8		616.85644	3.1334833		1.819348279			1.119598941			0.867689179	
	9		616.63245	3.143712		1.396953406			0.977867384			0.810232976	
	10		616.41502	3.15694		1.393761538			0.975633076			0.808381692	
	11	0.13142952	615.9912	3.168109	0.1	1.390809034	1.381390208	0.07	0.973566324	0.96697315	0.058	0.80666924	0.801206321
	12		615.599	3.19438		1.384605142			0.969223599			0.803070982	
	13		615.1984	3.2071		1.381390208			0.966973146			0.801206321	
Reboiler	14		615.2584	3.2248		1.790934045			1.102113258			0.854137775	

The following equations were used to calculate the net area for rectifying and stripping sections separately. The flooding percentage relevant to our column design was 80%.

$$\therefore A_{n,max,rectifying} = \frac{V_{max}}{U_{a,min}} = \frac{(Molar\ volume)_{max,rectifying} \times V}{\left(\begin{matrix}Flooding\\ \%\end{matrix}\right) \times K_{1,rectifying} \times \left(\sqrt{\frac{\rho_L - \rho_V}{\rho_L}}\right)_{min,rectifying}}$$

$$\therefore A_{n,max,stripping} = \frac{V'_{max}}{U_{a,min}} = \frac{(Molar\ volume)_{max,stripping} \times V'}{\left(\begin{matrix}Flooding\\ \%\end{matrix}\right) \times K_{1,stripping} \times \left(\sqrt{\frac{\rho_L - \rho_V}{\rho_L}}\right)_{min,stripping}}$$

With the assumption that downcomer area = 12% of the total column cross sectional area, the following equation was used to determine column area and downcomer area.

$$A_n = A_c - A_d = 0.88A_c$$

Column diameter was calculated using the equation $A_c = \pi d^2 / 4$

All calculated results relevant to our selected tray spacing value of 0.45m are outlined in the table below.

Table 11: Diameter calculations

	Plate	Average F_{LV}	K1	K1 *((ρL-ρV)/ρV)^0.5	K1*[((ρL-ρV)/ρV)^0.5] min	Average vapor molar volume (m3/kmol)	Maximum vapor molar volume (m3/kmol)	An	Ac	Ad	dc	Column diameter
Rectifying section	1	0.05587384	0.08	1.119283339	1.119283339	27.76061047						
	2			1.140054456		28.91063029						
	3			1.135109748		29.10936896						
	4			1.13311918		29.50093795						
	5			1.128968412		29.78486807						
	6			1.124025363		29.9280372						
	7			1.121800732		30.08569483	30.08569483	0.453311	0.515126432	0.06181517	0.809863781	
Feed	8			1.119598941		30.12953091						
Stripping section	9	0.13142952	0.07	0.977867384	0.966973146	30.23655103						
	10			0.975633076		30.35786236						
	11			0.973566324		30.53293104						
	12			0.969223599		30.57768771						
	13			0.966973146		30.74106946	30.74106946	0.536144	0.609253996	0.07311048	0.880753246	0.880753246
Reboiler	14			1.102113258		30.82784615						

Since the diameter of the stripping section was larger than that required for the rectifying section, it was decided that the column should be designed to fit the requirements of the stripping section.

The calculated column diameter was checked against the selected tray spacing value to ensure that they correspond with each other. According to the guidelines listed below, a column with diameter less than 1m should have a tray spacing of 0.5 m or less. Our selected value of 0.45m was in agreement with this.

1. Tray spacing

Tower diameter T		Tray spacing t	
m	ft	m	in
		0.15	6 minimum
1 or less	4 or less	0.50	20
1-3	4-10	0.60	24
3-4	10-12	0.75	30
4-8	12-24	0.90	36

Figure 8: Recommended general conditions and dimensions for tray tower

Section 11: Determining suitable liquid flow arrangements

Table 12: Determining suitable liquid flow arrangements

	Plate	Temp (^0C)	$\rho_{L,i}$ (kg/m^3)	min $\rho_{L,i}$	$L_{w,I}$ (kg/hr)	max L_w
Rectifying section	1	71	615.6284	615.1984	3284.832374	8692.379146
	2	73.5	618.6702		3342.489216	
	3	76.8	617.3272		3415.346498	
	4	80.7	617.3723		3489.776239	
	5	84.2	617.25125		3549.529693	
	6	86.8	617.069958		3589.889482	
	7	88.3	616.94		3614.000525	
Feed	8	89.2	616.85644		3627.628506	
Stripping Section	9	90.3	616.63245		8517.182252	
	10	91.7	616.41502		8561.287764	
	11	93.1	615.9912		8606.618429	
	12	94.8	615.599		8651.949094	
	13	96.1	615.1984		8692.379146	
Reboiler	14	97.2	615.2584		8726.683433	

Maximum liquid volume flow rate ($L_{w,\,max}$) = $\frac{(L_W)max}{(\rho_{L,i)}min}$

$L_{w,\,max}$ = 8692.379146/615.1984 = 14.12939167 m^3/hr = 0.00392483 m^3/s

The following diagram was used to find the relevant flow arrangement for a column with diameter 0.88075m and a maximum liquid flowrate of 3.98*10^{-3} m^3/s. Accordingly, the desired liquid flow arrangement was determined to be cross flow (single pass).

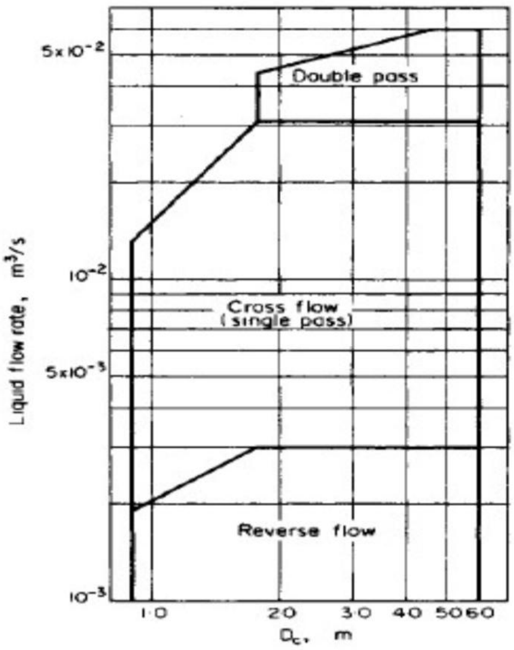


Figure 9: Selection of liquid flow arrangement

Section 12: Determining size and number of holes in a tray

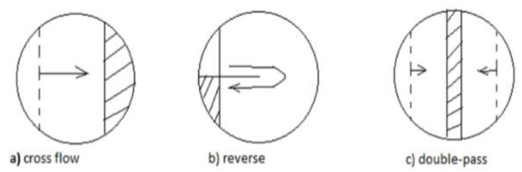


Figure 10: Choice of plate type

For a cross flow single pass plate type, the following equation may be used to determine the active area.

$A_a = A_c - 2A_d$

Based on the previously calculated values for column diameter, column cross sectional area and downcomer area, the following results were obtained for active area.

Table 13:Column diameter, Column cross sectional Area, Downcomer area and active area calculation

		Rectifying	Stripping
Column Diameter	d_c (m)	0.809863781	0.880753246
Column Cross Sectional Area	A_c (m2)	0.515126432	0.609253996
Downcomer Area	A_d (m^2)	0.061815172	0.07311048
Active Area	A_a (m^2)	0.391496088	0.463033036

In order to calculate hole area, it was assumed that the total hole area was 10% of the active area.

The hole diameter was assumed to be 5mm, in keeping with the requirement that hole diameter should be less than 6.5mm to prevent entrainment.

The area of a single hole was then calculated using the equation

$$\text{Single hole area} = \frac{\pi\, d_h^2}{4}$$

The number of holes was determined by dividing the total hole area (A_h) by the area of a single hole. The results of these calculations are outlined below.

Table 14:Hole area, Hole diameter, Area of a single hole and Number of holes calculation.

	Rectifying	Stripping
Hole area (Ah)	0.039149609	0.046303304
Hole diameter (D_h)	0.005	0.005
Area of a Single hole	0.000019635	0.000019635
Number of Holes	1993.8732	2358.207888
	1994	2359

Section 13: Weir height, weir length and weir liquid crest calculations

For distillation columns operating at atmospheric pressure, the recommended range for weir height is 40-90mm. In order to use Figure 8, it was required that weir height should be less than 15% of the selected tray spacing. Since our column tray spacing was 0.45m, the maximum possible weir height was 0.0675m.

Keeping with both these requirements, a weir height of 0.067m (67mm) was assumed for our calculations.

The following chart was used to calculate the weir length.

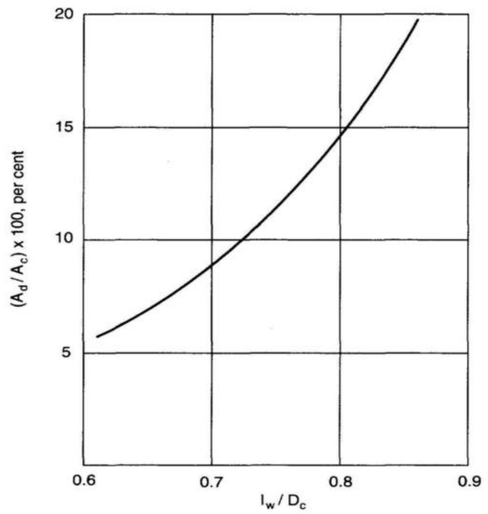


*Figure 11: plot of A_d/A_c * 100 vs l_w/d_c*

The l_w/d_c value corresponding to the A_d/A_c ratio was read from the curve, for rectifying and stripping sections separately. The calculated column diameter for each section was then used to obtain the value for weir length. The relevant values are outlined in the table below.

Table 15:Weir length calculation.

	Rectifying	**Stripping**
Ad (m)	0.061815172	0.07311
A_c (m^2)	0.515126432	0.609254
A_d/A_c	0.12	0.12
l_w/D_c	0.76	0.76
D_c (m)	0.809863781	0.880753
Weir length (m)	0.615524	0.669372

The height of the weir liquid crest was calculated using the Francis weir formula:

$h_{ow} = 750 \left(\frac{L_W}{\rho_L l_W}\right)^{\frac{2}{3}}$

Where h_{ow} = weir crest height in mm liquid, l_w = weir length in m and L_w = liquid flow rate in kg/s

L_w was taken as the average liquid flow rate and ρ_L was taken as the average liquid density across the trays in the rectifying and stripping sections separately.

Table 16:liquid crest height calculation

	Rectifying	**Stripping**
Ad (m)	0.061815172	0.07311048
Ac (m^2)	0.515126432	0.609253996
Ad/Ac	0.12	0.120000001
lw/Dc	0.76	0.76
Dc (m)	0.809863781	0.880753246
weir length (m)	0.615524	0.669372467

$$h_{ow} = 750\left[\frac{L}{\rho_L l_w}\right]^{2/3}$$

how (mm liquid	13.95057468	24.20506602

Section 14: Column entrainment

Entrainment refers to the liquid carried by the vapor from one plate to the plate above due to high vapor velocity.

The fractional entrainment can be determined using the percent flood and F_{lv} values, based on the following graph, which is a correlation provided by Fair (1961).

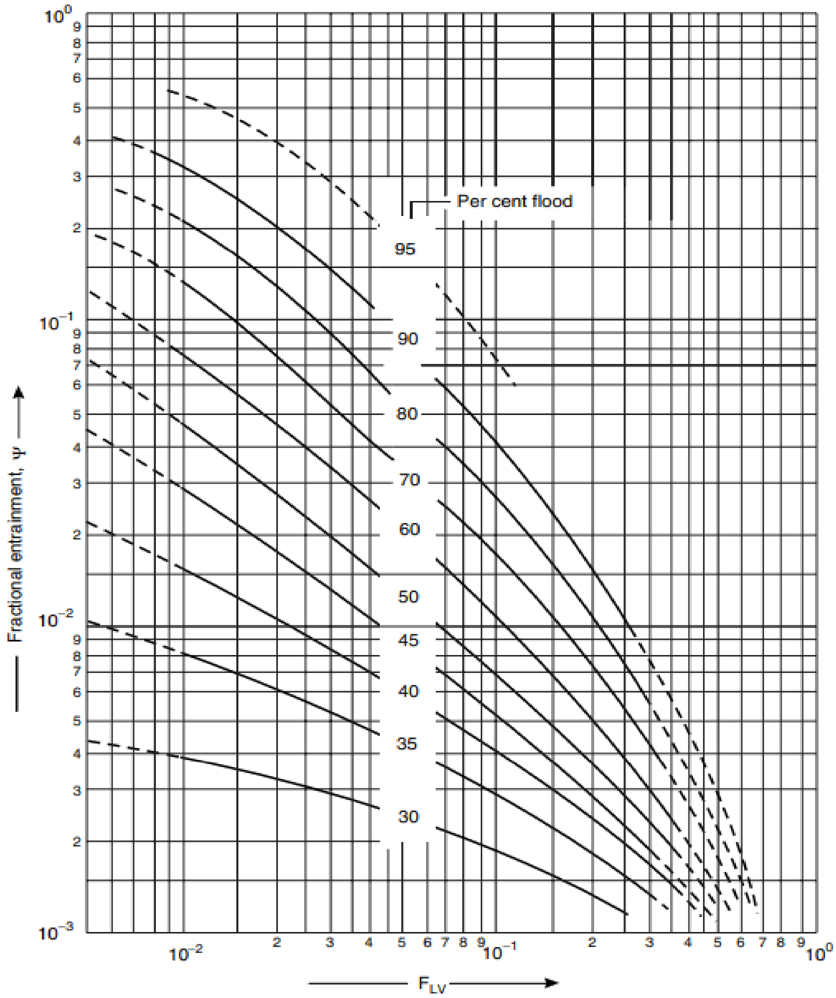


Figure 12: Fractional entrainment vs Flv

The fractional entrainment value corresponding to calculated F_{lv} values for rectifying and stripping sections were read from the chart using the 80% flooding curve.

The resulting fractional entrainment was 0.05 for the rectifying section and 0.019 for the stripping section.

An entrainment value greater than 0.1 would not be advisable since it would lead to reduced efficiency in the column. [2] Since our entrainment values were less than 0.1 for both rectifying and stripping sections of the column, it was determined that the entrainment in our design was at a satisfactory level.

Table 17:Fractional entrainment calculation

	Plate	Temp (^{0}C)	Average Flv	Fractional Entrainment Ψ(by graph)
Rectifying section	1	71	0.055873839	0.05
	2	73.5		
	3	76.8		
	4	80.7		
	5	84.2		
	6	86.8		
	7	88.3		
Feed	8	89.2		
Stripping Section	9	90.3	0.131429519	0.019
	10	91.7		
	11	93.1		
	12	94.8		
	13	96.1		
Reboiler	14	97.2		

Section 15: Weeping

Weeping occurs when liquid leaks through the tray perforations (holes) rather than flowing properly through the downcomer. This phenomenon can be prevented if the actual vapor velocity ($U_{a,min}$) is maintained above the minimum vapor velocity required to keep the liquid supported on the tray ($U_{h,min}$).

The following equation was used to calculate $U_{a,min}$.

$$U_{a,min} = \frac{\min \quad vapor\ molar\ flow\ rate \cdot \min \quad molar\ volume\ at\ a\ tray}{Ah}$$

Since constant molal overflow is assumed under the McCabe-Thiele method, the minimum vapor molar flow rate is the same the overall vapor molar flow rate (calculated as V under material balances).

Table 18:Minimum actual vapor velocity calculation.

Plate	Temp (°C)	Average vapor molal volume (m3/kmol)	Minimum vapor molal volume (m3/kmol)	Minimum vapor flow rate (kmol/s)	Ah	Ua,min
1	71	27.76061047				
2	73.5	28.91063029				
3	76.8	29.10936896				
4	80.7	29.50093795	27.76061047	0.013491694	0.039149609	9.566830346
5	84.2	29.78486807				
6	86.8	29.9280372				
7	88.3	30.08569483				
8	89.2	30.12953091				
9	90.3	30.23655103				
10	91.7	30.35786236				
11	93.1	30.53293104	30.23655103	0.013491694	0.046303304	8.810220145
12	94.8	30.57768771				
13	96.1	30.74106946				
14	97.2	30.82784615				

To determine the minimum design vapor velocity ($U_{h,min}$) in m/s, the following equation was used.

$$U_{h,min} = \frac{K2 - 0.90(25.4 - dh)}{\rho_c c^{0.5}}$$

Where dh is the hole diameter in mm and K2 is a constant dependent on the depth of clear liquid on the plate.

The following chart was used to find the value for K2, using calculated values for $h_w + h_{ow}$.

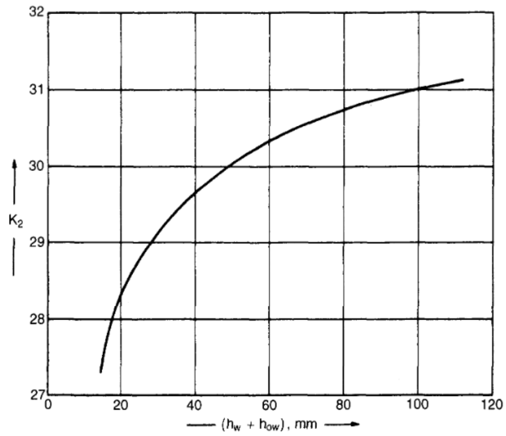


Figure 13: Plot of K2 vs ($h_w + h_{ow}$)

The results of the relevant calculations are listed below.

Table 19:minimum design vapor velocity calculation

	Rectifying	Stripping	
h_w+h_{ow}	80.95057468	91.20506602	mm
K_2	30.8	30.95	
U_h	7.076457092	4.540898169	m/s
$U_{a,min}$	9.566830346	8.810220145	m/s

The calculated values for $U_{h,min}$ are less than the values for $U_{a,min}$ in both rectifying and stripping sections. Therefore, weeping will not occur in the column.

Section 16: Calculating the total pressure drop across each tray in a distillation column

Hole diameter = 0.005m

Plate thickness = 0.005m (For our design, the thickness range is typically 5-10mm)

Plate thickness/hole diameter = 1

Assume Ah/Ap = Ah/Aa

Ah/Ap = 0.1

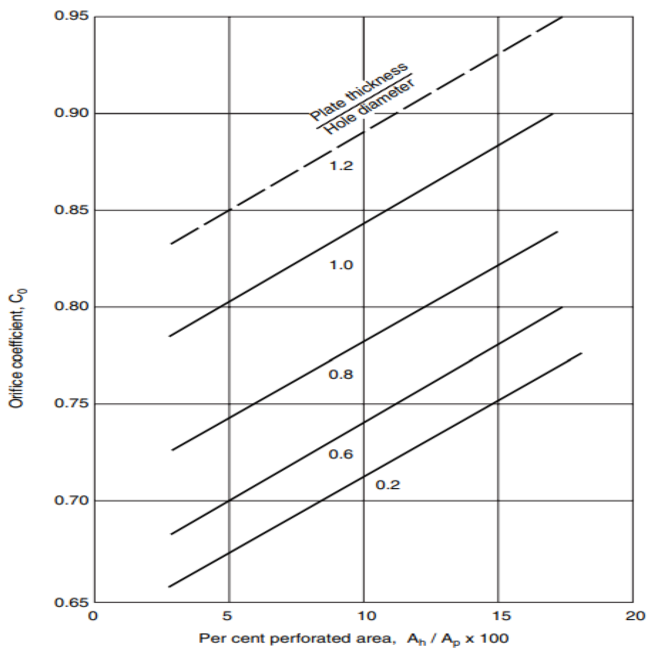


Figure 14: Plot of Ah/Aa vs C_0

Using the curve for which the ratio of plate thickness/hole diameter = 1 on the above chart, the C_0 value was found to be 0.84.

The pressure drop caused by vapor moving through dry tray holes (h_d) can be calculated using the following equation.

$$h_d = 51\left(\frac{U_h}{C_0}\right)^2 \frac{\rho_V}{\rho_L}$$

U_h=vapor velocity through tray holes

C_o=office coefficient

ρ_V = vapor density

ρ_L = liquid density

Minor pressure drop due to surface tension and liquid resistance across tray outlets (h_r) can be calculated using

$$h_r = \frac{12.5\times10^3}{\rho_L}$$

The total pressure drop across each plate can then be found using

$h_t = h_d + (h_w + h_{ow}) + h_r$

where h_w = weir height and h_{ow} = height of liquid over weir

The results obtained by applying the above equations are tabulated below.

Table 20:Pressure drop values

	Plate	h_d (mm liquid)	$h_w + h_{ow}$ (mm liquid)	h_r (mm liquid)	h_t (mm liquid)
Rectifying section	1	18.39629122	80.95057468	20.25341392	119.6002798
	2	17.73531348			118.9393021
	3	17.88940003			119.0933886
	4	17.95199639			119.155985
	5	18.08358304			119.2875716
	6	18.24217922			119.4461678
	7	18.31423625			119.5182248
Feed	8				
Stripping Section	9	7.598213753	91.20506602	20.29328788	119.0965677
	10	7.632876667			119.1312306
	11	7.665151394			119.1635053
	12	7.733637333			119.2319912
	13	7.769488607			119.2678425
Reboiler	14				

Section 17: Downcomer Design.

The following equations were used in our calculations.

$h_{ap} = h_w - (5 \text{ to } 10 \text{ mm})$

The height of the bottom edge of the apron above the plate was set to 8 mm to obtain a value of 59 mm for h_{ap}.

$A_{ap} = h_{ap} * l_w$

Using the calculated values for h_{ap} and l_w,

A_{ap} (rectifying) = 0.03632m^2

A_{ap} (stripping) = 0.03949m^2

A_d (rectifying) = 0.06182 m^2

A_d (stripping) = 0.07311 m^2

$$h_{dc} = 166\left[\frac{L_{wd}}{\rho_L A_m}\right]^2$$

since $A_d > A_{ap}$

h_{dc} (rectifying) = 0.306899775 mm

h_{dc} (stripping) = 1.603020548 mm

$$h_b = (h_w + h_{ow}) + h_t + h_{dc}$$

Table 21: Downcomer design calculations

	Plate	hw + how (mm liquid)	ht (mm liquid)	hdc(mm liquid)	hb(mm liquid)
Rectifying section	1	80.95057468	119.6002798	0.306899775	200.8577543
	2		118.9393021		200.1967765
	3		119.0933886		200.3508631
	4		119.155985		200.4134595
	5		119.2875716		200.5450461
	6		119.4461678		200.7036423
	7		119.5182248		200.7756993
Feed	8				
Stripping Section	9	91.20506602	119.0965677	1.603020548	211.9046542
	10		119.1312306		211.9393171
	11		119.1635053		211.9715919
	12		119.2319912		212.0400778
	13		119.2678425		212.0759291
Reboiler	14				

$0.5 * (l_t + h_w) = 258.5$ mm

$h_b < 0.5*(l_t + h_w)$

Our calculated values for h_b satisfy the above inequality. Therefore, the values are acceptable.

Section 18: Residence time calculation

The following equation can be used to calculate h_{bc}.

$$h_b = \frac{h_{bc}}{\phi_{dc}}$$

Where ϕ_{dc} is the average relative density = 0.655

Table 22: hbc calculations

	Plate	hb(mm liquid)	hbc(mm liquid)
Rectifying section	1	200.8577543	131.5618291
	2	200.1967765	131.1288886
	3	200.3508631	131.2298153
	4	200.4134595	131.2708159
	5	200.5450461	131.3570052
	6	200.7036423	131.4608857
	7	200.7756993	131.508083
Feed	8		
Stripping Section	9	211.9046542	138.7975485
	10	211.9393171	138.8202527
	11	211.9715919	138.8413927
	12	212.0400778	138.886251
	13	212.0759291	138.9097335
Reboiler	14		

The following equation was used to calculate residence time.

$$t_r = \frac{A_d h_{bc} \rho_L}{L_{wd}}$$

Table 23: Residence time calculations

	Plate	hbc(mm liquid)	hbc(m)	Ad (m2)	$\rho_{L,i}$ (kg/m^3)	L (kg/s)	tr(s)
Rectifying section	1	131.5618291	0.131561829	0.061815172	617.1799011	0.963724763	5.208153064
	2	131.1288886	0.131128889				5.191014202
	3	131.2298153	0.131229815				5.195009598
	4	131.2708159	0.131270816				5.196632695
	5	131.3570052	0.131357005				5.200044678
	6	131.4608857	0.131460886				5.204157008
	7	131.508083	0.131508083				5.206025414
Feed	8						
Stripping Section	9	138.7975485	0.138797549	0.07311048	615.967214	2.390523149	3.024725313
	10	138.8202527	0.138820253				3.025153024
	11	138.8413927	0.138841393				3.025551267
	12	138.886251	0.138886251				3.026396326
	13	138.9097335	0.138909734				3.0268387
Reboiler	14						

Section 19: Efficency Calculation and Actual Number of Plates.

The following equation was applied to calculate the relative volatility (α)

$$\alpha = \frac{y_A / (1 - y_A)}{x_A / (1 - x_A)} = \frac{y_A . (1 - x_A)}{x_A . (1 - y_A)}$$

$$E = 51 - 32.5 \log (\mu_a \alpha_a)$$

Table 24: Efficiency calculations

Plate	Temperature	n-hexane composition in	n-heptane composition	Viscosity n-hexane	Viscosity n-heptane	viscosity (mPas)	n-hexane composition	n-heptane composition	alpha value	avg liq viscosity	avg alpha	Efficency
1	71	0.88	0.12	0.00019662	0.000249382	0.030098878	0.95	0.05	2.590909091	0.141149848	2.440796957	66.04060775
2	73.5	0.77	0.23	0.000192702	0.000243803	0.056222978	0.896	0.104	2.573426573			
3	76.8	0.631	0.369	0.000187921	0.000236994	0.087569327	0.812	0.188	2.525778063			
4	80.7	0.489	0.511	0.000181434	0.000227754	0.116471066	0.704	0.296	2.485381087			
5	84.2	0.375	0.625	0.000176056	0.00022009	0.137622146	0.596	0.404	2.458745875			
6	86.8	0.298	0.702	0.000172559	0.000215105	0.151054782	0.508	0.492	2.432312981			
7	88.3	0.252	0.748	0.000169981	0.000211428	0.158191129	0.448	0.552	2.409017713			
8	89.2	0.226	0.774	0.00016913	0.000210215	0.162744246	0.413	0.587	2.409597323			
9	90.3	0.195	0.805	0.00016744	0.000207804	0.167314951	0.367	0.633	2.393445943			
10	91.7	0.159	0.841	0.000165767	0.000205416	0.172781549	0.311	0.689	2.38748163			
11	93.1	0.122	0.878	0.000163286	0.000201876	0.1772674	0.247	0.753	2.360677509			
12	94.8	0.085	0.915	0.000161651	0.000199544	0.182596134	0.18	0.82	2.362984218			
13	96.1	0.052	0.948	0.000159228	0.000196084	0.185896291	0.115	0.885	2.368970013			
14	97.2	0.024	0.976	0.000158428	0.000194942	0.190266999	0.056	0.944	2.412429379			

Step by step using above equations relative volatility and liquid viscosity for each tray were calculated and average values obtained.

Efficency of the column = 66.04%

To calculate the actual number of stages, the equation E = $\frac{number\ of\ ideal\ stages\ x\ 100}{number\ of\ actual\ stages}$ was used.

$$\begin{aligned} &Column\ Height \\ &= Distance\ between\ topmost\ and\ bottommost\ trays + Total\ allowance \end{aligned}$$

$$\begin{aligned} &Distance\ between\ topmost\ and\ bottommost\ trays \\ &= (No.of\ actual\ trays - 1) \times Tray\ Spacing \end{aligned}$$

$$\begin{aligned} &Allowance\ for\ each\ top\ or\ bottom \\ &= 15\%\ of\ Distance\ between\ topmost\ and\ bottommost\ trays \end{aligned}$$

Now according to these equations step by step we calculated below values.

No. of actual stages=	21.19907808
	22
No.of actual trays	21
Tray Spacing (m)	0.45
Distance between first and last plate (m)	9
Allowance for each top and bottom(m)	1.35
Total Allowance(m)	2.7
Column Height(m)	11.7

Column height equals to 11.7m

(1)- https://archive.nptel.ac.in/content/storage2/courses/103103027/pdf/mod7.pdf

(2)- https://aiche.onlinelibrary.wiley.com/doi/abs/10.1002/aic.690070414

(3)- https://ndl.ethernet.edu.et/bitstream/123456789/53196/1/Coulson_%26_Richardson%27s_Chemical_Engineering%2C_Volume_6.pdf