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B.sc Degree In Refining And Transportation Engineering

Project title:

Design of Binary Distillation Column for Benzene-Toluene mixture

تصميم عمود تقطير ثنائي بنزين-تولوين

Prepared by:

1. Aanab Ismail mohammedsaeed
2. Fatima mohammed Alimohammed
3. Sabreenyousifsuliman

Supervisor:

Dr.Musa A. Garalnabi

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The project include:

Subject	Page number
الاستهلال	i
الاهداء	ii
الشكر والعرفان	iii
التجريد	iv
Abstract	V
Content	vi
Figures list	xi
Tables list	xii
Symbols list	xiii

الإستهلال

قال تعالى:

﴿اقرأ باسم ربك الذي خلق (1) خلق الإنسان من علق (2) اقرأ
وربك الأكرم (3) الذي علم بالقلم (4) علم الإنسان ما لم يعلم
(5)﴾

صدق الله العظيم

سورة العلق

الايه(1-5)

الإهداء

إلى ... أمي الغالية

التي ارضعتني الحب ممزوجا بالعنان والدفع والطيبه

إلى ... أبي العزيز

الذي حفني برعايته وعطفه وقادني إلى الطريق القويم

ووقف بجانبني في كل مراحل الدراسة

إلى كل من ساهم في تحقيق ما نرناه إليه من أساتذة وزملاء

نهديهم جهدنا المتواضع.

ونضع بين أيديكم ما نتمنى أن يجد القبول والرضاء

مع الإخلاص دوما

الشكر والعرفان الشكر والعرفان

لله نور السماوات والأرض

الحمد لله بادئ القول حمدا كثيرا طيبا مبارك فيه

أفضل الناس في الورى رجلا

تقضى على يده للناس حاجات

أجزله الأستاذ/

د. موسى عبد الله جار النبي

وكافة الأساتذة بقسم هندسة النقل والتكرير

ولكل من ساهم بفكره وتشجيعه لإنجاز هذا العمل المتواضع

وفتكم الله وسدد خطاكم

التجريد

في هذا المشروع تم تصميم عمود تقطير لفصل خليط ثنائي من البنزين والتولوين و الوصول الى نقاوه بنسبه 95% باستخدام طريقه مكايي ذيبيل وهي طريقه رسم بياني مع استخدام بعض المعادلات التحليليه كمعادله انتونيون وقد كانت نتيجته البحث ان عدد الصواني النظري اثنا عشر صنيه ونسبه لان الكفاءه 75% فان العدد الفعلي للصواني خمس عشر صنيه وان افضل وضع للتغذيه هو في الصنيه رقم خمس وان قطر العمود 7.5 متر و ارتفاعه 9 متر كما موضح بالرسم والمعادلات التحليليه.

Abstract:

This project was carried out to design a distillation column for separating a binary mixture of benzene and toluene using McCabe Thiele's procedure. It determined the number of trays required for the desired degree of 95% purity. A number of analytical equations, such as Antoine equation were used. Some other column parameters such as the height and diameter of the column were also determined. The separation is achieved by using sieve tray with 75% overall column efficiency. Finally, the conclusion showed that the distillation of benzene-toluene mixture required 12 theoretical sieve trays and a distillation column of 9 meters high and 7.5 meters in diameter.

Key words:

Distillation, feed, efficiency, material and energy balance,

Reflux, stages, duty, flooding, weeping, entrainment, foaming, liquid backup.

project including:

الاستهلال.....	i
الاهداء.....	ii
الشكر والعرفان.....	iii
التجريد.....	iv
Abstract.....	V
Key word.....	vi
Content.....	x
List of Figures.....	xi
List of tables.....	xii
List of Symbols.....	xiii

Contents

Chapter 1

Introduction to distillation

1.1. Separation process.....	1
1.2. Distillation separation process requirements.....	2
1.3. Limitations.....	3
1.4. Objective.....	3

Chapter 2

Literature review.....	4
2.1.Trays or plates.....	5
2.1.1. Tray efficiency.....	5
2.1.2. Advantages of trays.....	6
2.2. Reflux considerations.....	6
2.3 Distillation principles.....	6
2.4. Dew point and bubble point.....	8

Chapter 3

The design Methodology.....	9
3.1.McCabe and Thiele method.....	9
3.1.1. Material and energy balances.....	9
3.1.2. The Feed Line (q-line).....	13
3.1.3. Calculation of minimum reflux ratio (R_{min}).....	14
3.1.4. Calculation of actual reflux ratio.....	15
3.1.5. Minimum number of stages	
3.1.6. Calculation the number of stages.....	16
3.2. Calculation location of feed.....	17
3.3. Calculate the column diameter.....	18
3.4. Check Weeping.....	21
3.5. Weir liquid crest.....	21
3.6. Check Entrainment.....	22
3.7. Weir dimensions.....	24
3.8. Hole size.....	25

3.9. Hole pitch.....	25
3.10. Plate Pressure Drop (P.D).....	26
3.11. Downcomer Liquid Backup.....	28
3.12. Downcomer residence time.....	29
3.13. Numbers of Holes.....	30
3.14. Height of Distillation Column.....	30

chapter4.

The specification and calculations of the Design.....31

4.1 .design specifications.....	31
4.2. Assumptions made during the design.....	31
4.3. Design procedure.....	32
4.4. Construct X-Y diagram.....	32
4.5. Physical properties.....	35
4.6. Design calculations.....	38
4.6.1. Flow sheet.....	38
4.7. Minimum reflux ratio R_{min}	39
4.8. Minimum number of stage.....	39
4.9. Actual reflux ratio.....	40
4.10. Theoretical number of trays.....	41
4.11. Determining the location of feed.....	42
4.12. Condenser and reboiler duties.....	42

4.13. Calculation the column diameter.....	43
4.14. Plate layout.....	44
4.15. Check weeping.....	44
4.16. Plate Pressure Drop (P.D).....	45
4.17. Downcomer liquid back up.....	45
4.18. Downcomer residence time(t_r).....	46
4.19. Check entrainment.....	46
4.20. Number of holes.....	46
4.21. Height of column.....	46

Chapter 5

Materials of construction.....	47
5.1. Materials properties.....	47
5.1.1. Strength.....	47
5.1.2. Hardness.....	47
5.2. Effect of temperature on the mechanical properties.....	47
5.3. Corrosion resistance.....	47
5.4. Selection for corrosion resistance.....	48
5.5. Material costs.....	48
5.6. Contamination.....	49
5.7. Carbon steel.....	49

Chapter 6

discussion and results.....	50
6.1. Discussion of results.....	50
6.2. Column parameters.....	51
6.3. Tray parameters.....	51

Recommendation.....52
References.....53

List of Figures:

Figure 2.1. Boiling point diagram.	7
Figure 3.1 material balances in top section distillation column	10
Figure 3.2 material balances in bottom section of distillation column	11
Figure 3.3 rectifying section with equilibrium curve	12
Figure 3.4 stripping section with equilibrium curve	12
Figure 3.5 five cases of q lines	13
Figure (3.6) McCabe-Thiele diagram for Minimum reflux for a binary system	15
Figure (3.7) McCabe-Thiele diagram for Minimum number of stages for a binary system	16
Figure (3.8) Distillation column efficiencies (after O'Connell, 1946)	17
Figure (3.9) Flooding velocity, sieve plates	19
Figure (3.10) Weep-point correlation	22
Figure (3.11) entrainment correlation for sieve plates	23
Figure (3.12) Relation between downcomer area and weir length	24
Figure (3.13) Relation between hole area and pitch	26
Figure(3.14) Discharge coefficients, sieve plates	27
Figure (4.1)equilibrium curve (X_B versus y_B)	35
Figure (4.2)a typical distillation column	38
Figure (4.3) Minimum reflux ratio	39
Figure (4.4) Minimum number of stage	40
Figure (4.5) theoretical number of stage N_{theo}	41

List of Tables:

Table (4.1) the design specifications	31
Table (4.2) The constants A, B ,C for this antonion equation	33
Table (4.3) T-X-Y data for benzene	34
Table 5.1. The price of some metals	49
Table 6.1 the column parameters	51
Table 6.2 the tray parameters	51

List of Symbols:

E_0	Tray efficiency
y^*	Composition of vapor in equilibrium with the liquid at point n
y_n	Actual vapor composition at that point
A	Relative volatility
X	Mole fraction in liquid phase
Y	Mole fraction in gas phase
q_c	Heat removed in the condenser
H	Enthalpy
R	Reflux ratio
R_{min}	Minimum reflux ratio
N _{th}	Number of theoretical plates
N _{ac}	Actual number of stages
E	Overall column efficiency
α_{avg}	Average relative volatility of more volatile component mNs/m ²
μ_{avg}	Molar average liquid viscosity of feed evaluated at average temperature of column
N_D	Number of Plates above the feed tray
N_B	Number of Plates below the feed tray
B	Bottom flow rate
D	Distillate flow rate
x_{Hk}	Mole fraction of heavy component
x_{Lk}	Mole fraction of light component
X_{LkB}	<i>Mole fraction of light component in the bottom</i>
X_{HKD}	<i>Mole fraction of heavy component in the distillate</i>
u_f	Flooding vapour velocity, m/s
ρ_l	<i>Liquid density kg/m³</i>
ρ_v	<i>Vapor density kg/m³</i>
F_{LV}	<i>Liquid – vapor flow factor</i>
L_w	<i>Liquid mass flow rate, kg/s</i>
V_w	<i>Vapor mass flow rate, kg/s</i>
Q _v	Volumetric flow rate of vapors
A _n	Net column area
V_n	Maximum vapor velocity
A_c	<i>Actual area or column area m²</i>
A_n	<i>Net area, in m²</i>
A_d	<i>Downcomer area, m²</i>
D	<i>Dieter of column, m</i>
\check{u}_{min}	Minimum vapour velocity through the holes, m/s.
d_h	Hole diameter, mm
l_w	Weir length, m
h_{ow}	Weir crest, mm liquid

L_w	Liquid flow – rate, kg/s
u_n	Actual velocity, (m/s)
u_f	Flooding velocity
A_h	Hole area
A_p	Perforated area
l_p	Holepitch(distancebetweentheholecentres)
A_a	Active area
h_d	Pressure drop through the dry plate
C_o	Orifice coefficient
u_h	Velocity through the holes
h_r	Residual head
h_t	Total plate pressure drop
h_w	<i>Weirheight</i>
h_b	Downcomer back-up, measured from plate surface, mm
h_{dc}	Head loss in the downcomer, mm
h_t	<i>Totalplatepressuredrop</i>
h_{ow}	<i>Weircrest, mliquid</i>
l_{wd}	The liquid flow rate in downcomer
A_{ap}	Clearance area under the downcomer
h_{ap}	Height of the bottom edge of the apron above the plate
t_r	Residence time, s
h_{bc}	Clear liquid back-up, m
l_{wd}	The liquid flow rate in downcomer
H_c	Height of Distillation Column
H_s	<i>Trayspacing, m</i>
L	The Liquid flow rate from the top section <i>kmol/hr</i>
\dot{L}	Liquid flow rate from the bottom section <i>kmol/hr</i>
G	Vapor flow rate in the column
V_w	Vapor mass flow rate
L_w	Liquid mass flow rate

Chapter 1

Introduction

Chapter 1

Introduction to distillation

Introduction:

This project discusses the design of distillation column to separate a binary mixture of benzene (C_6H_6) and toluene ($C_6H_5CH_3$). Benzene and toluene are aromatics produce from different source such as Primary sources of aromatics are from refinery catalytic reformers, pyrolysis gasoline from olefins plants, Secondary sources include toluene disproportionation (TDP) and toluene hydrodealkylation (THDA) units.

About 70% of the global production of benzene is by extraction from either reformat or pyrolysis gasoline.

1.1. Separation process:

The separation of benzene from toluene is done for many reasons :

- i. The Existence benzene and toluene as mixture is not required because we use benzene and toluene apart.
- ii. also toluene has harmful effects. The methyl group makes it around 25 times more reactive than benzene in reactions.
- iii. Benzene has harmful effects too (its causes cancer).

The separation process is achieved to produce high purity of benzene that it has important uses in chemical industry some of there uses are:

- a. Benzene is used mainly as an intermediate to make other chemicals used in the production of more than 250 different products.
- b. About 80% of benzene is consumed in the production of three chemicals, ethyl benzene, cumene, and cyclohexane ,and Smaller amounts of benzene are used to make some types of rubbers, lubricants, dyes, detergents, drugs, explosives, and pesticides(9). Toluene also is produced from the separation it is used as solvent to dissolve paints, and as an octane booster in gasoline fuels used in internal combustion engines, Toluene is used as a jet fuel surrogate for its content of aromatic compounds.(9)

1.2. Distillation separation process requirements:

The separation process requires three things:

- A. Firstly, both liquid and vapor phases are present and in contact with each other on each stage within the separation column .
- B. Secondly, the components must have different volatilities where the higher difference the easier the separation process.
- C. Thirdly, the two phases can be separated by gravity or by mechanical means. (2)

Distillation is based on the fact that the vapor of a boiling mixture will be richer in the components that have lower boiling points. There are many types of distillation columns, each one of them is designed to be used in specific kind of separation.

Depending on how they are operated they can be classified to: Continuous or Batch distillation columns.

In continuous columns processes, a continuous feed stream is entered to the column. This feed can contain two components in this case called binary mixture, if the feed contains more than two components it is called multi component mixture.

In designing a column for a given separation, the number of stages required and the Flow rates of the liquid and vapor streams must first be determined using the general methods. In the mechanical design of the column, tower diameter, tray spacing, and the detailed layout of each tray is considered. Initially, a diameter is established, based on the criterion of absence from liquid entrainment in the vapor stream, and then the weirs and the downcomers are designed to handle the required liquid flow. It is then possible to consider the tray geometry in more detail, and, finally, to examine the general operating conditions for the tray and to establish its optimum range of operation.

In this work we applied a designing procedure for a distillation column to separate binary mixture (benzene toluene) by using McCabe Thiele which is the simplest graphical method for calculation the number of stages in the distillation column.

1.3. Limitations:

- Azeotropes: An azeotrope is a liquid mixture which when vaporized produces the same composition as the liquid. If the mixture is azeotropic, then more advanced types of separation must be considered.
- Solids: If the material to be separated is high in solids, or contains tars or resins that could plug or foul a continuous unit, then a batch separation should be considered (Perry's, 1997).

- **1.4. Objective:**

Design a distillation column to separate mixture of benzene and toluene by using McCabe Thiele because Benzene and toluene are part in chemical industry

Chapter 2

Literature review

Chapter 2

Literature review:

Distillation is one of the oldest unit operations. While the first technical publication Was developed in 1597, distillation already had been practiced for many Centuries specifically, for the concentration of ethyl alcohol for beverages. Today, distillation is one of the most used unit operations and is the largest Consumer of energy in the process industries.(1)

Distillation is a separation process used to separate two or more components into an overhead distillate and bottoms where the bottoms product is liquid, and the distillate may be liquid or a vapor or both.

Now, we take continues distillation column to separate binary mixture of benzene-toluene as case study, Distillation columns are made up of several components each of which is used to transfer heat energy or enhance material transfer a typical distillation contains several major components they are:

- Vertical shell
- Column internals (trays or packing)
- Reboiler: it is divided into two types according to extent of boil,
 - ❖ Partial reboiler only partially vaporize the bottom liquid coming in.
 - ❖ Total reboiler would completely vaporized stream.
- Condenser : it is differentiated by extent of condensation
 - ❖ Total condenser, which all of the vapor leaving the top of the column is condensed. Consequently, the composition of the vapor leaving the top tray is the same as that of liquid distillate product and reflux X_D .
 - ❖ Partial condenser, which the vapor is only partially liquefied. The liquid produced is required to the column as liquid and vapor product stream is removed.
- Reflux drum: used to hold up of liquid from condenser. (Atypical distillation column is shown in figure(4.2))

2.1. Trays or plates:

Trays are used inside the distillation column for vapor liquid equilibrium and efficiency of the column, there are many types of trays:

- Bubble cap trays.
- Valve trays.
- Sieve trays.

If we assume the trays are properly designed installed and operated, Bubble cap, Valve Sieve trays have comparable capacity, efficiency, entrainment and pressure drop the bubble cap tray have lower capacity and efficiency and high entrainment and pressure drop than sieve and valve trays. The turn down of valve trays is much better than sieve trays, but not as good as bubble cape trays. Bubble cape trays are most suitable to handle low liquid rate applications, the cost of bubble cap trays is by far the highest, sieve trays are the least expensive, but valve trays are only slightly more expensive.(3)

2.1.1. Tray efficiency:

Three principal definitions of efficiency are used:

1. Overall column efficiency is defined by equation (2.1).

$$E_0 = \frac{\text{number of ideal stages}}{\text{number of real stages}} \quad (2.1)$$

Once the tray efficiency is known the number of actual tray can be obtained from equation (2.1). (4)

2. Point efficiency: is the ratio of change of composition at a point to the change that would accure on a theoretical stage. Point efficiency defined by equation(2.2).(3)

$$E_{OG} = \frac{y_n - y_{n-1}}{y^* - y_{n-1}} \quad (2.2)$$

y^* Is the composition of vapor in equilibrium with the liquid at point n. y_n is the actual vapor composition at that point.

3. Murphree plate efficiency: is the same as point efficiency, except that it applies for the entire tray instead of to a single point. And is defined by same eq.(2.2)where y^* is the composition of vapor in equilibrium with the liquid

that leaving the tray and y_n is the actual vapor composition leaving the tray.(3).

Tray efficiencies are affected by fouling, wear, tear and corrosion, and the rates at which these occur depends on the properties of the liquids being processed. Thus appropriate materials should be specified for tray construction.

2.1.2. Advantages of trays:

- Column can be handle wide range of gas and liquid flow rate.
- Plate efficiency can be predicted more accurately.
- Plate column operation is considerably smooth.
- Cooling arrangement can be provided in plate column.
- Plate column provide stage wise contact.
- When liquid cause fouling or deposition of solid even through the tray is blocked up It is easier to clean and manholes can be provided over the plate

2.2 Reflux considerations:

Reflux refers to the portion of the condensed over head liquid product from a distillation tower that is returned to the upper part of the tower. There are three type of reflux: Total reflux, Minimum reflux, Optimum reflux.(4)

2.3 Distillation principles:

Separation of components from a liquid mixture via distillation depends on the differences in boiling points of the individual components. Also, depending on the concentrations of the components present, therefore, distillation processes depends on the vapor pressure characteristics of liquid mixtures.

- **K-value and Relative Volatility:**

K-value is measure of the tendency of component i to vaporize.

$$K_i = \frac{\text{mole fraction of component i in vapor phase}}{\text{mole fraction of component i in liquid phase}} \quad (2.3)$$

The K-value is a function of temperature, pressure and composition

The relative volatility of component I and j defined by equation (2.4) (3):

$$\alpha = \frac{k\text{-value of component } i}{k\text{-value of component } j} \quad (2.4)$$

$$\alpha = \frac{K_1}{K_2} = \frac{y_1(1-x_1)}{x_1(1-y_1)} \quad (2.5)$$

This equation can be rearranged to give equation:

$$y = \frac{\alpha x}{1+x(\alpha-1)} \quad (2.6)$$

Where:

$\alpha \equiv$ relative volatility.

$x \equiv$ mole fraction in liquid phase.

$y \equiv$ mole fractio in gas phase.

- **Phase diagram:-**

Phase diagram are use to describe binary system by plotting tow out of three variables: composition, temperature, pressure, at constant value of the remaining one. Consider an example of a liquid mixture containing 2 components (A and B) -a binary mixture. This has the following boiling point diagram. Figure 2.1.

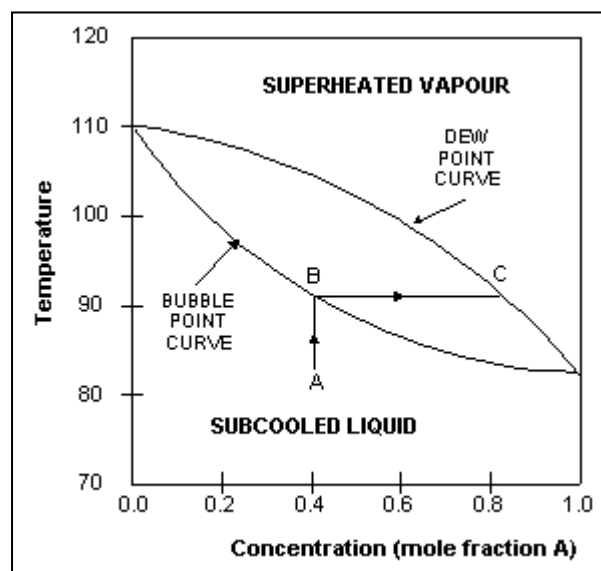


Figure 2.1. Boiling point diagram.

- **Ideal system:-**

An ideal system is one where the vapor obeys the ideal gas law and the liquid obeys Raoult's law, and ideal gas mixture obeys Dalton's law

$$p_A = y_A p \quad (2.7)$$

An ideal solution obeys Raoult's law which states partial pressure of a component in solution is equal to the product of its mole fraction and the vapor pressure of the pure component, thus

$$p_A = x_A p_A^\circ \quad (2.8)$$

From (2.7) and (2.8): $y_A = x_A p_A^\circ / p$ and $y_B = x_B p_B^\circ / p$

But $y_A + y_B = 1$ So: $\frac{x_A p_A^\circ}{p} + \frac{x_B p_B^\circ}{p} = 1$

Giving: $X_A = \frac{p - p_B}{p_A - p_B}$ (2.9)

From equation (2.5) and (2.6) and the definition of K-value, one obtains :

$$k_i = \frac{y_i}{x_i} = \frac{p_i^\circ}{p} \quad (2.10)$$

2.4. Dew point and bubble point:

- **dew point:**

The temperature at which the saturated vapor starts to condense.

Dew point of mixture is calculated from:

$$\sum_{i=1}^c \frac{y_i}{k_i} = 1 \quad (2.11)$$

- **bubble point:**

The temperature at which the liquid starts to boil.

Bubble point of mixture is calculated from:

$$\sum_{i=1}^c x_i k_i = 1 \quad (2.12)$$

Chapter 3

Methodology

Chapter 3

methodology

The design of multistage columns can be accomplished by graphical techniques when the feed mixture contains only two components. The x - y diagram method Developed by McCabe and Thiele uses only phase equilibrium and mole balance relationship. However, the approach is simple and gives a useful first Estimate of the column design which can be refined by using the enthalpy composition diagram method of Ponchon and Savarit. (5)

3.1. McCabe and Thiele method:

Operating Lines the McCabe-Thiele method is based upon representation of the material balance equations as operating lines on the y - x diagram. The lines are made straight by the assumption of constant molar overflow, The liquid-phase flow rate is assumed to be constant from tray to tray in each section of the column between addition (feed) (product) points. It assumes constant molar overflow and this implies that:

- The molar heats of vaporization of the feed components are equal.
- For every mole of liquid vaporized, a mole of vapor is condensed.
- Heat effects such as heats of solution are negligible.

3.1.1. Material and energy balances:

As mentioned the operating lines for rectifying and stripping sections is constructed from a material balance around each section (top and bottom sections).

I. Material and energy balance around top section:

From figure 3.1 below take the top section that bounded by scattered line and made material and energy balances around it:(5)

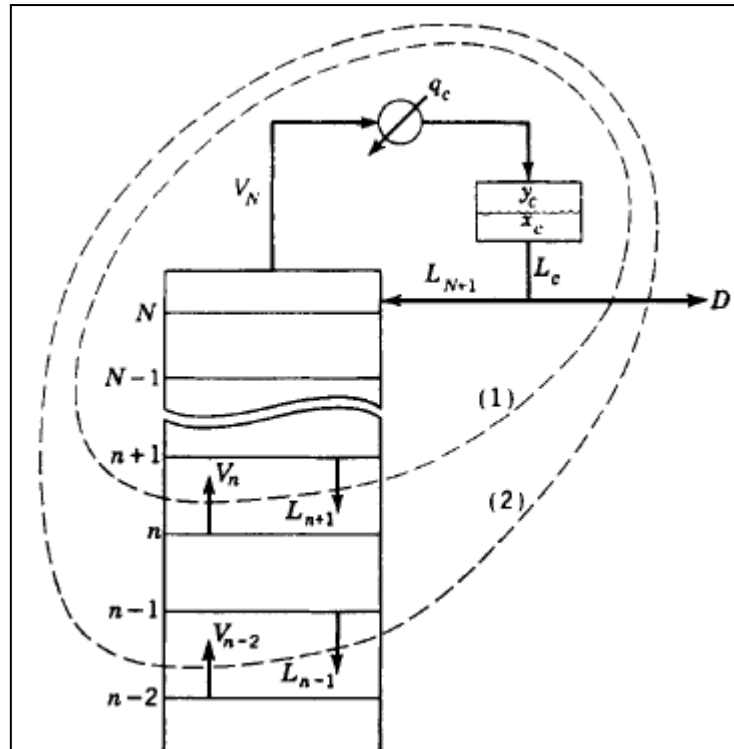


Figure 3.1 material balances in top section of distillation column (5)

- **Material balances**

Total flow:
$$v_n = l_{n+1} + D \quad (3.1)$$

For either component:
$$v_n \cdot y_n = L_{n+1} x_{n+1} + D x_d \quad (3.2)$$

Combining equations 3.1 & 3.2 gives:

$$y_n = \frac{L_{n+1}}{L_{n+1} + D} x_{n+1} + \frac{D}{L_{n+1} + D} x_d \quad (3.3)$$

If the liquid flow rate L_{n+1} is assumed to be identical to L_{n-1} , then $v_n = v_{n-1}$ and take equations 3.1 & 3.3 we come up to:

$$y_n = \frac{L}{v} x_{n+1} + \frac{D}{v} x_d \quad (3.4)$$

- **Energy balance:**

Total enthalpies:
$$H_n v_n = L_{n+1} h_{n+1} + D H_d + q_c \quad (3.5)$$

Where q_c is the heat removed in the condenser.

Combining equations 3.1 & 3.5 gives:

$$H_n v_n = (L_{n+1} + D)H_{n+1} = L_{n+1} h_{n+1} + DH_d + q_c \quad (3.6)$$

Where y and x have a stage subscript n or $n + 1$, but L and V need be identified only with the section of the column to which they apply. Equation (3.4) has the analytical form of a straight line where L/V is the slope and DxD is the y intercept at $x = 0$.

Also equation 3.4 can be written in terms of reflux ratio as follow:

$$y_n = \frac{R}{R+1} x_d + \frac{x_d}{1+R} \quad (3.7)$$

II. Material and energy balances around bottom section:

Similar equations can be written for the bottom section of the column.

$$y_m = \frac{L'}{v'} x_{m+1} - \frac{B}{v'} x_B \quad (3.8)$$

$$h_n l_n = (v_{n+1} + B)h_{n+1} = v_{n+1} H_{n+1} + BH_b - q_r \quad (3.9)$$

Where the subscript m is used to identify the stage number in the bottom Section, and q_r is the Reboiler duty.

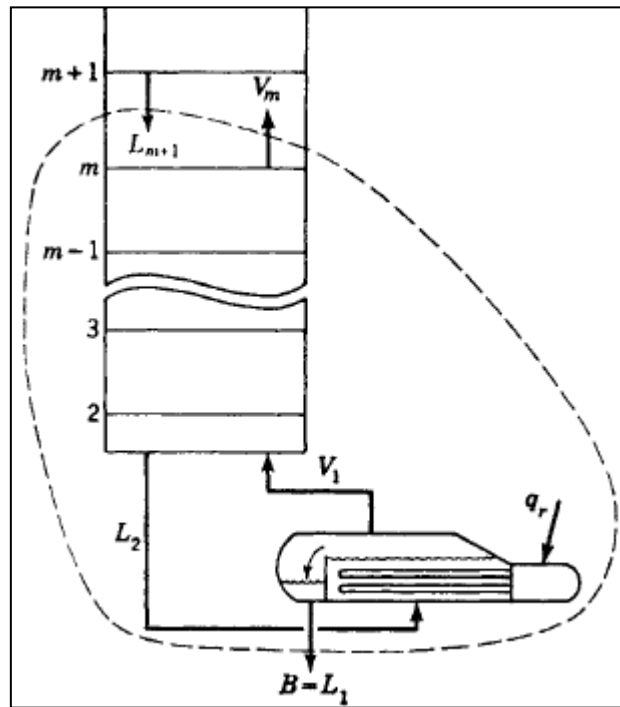


Figure 3.2 material balances in bottom section of distillation column (5)

- Equations 3.4 & 3.7 are used to construct the rectifying operating line as follow:

First the desired top product composition is located on the VLE diagram, and a vertical line produced until it intersects the diagonal line that splits the VLE plot in half. A line with slope $R/(R+1)$ is then drawn from this intersection point as shown in the diagram 3.3 below.

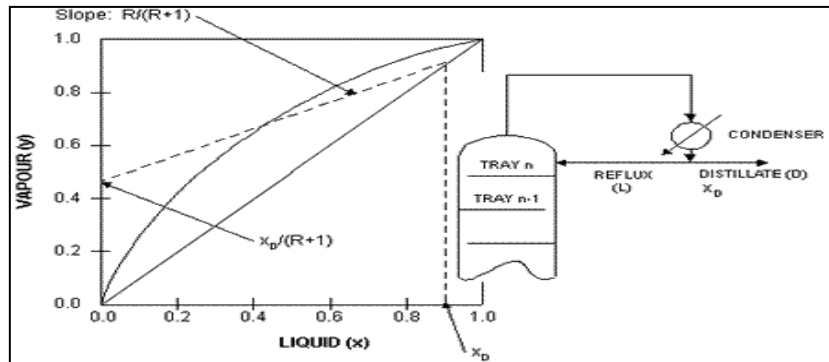


Figure 3.3 rectifying section with equilibrium curve

- The operating line for the stripping section is constructed using equation (3.8)

In a similar manner. However, the starting point is the desired bottom product composition. A vertical line is drawn from this point to the diagonal line, and a line of slope L'/V' is drawn as illustrated in the diagram (3.4) below.

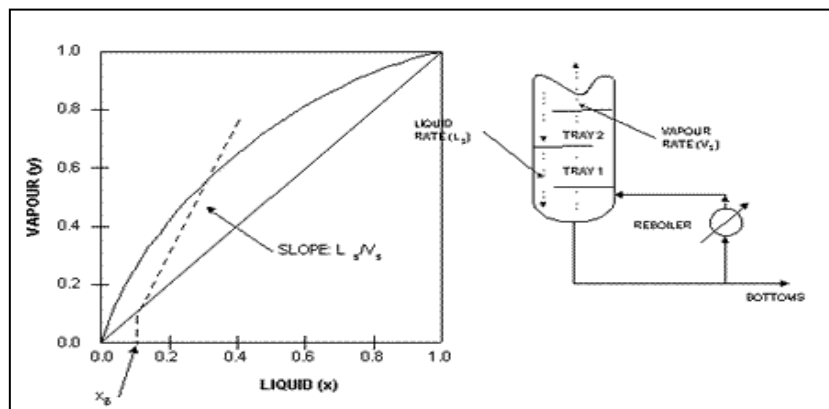


Figure 3.4 stripping section with equilibrium curve

3.1.2. The Feed Line (q-line):

The point of intersection of the two operating lines is dependent on the phase condition of the feed. The line on which the intersection occurs is called the q line. The q line is found as follows:

$$q = \frac{\text{heattovaporise 1moloffeed}}{\text{molarlatenheatoffeed}} \quad (3.10)$$

Also defined by the following equations for a feed stream F:

$$L' = L + qF \quad (3.11)$$

$$v = v' + (1 - q)F \quad (3.12)$$

Subtracting one equation from other gives for feed stage:

$$(v - v')y = (L - L')X + Fz_f \quad (3.13)$$

Which when combined with Eqs3.10 &3.11 gives the q line equation:

$$y = \frac{q}{q-1}x - \frac{z_f}{q-1} \quad (3.14)$$

Where $\frac{q}{q-1}$ is the slope of q line.

Depending on the state of the feed, the feed lines will have different slopes. For example:

- $q=0$ (saturated vapor)
- $q=1$ (saturated liquid)
- $0 < q < 1$ (mix of liquid and vapor)
- $q > 1$ (sub cooled liquid)
- $q < 0$ (superheated vapor)

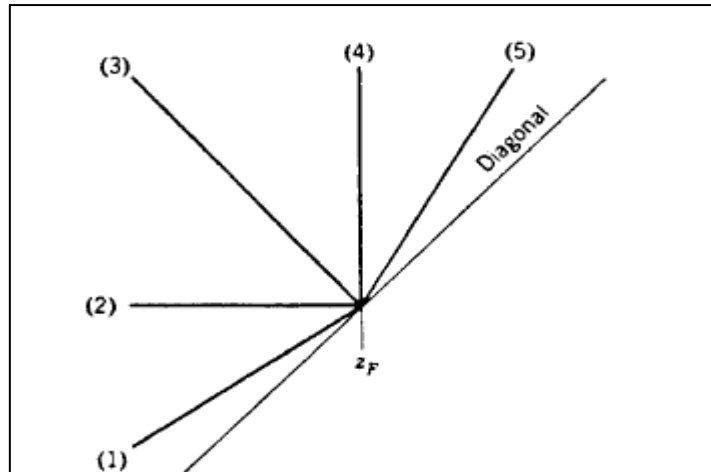


Figure 3.5 five cases of q lines (5)

Figure 3.5 all five cases of q lines: (1) superheated vapor feed, (2) saturated vapor feed, (3) partially vaporized feed, (4) saturated liquid feed, and (5) sub cooled liquid feed.

If we have information about the condition of the feed mixture, then we can construct the q-line and use it in the McCabe-Thiele design. However, excluding the equilibrium line, only two other pairs of lines can be used in the McCabe Thiele procedure. These are:

- Feed-line and rectification section operating line
- Feed-line and stripping section operating line
- stripping and rectification operating lines

This is because these pairs of lines determine the third.

3.1.3. Calculation of minimum reflux ratio (R_{min}):

The separation is theoretically possible if the operating lines intersect at point just below the equilibrium curve, the corresponding reflux ratio is termed minimum reflux ratio the separation at minimum reflux ratio requires infinite number of stages.

To determine minimum reflux construct the q-line and identify its point of intersection with the equilibrium curve. The draw a line from product composition point on the 45° diagonal line to this in intersection point. From equation (3.7) the slope of line is $R_{min}/R_{min} + 1$ and the intercept of this line on the y axis is

$X_d/R_{min} + 1$ minimum reflux can be determined from either of these.

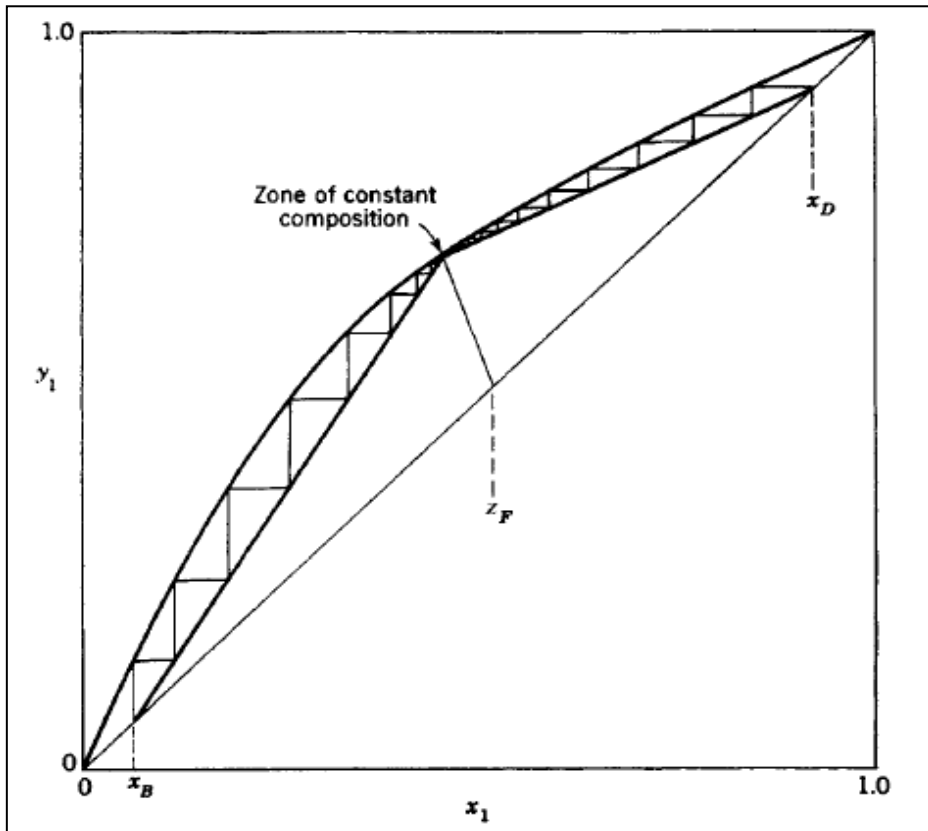


Figure (3.6) McCabe-Thiele diagram for Minimum reflux for a binary system (5)

3.1.4. Calculation of actual reflux ratio:

The rule of thumb is:

$$R = (1.2 \text{ ----- } 2) R_{\min}$$

In our calculations assume $R = 2R_{\min}$

3.1.5. Minimum number of stages:

A column operating at total reflux is represented in Fig. (3.7). enough material has been charged to the column to fill the reboiler, the trays, and the overhead condensate drum to their working levels. The column is then operated with no feed and with all the condensed overhead stream returned as reflux ($L_{N+1} = V_N$ and $D = 0$). Also all the liquid reaching the reboiler is vaporized and returned to the column as vapor. Since F , D , and B are all zero, $l_{n+1} = v_n$ at all points in the column. With a slope of unity ($L/V = 1.0$), the operating line must coincide with the diagonal throughout the

column. Total reflux operation gives the minimum number of stages required to effect a specified separation between x_B and x_D .

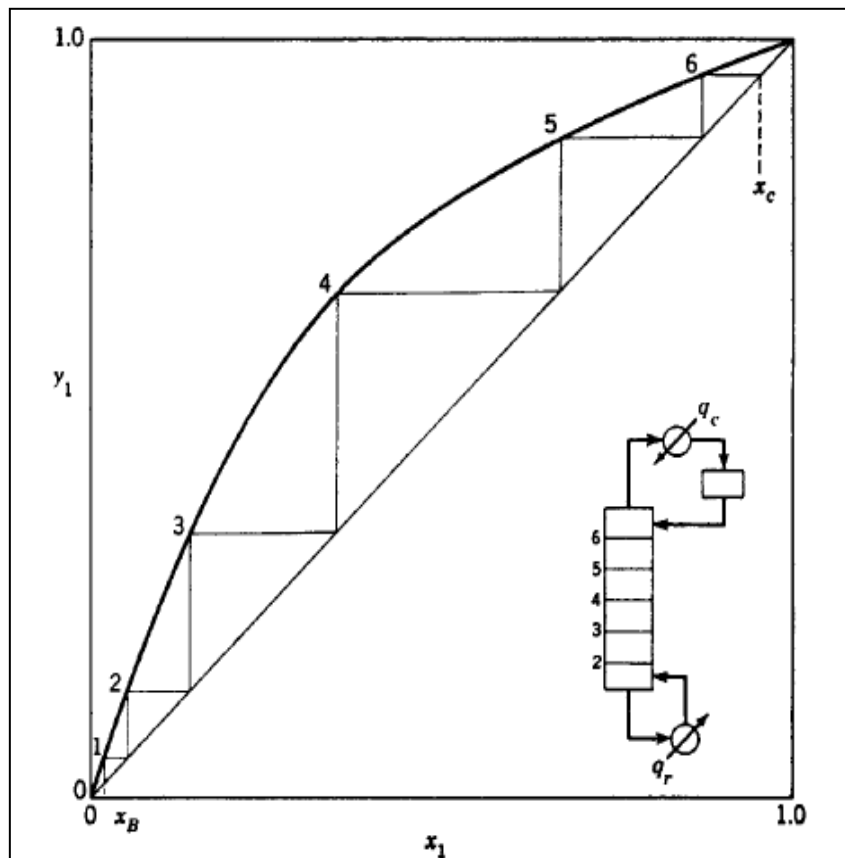


Figure (3.7) McCabe-Thiele diagram for Minimum number of stages for a binary system (5)

3.1.6. Calculation the number of stages:

- **Theoretical number:**

By draw the steps between operating lines and the equilibrium line and count them. Those steps represent the [theoretical plates](#) (N_{th}) (or equilibrium stages).

- **Actual number:**

The actual number of stages is N_{ac} calculated by:

$$N_{ac} = \frac{N_{th}}{E} \quad (3.15) (6)$$

Where $E \equiv$ the overall column efficiency .

the overall column efficiency can be obtained from the correlation given by O'Connell (1946) which is shown in Figure (3.8). The overall column efficiency is correlated with the product of the relative volatility of the light key component and the molar average viscosity of the feed, estimated at the average column temperature.

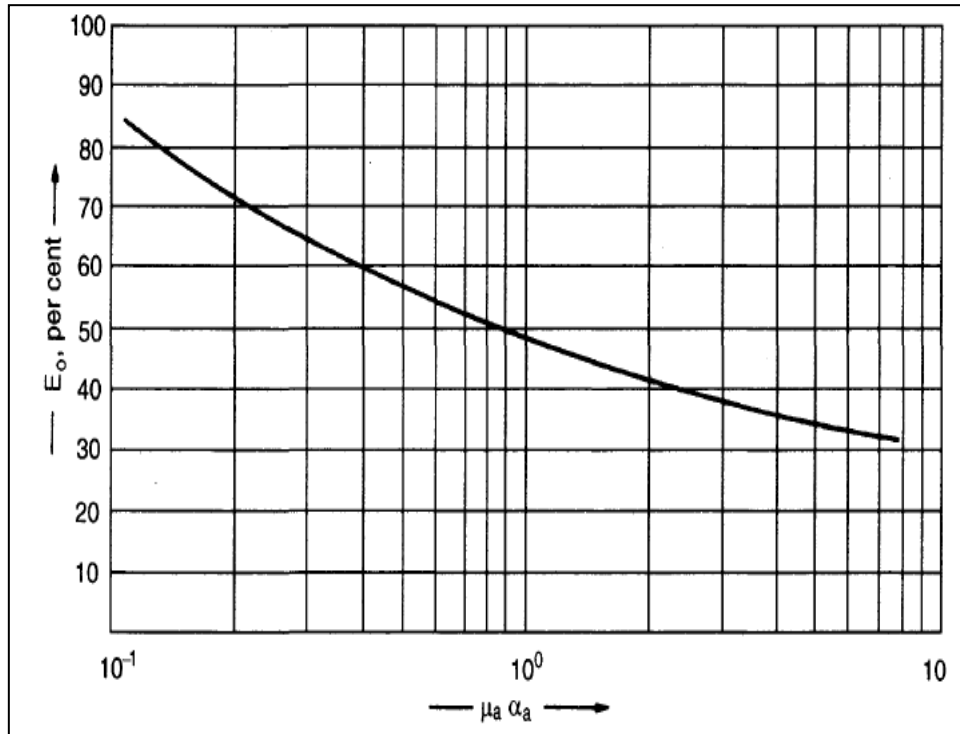


Figure (3.8) Distillation column efficiencies (after O'Connell, 1946) (6)

Eduljee (1958) has expressed the O'Connell correlation in the form of an equation:

$$E_o = 51 - 32.5[\log(\mu_{avg} \alpha_{avg})] \quad (3.16) \quad (6)$$

$\alpha_{avg} \equiv$ Average relative volatility of more volatile component mNs/m².

$\mu_{avg} \equiv$ Molar average liquid viscosity of feed evaluated at average temperature of column

3.2. Calculation location of feed:

The Kirk bride method is used to determine the ratio of trays above and below the feed point the Kirk bride equation is:

$$\log\left(\frac{N_D}{N_B}\right) = 0.2061\left[\left(\frac{B}{D}\right)\left(\frac{x_{Hk}}{x_{lk}}\right)\left(\frac{x_{lkB}}{x_{HkD}}\right)^2\right] \quad (3.17) \quad (6)$$

Where:

$N_D \equiv$ Number of Plates above the feed tray .

$N_B \equiv$ Number of Plates below the feed tray .

$B \equiv$ bottom flow rate.

$D \equiv$ distillate flow rate.

$x_{Hk} \equiv$ mole fraction of heavy component

$X_{lk} \equiv$ mole fraction of light component .

$X_{lkB} \equiv$ mole fraction of light component in the bottom.

$X_{HkD} \equiv$ mole fraction of heavy component in the distillate

3.3. Calculation of the column diameter:

the column diameter must be selected so that that flooding does not occur and the velocity will normally be between 70 to 90 percent of that which would cause flooding.

Flooding is brought about by excessive vapor flow, causing liquid to be entrained in the vapor up the column. The increased pressure from excessive vapor also backs up the liquid in the down-comer, causing an increase in liquid holdup on the plate above. Depending on the degree of flooding, the maximum capacity of the column may be severely reduced. Flooding is detected by sharp increases in column differential pressure and significant decrease in separation efficiency. The flooding vapor velocity is calculates by below equation:

$$u_f = K1 \left(\frac{\rho_l - \rho_v}{\rho_v} \right)^{0.5} \quad (3.18) (6)$$

Where:

$u_f \equiv$ flooding vapour velocity, rn/s.

$\rho_l \equiv$ liquid density.

$\rho_v \equiv$ vapor density.

$K_1 \equiv$ a constant obtained from Figure (3.9) by Assuming the spacing between the trays, normally from 0.15 m (6 in.) to 1 m (36 in.) are used. The spacing chosen will depend on the column diameter and operating conditions.

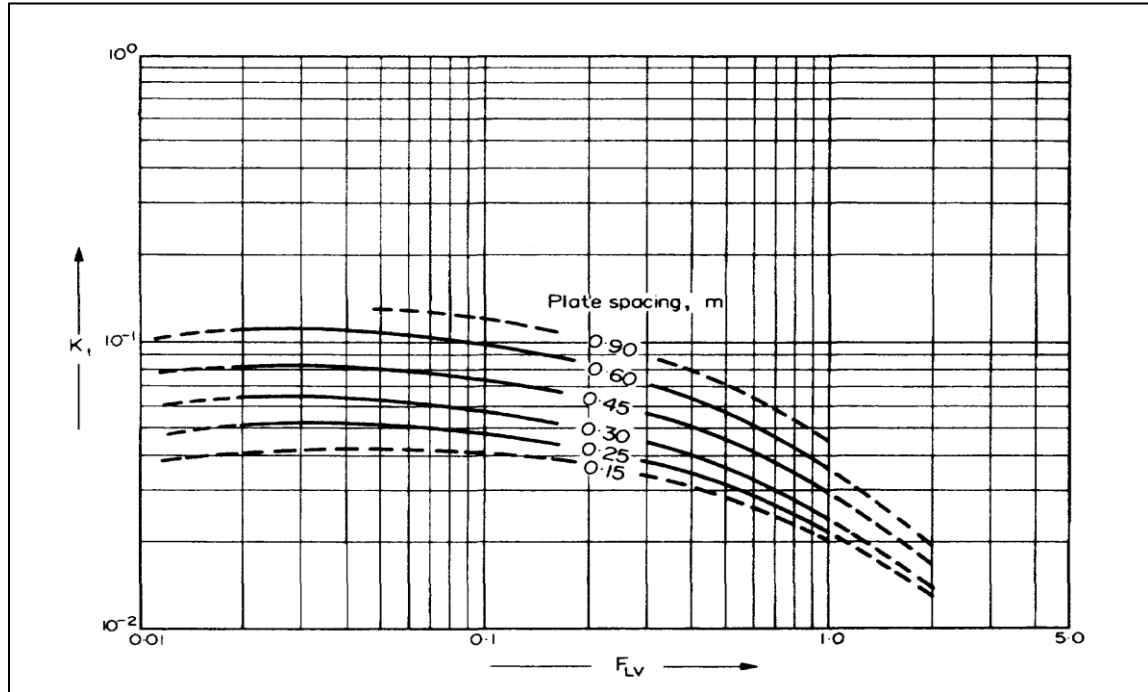


Figure (3.9) Flooding velocity, sieve plates (6)

Flow parameter F_{lv} :

$$F_{lv} = \left(\frac{L_w}{V_w} \right) \left(\frac{\rho_v}{\rho_l} \right)^{0.5} \quad (3.19) (6)$$

$F_{LV} \equiv$ liquid – vapor flow factor.

$L_w \equiv$ liquid mass flow rate, kg/s.

$V_w \equiv$ vapor mass flow rate, kg/s.

$\rho_l, \rho_v \equiv$ density of liquid and vapor kg/m³.

To calculate the column diameter we need to find the net column area A_n and down-comer area A_d since the cross sectional area A_c produce from summation of this area.

- **Calculate Net column area used in separation**

Net column area the total cross-section area less the area at top of the down-comer area and calculated from:

$$A_n = \frac{Q_v}{V_n} \quad (3.20) (6)$$

Where:

$Q_v \equiv$ Volumetric flow rate of vapors.

$A_n \equiv$ Net column area .

$V_n \equiv$ Maximum vapor velocity.

$$Q_v \equiv \frac{(\text{mass vapor flow rate})}{(\text{vapor density})3600} \quad (3.21) (6)$$

Calculate the cross sectional area of the column A_c :

$$A_c = A_n + A_d \quad (3.22) (6)$$

Where:

$$A_c \equiv \text{actual area, } m.^2$$

$$A_n \equiv \text{net area, in } m.^2$$

$A_d \equiv$ downcomer area, $m.^2$ (Downcomers are conduits having circular, segmental, or rectangular cross sections that convey liquid from upper tray to a lower tray in a distillation column.) (7)

A_d Can be taken as percent from a cross-sectional area,

$$A_d = \text{percent} \times A_c \quad (3.23) (6)$$

Where:

$$A_c \equiv \text{column area, } m.^2$$

So, diameter of column is

$$A_c = \frac{\pi D^2}{4} \quad (3.24) (6)$$

$D \equiv$ diameter of column.

3.4. Check Weeping:

Weeping Phenomenon is caused by low vapor flow. The pressure exerted by the vapors insufficient to hold up the liquid on the tray. Therefore, liquid starts to leak through perforations. Excessive weeping will lead to dumping. That is the liquid on all trays will crash (dump) through to the base of the column (via a domino effect) and

the column will have to be re-started. Weeping is indicated by a sharp pressure drop in the column and reduced separation efficiency

The vapor velocity at the weep point is the minimum value for stable operation. The hole area must be chosen so that at the lowest operating rate the vapor flow velocity is still well above the weep point.

The minimum design vapor velocity is given by:

$$\check{u}_{min} = \frac{[K_2 - 0.90(25.4 - d_h)]}{\rho_v^{0.5}} \quad (3.25) (6)$$

$\check{u}_{min} \equiv$ minimum vapour velocity through the holes (based on the hole area), m/s.

$d_h \equiv$ hole diameter, mm.

$\rho_v \equiv$ density of vapour.

$K_2 \equiv$ A constant, dependent on the depth of clear liquid on the plate, obtained from Figure (3.10).

The clear liquid depth is equal to the height of the weir h_w plus the depth of the crest of liquid over the weir h_{ow} .

3.5. Weir liquid crest:

Weir is small dam used to hold up the liquid on the tray. Weir crest is the height of the liquid crest over the weir and can be estimated using the Francis Formula.

$$h_{ow} = 750 \left[\frac{L_w}{\rho_l l_w} \right]^{2/3} \quad (3.26) (6)$$

Where:

$l_w \equiv$ weir length, m.

$h_{ow} \equiv$ weir crest, mm liquid.

$L_w \equiv$ liquid flow – rate, kg/s.

$\rho_l \equiv$ density of liquid.

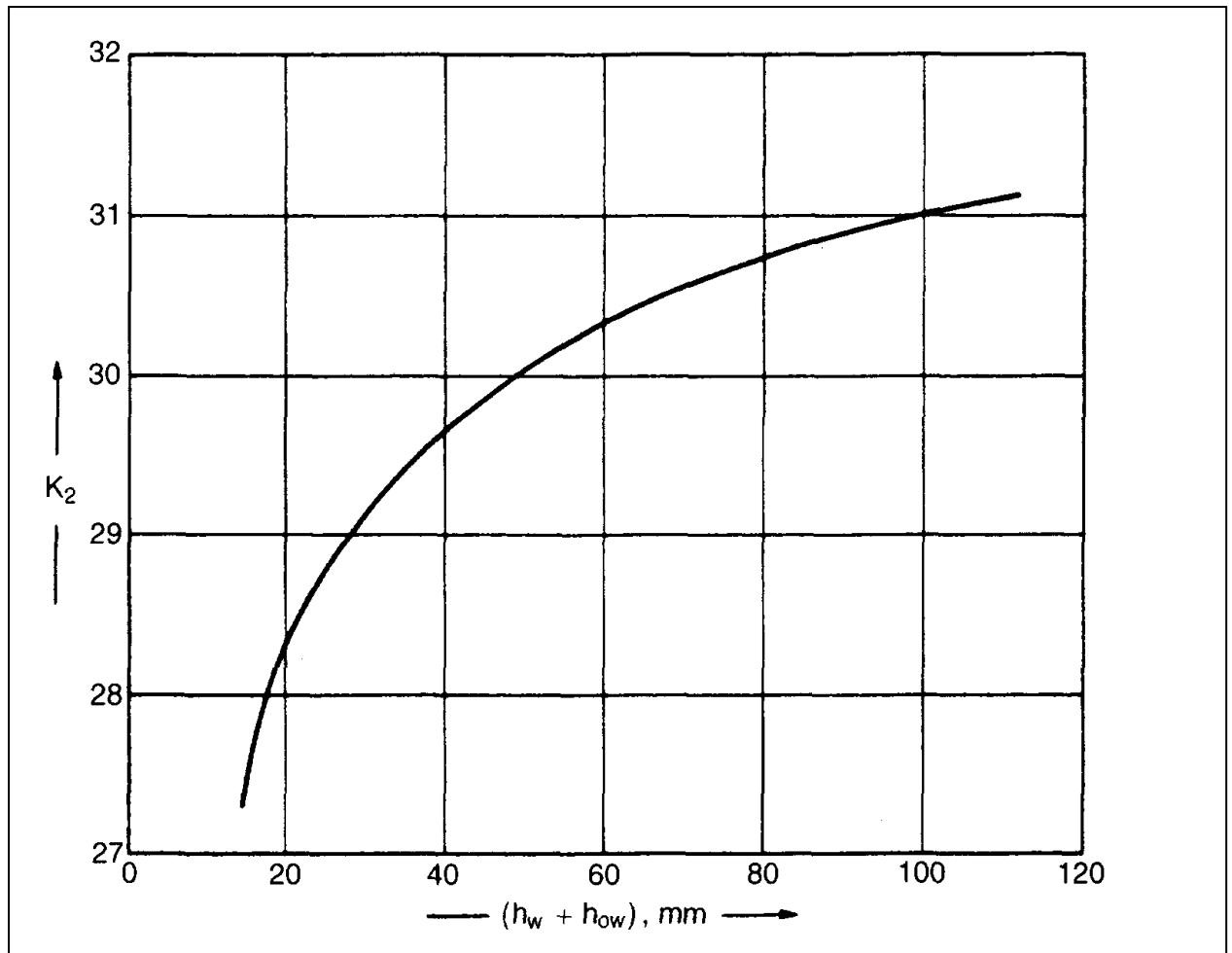


Figure (3.10) Weep-point correlation (6)

3.6. Check Entrainment:

Entrainment refers to the liquid carried by vapor up to the tray above and is again caused by high vapor flow rates. It is detrimental because tray efficiency is reduced. Excessive entrainment can lead to flooding. Entrainment can be estimated from correlation that given by Fair in figure (3.11) which gives the fractional entrainment (ψ) as function of liquid flow parameter with percentage flooding as a parameter.

The percentage flooding given by:

The percentage flooding = u_n / u_f

where:

u_n ≡ is the actual velocity. (m/s).

u_f ≡ flooding velocity from eq(3.19).

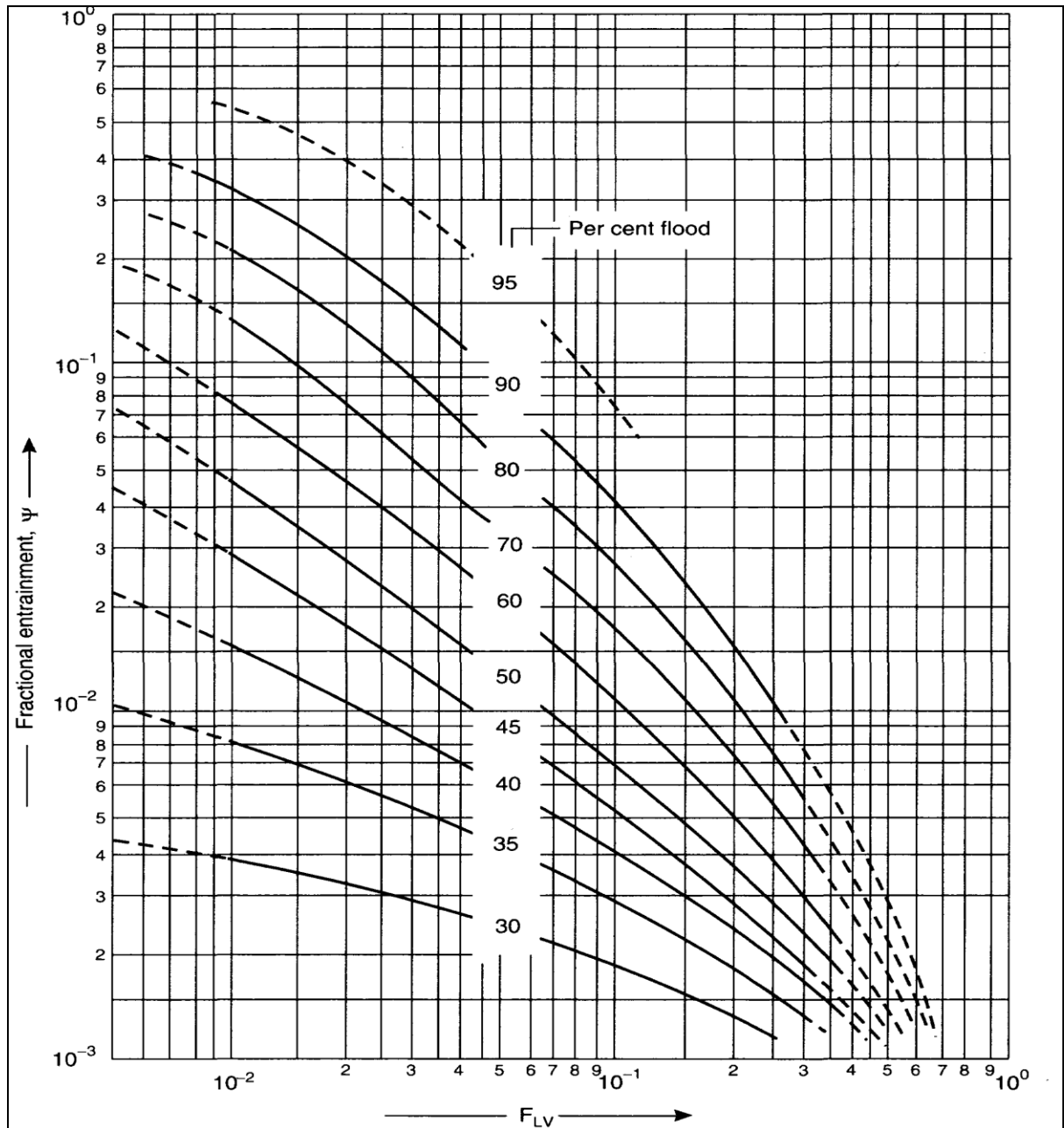


Figure (3.11) entrainment correlation for sieve plates

The effect of entrainment on plate efficiency is determined as upper limit of fraction entrainment is 0.1 below this value the effect of entrainment is small.

3.7. Weir dimensions:

- Weir height

The height of the weir determines the volume of liquid on the plate and is an important factor in determining the plate efficiency. A high weir will increase the

plate efficiency. For columns operating above atmospheric pressure the weir heights will normally be between 40 mm to 90 mm (1.5 to 3.5 in.); 40 to 50 mm is recommended. For vacuum operation lower weir heights are used to reduce the pressure drop; 6 to 12 mm is recommended.

- **Weir length**

the length of the weir fixes the area of the down-comer. The chord length will normally be between 0.6 to 0.85 of the column diameter. A good initial value to use is 0.77, equivalent to a downcomer area of 12 per cent.

The relationship between weir length and down-comer area is given in Figure 3.12

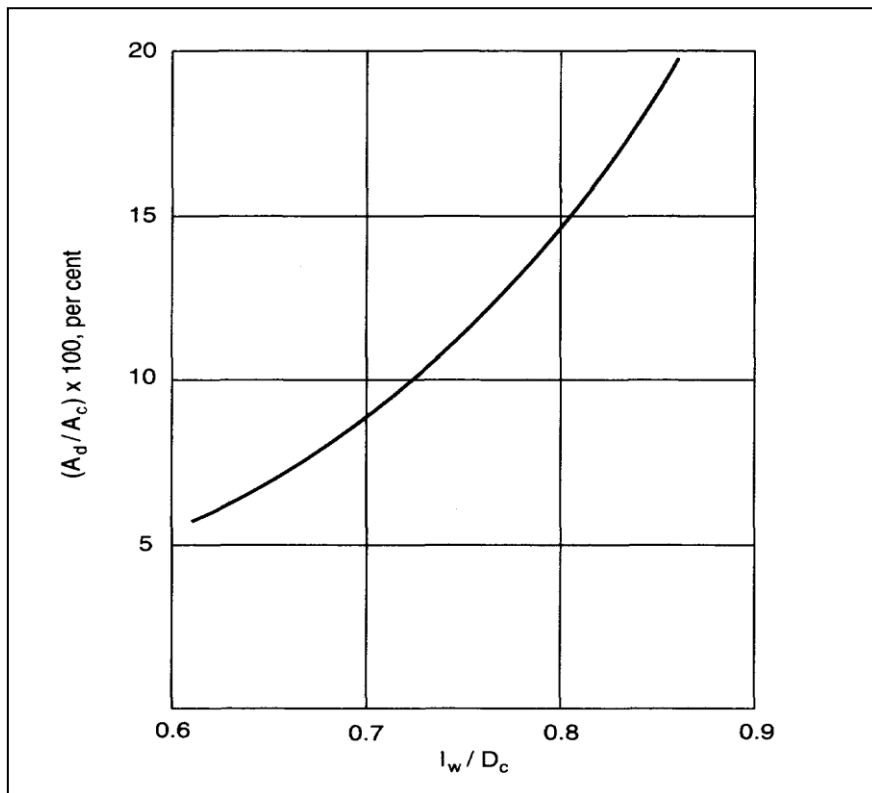


Figure (3.12) Relation between downcomer area and weir length (6)

3.8. Hole size

The hole sizes used vary from 2.5 to 12 mm; 5 mm is the preferred size. Larger holes are occasionally used for fouling systems. The holes are drilled or punched. Punching is cheaper, but the minimum size of hole that can be punched will depend on the plate thickness, For carbon steel, hole sizes approximately equal to the plate thickness can

be punched, but for stainless steel the minimum hole size that can be punched is about twice the plate thickness.

3.9. Hole pitch

The hole pitch (distance between the hole centers) l_p should not be less than 2.0 hole diameters, and the normal range will be 2.5 to 4.0 diameters. Within this range the pitch can be selected to give the number of active holes required for the total hole area specified. Square and equilateral triangular patterns are used; triangular is preferred. The total hole area as a fraction of the perforated area A_p is given by the following expression equation for an equilateral triangular pitch equation (3.27) This equation is plotted in Figure 3.13.

$$\frac{A_h}{A_p} = 0.90 \left[\frac{d_h}{l_p} \right]^2 \quad (3.27) \quad (6)$$

Where:

- ✓ $A_h \equiv$ hole area
- ✓ $A_p \equiv$ Perforated area.
- ✓ $d_h \equiv$ hole diameter.
- ✓ $l_p \equiv$ holepitch(distancebetweentheholecentres).
- **Hole area** A_h is the total area of perforations on the tray and it taken as percent from active area.
- **The Active area** A_a is the total cross sectional area less the total of down-comer area.

$$A_a = A_c - 2A_d \quad (3.28) \quad (6)$$

$$A_h = \text{percent} \times A_a \quad (3.29)$$

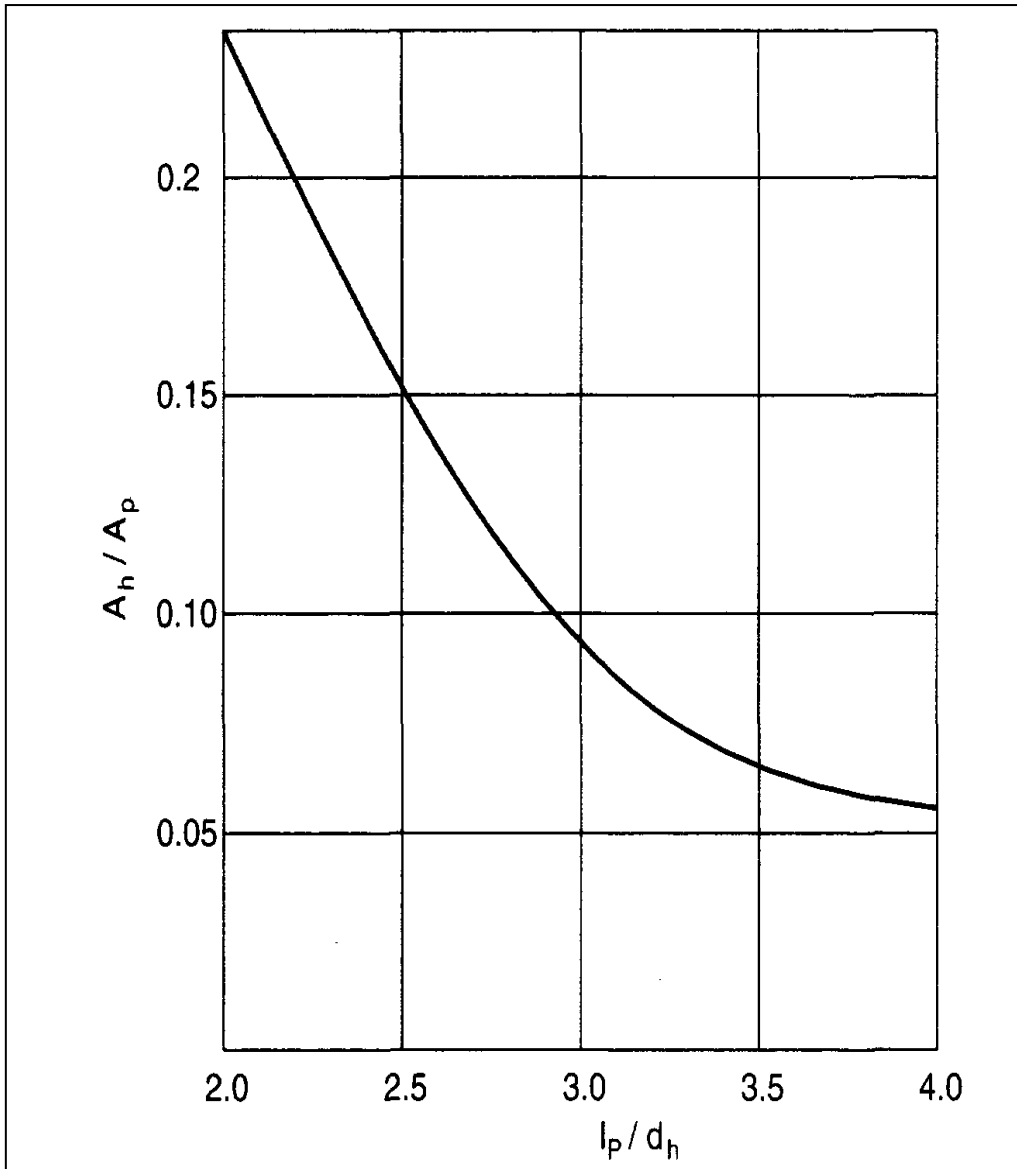


Figure (3.13) Relation between hole area and pitch (6)

3.10. Plate Pressure Drop (P.D):

The pressure drop over the plates is an important design consideration. There are two sources of pressure loss due to vapor flow through the holes (dry plate P.D (orifice loss)), and static head of liquid on the plate.

The total pressure drop taken as the sum of pressure drop due to flow of vapor through the dry plate h_d ; the head of liquid on the plate; minor sources of pressure drop residual loss h_r .

- **Dry plate drop**

The pressure drop through the dry plate can be estimated using expressions:

$$h_d = 51 \left[\frac{u_h}{c_o} \right]^2 \frac{\rho_v}{\rho_l} \quad (3.30) \quad (6)$$

Where:

$C_o \equiv$ the orifice coefficient.

C_o is a function of the plate thickness, hole diameter, and the hole to perforated area ratio. C_o can be obtained from Figure 3.14.

$u_h \equiv$ the velocity through the holes.

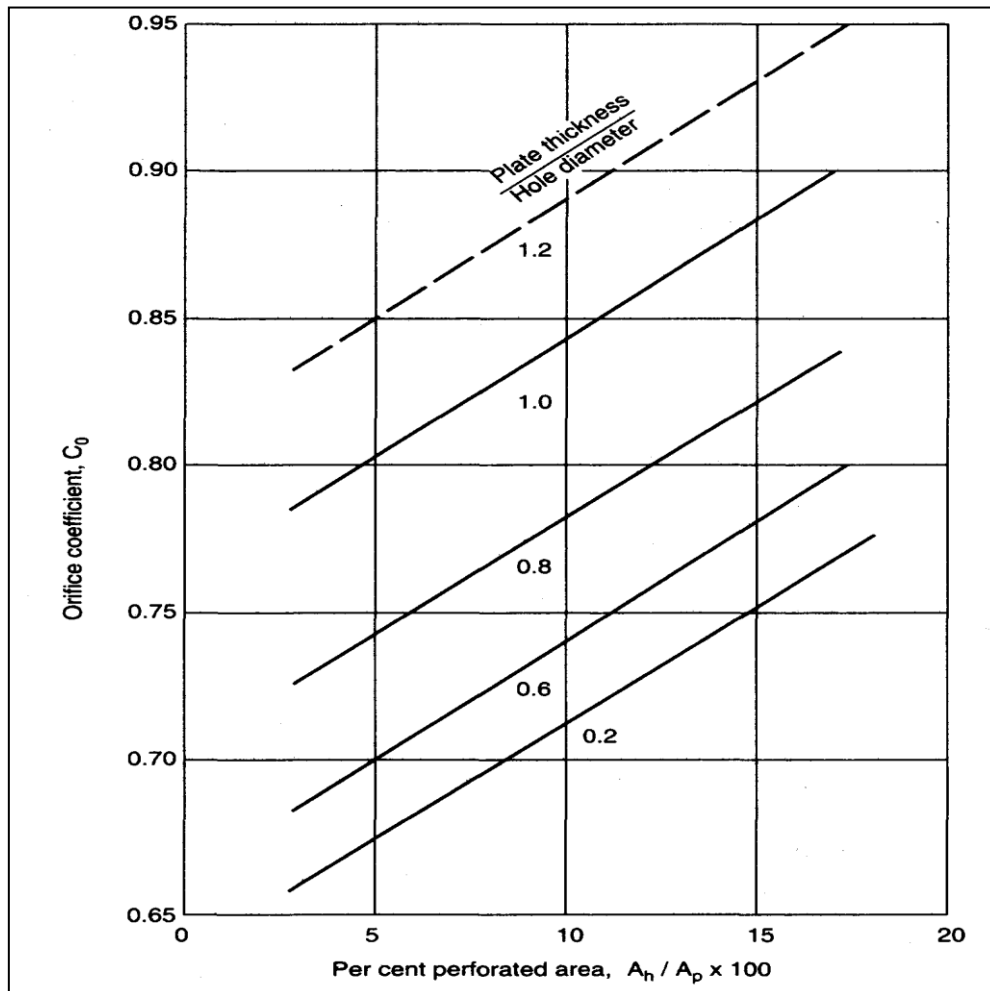


Figure 3.14 Discharge coefficients, sieve plates (6)

- **Residual head**

$$h_r = \frac{12.5 \times 10^3}{\rho_l} \quad (3.31) (6)$$

Where:

$h_r \equiv$ Residual head.

$\rho_l \equiv$ density of liquid.

Total drop

The total plate drop is given by:

$$h_t = h_d + (h_w + h_{ow}) + h_r \quad (3.32) (6)$$

$h_t \equiv$ total plate pressure drop.

$h_d \equiv$ dry plate pressure drop.

$h_{ow} \equiv$ weir crest, mmliquid.

$h_w \equiv$ weir height.

3.11. Downcomer Liquid Backup:

Caused by Pressure Drop over the plate and resistance to flow in the down-comer it self.

$$h_b = (h_w + h_{ow}) + h_t + h_{dc} \quad (3.33) (6)$$

Where:

$h_b \equiv$ downcomer back-up, measured from plate surface, mm.

$h_{dc} \equiv$ head loss in the downcomer, mm.

$h_t \equiv$ total plate pressure drop.

$h_{ow} \equiv$ weir crest, mmliquid.

$h_w \equiv$ weir height.

Head loss in the down comer can be estimated using the equation given as,

$$h_{dc} = 166 \left[\frac{l_{wd}}{\rho_l A_{ap}} \right]^2 \quad (3.34) (6)$$

Where:

$l_{wd} \equiv$ the liquid flow rate in downcomer.

$A_{ap} \equiv$ clearance area under the downcomer.

$\rho_l \equiv$ density of liquid.

$$A_{ap} = h_{ap} l_w \quad (3.35) (6)$$

Where:

$h_{ap} \equiv$ height of the bottom edge of the apron above the plate.

$l_w \equiv$ weir length.

This height is normally set at 5 to 10 mm below the outlet weir height.

$$h_{ap} = h_w - (5 - 10) \text{mm} \quad (3.36) (6)$$

Where:

$h_w \equiv$ weir height.

3.12. Downcomer residence time:

$$t_r = \frac{A_d h_{bc} \rho_l}{l_{wd}} \quad (3.37) (6)$$

Where:

$t_r \equiv$ residence time, s,

$h_{bc} \equiv$ Clear liquid back-up, m.

$\rho_l \equiv$ density of liquid.

$A_d \equiv$ downcomer area .

$l_{wd} \equiv$ the liquid flow rate in downcomer.

3.13. Numbers of Holes:

$$\text{Areaof1Hole} = \frac{\pi}{4} D_{\text{hole}}^2 \quad (3.38)$$

$$\text{NumberOfHoles} = \frac{\text{AreaofNHoles}}{\text{Areaof1Hole}} \quad (3.39)$$

3.14. Height of Distillation Column:

$$H_c = (N_{act} - 1)H_s + \Delta H + \text{plate thickness} \quad (3.40)$$

Where:

$H_s \equiv$ tray spacing, m.

$N_{act} \equiv$ actual number of plate.

$H_c \equiv$ Height of Distillation Column.

$\Delta H \equiv$ 0.5 meter each for liqued hold up and vapor disengagement .

Chapter 4

The specification and calculations of the Design

Chapter4

The specification and calculations of the Design

4.1 .design specifications

We are going to design a distillation column to separate benzene-toluene mixture with feed flow rate 3000Kmol/hr , the feed is saturated liquid, the feed has 60% mol fraction of benzene and the over head product has 0.95 mol fraction of benzene and the bottom product contain 0.05 mol fraction of benzene. The relative volatility of benzene to toluene equals 2.3.

The distillation column operates at atmospheric pressure ($p= 1\text{atm}$) and the operating reflux ratio 2 the minimum value. and the feed temperature can be calculated as follow:

$$T_{feed} = 0.6 \times 80.1 + 0.4 \times 110.6 = 92.3^{\circ}\text{C}$$

➤ Summary of design specifications:

Table (4.1) the design specifications

Feed rate F	3000kmol/hr
Feed composition	Benzene=60% ,toluene=40%
Feed temperature	92.3C
Column operating pressure	Atmospheric
column reboiler	Partial
Column condenser	Total
Distillate composition x_d	95% benzene
Bottom composition x_b	95% toluene

4.2. Assumptions made during the design:

we used the McCabe-Thiele method of column design, the following assumptions are inherent in our calculation:

- Constant vapor and liquid flow rates in given section of the tower.
- The latent heat of evaporation is approximately constant with composition and also does not vary much as we proceed from tray to tray
- The system is non-foaming and non corrosive, and thus we can use carbon steel rather than stainless steel as our material of construction.

4.3. Design procedure:

The following steps should be followed in design in the column by this method

1. Determine the vapor-liquid equilibrium curve (x-y diagram) from Antoine data.
2. Obtain the physical data of benzene and toluene required for the design.
3. Calculate the flow rate of various stream through the column
4. Calculate the minimum reflux ratio and the minimum number of trays required.
5. Using the physical data and flow rates calculate the reboiler and condenser duties
6. Calculate maximum and minimum liquid and vapor flow rates.
7. To start the iteration, select reasonable plate spacing and using the trial plate spacing calculate the column diameter.
8. Select a trial plate layout, select down-comer area, active, area and size, weir height and length.
9. From this data check that the weeping rate is satisfactory
10. Calculate the plate pressure drop.
11. Check that the down-comer area backup is acceptable.
12. If at any stage some of the values are too high or low select new trial values and repeat the iterations above.

4.4. Constructing the XY diagram:

Step1: use the Antoine equation to calculate vapor pressure of benzene and toluene with temperature range for boiling points of benzene and toluene:

Where:

$$T_{benzen}^{bp} = 80.1^{\circ}C \quad T_{toluene}^{bp} = 110.6^{\circ}C$$

Table (4.2) The constants A, B, C for this antonion equation

Compound	A	B	C
Benzene	6.90565	1211.033	220.79
Toluene	6.95334	1343.943	219.377

Example for calculation the vapor at T=80.1°C:

- For benzene : $\text{Log}P_{vap}^{\circ} = A - \frac{B}{T+C} = 6.90565 - \frac{1211.033}{(80.1+220.79)}$

$$p_B^{\circ} = 760 \text{ mmHg}$$

- For toluene: $\text{Log}P_{vap}^{\circ} = A - \frac{B}{T+C} = 6.905334 - \frac{1343.943}{(80.1+219.377)}$

$$p_T^{\circ} = 292.2 \text{ mmHg}$$

Step2: use equation (2.9) $X_A = \frac{p - p_B}{p_A - p_B}$ (for mixture A,B)

Where:

$$P = 1 \text{ atm} = 760 \text{ mmHg}$$

$$p_B \equiv \text{vapor pressure of component B (toluene } P_T^{\circ})$$

$$p_A \equiv \text{vapor pressure of component A (benzen } P_B^{\circ})$$

$$\text{At } T = 80.1^{\circ}C \quad p_B = 292.2 \text{ mmHg} \quad p_A = 760 \text{ mmHg}$$

$$X_A = \frac{760 - 292.2}{760 - 292.2} = 1$$

Step3: using relative volatility $\alpha = 2.3$ and applied the equation (2.4)

$$\text{At } T = 80.1^{\circ}C \quad y_B = (2.3 * 1) / (1 + (2.3 - 1) * 1) = 1$$

Applying the above steps for the other temperature between (80.1- 110.6) we obtained the following results, listed in table (4.3) below:

table (4.3) T-X-Y data for benzene

Temperature(°C)	P_B°	P_T°	X_B	y_B
80.1	760.0	292.2	1.000	1.00
82	805.5	311.9	0.908	0.957
84	855.7	333.7	0.817	0.911
86	908.3	356.8	0.731	0.862
88	963.3	381.1	0.651	0.81
90	1021.0	406.7	0.575	0.756
92	1081.3	433.7	0.504	0.700
94	1144.3	462.1	0.437	0.641
96	1210.1	492.0	0.373	0.578
98	1278.8	523.4	0.313	0.511
100	1350.5	556.3	0.256	0.442
102	1425.2	590.9	0.203	0.369
104	1503.1	627.2	0.152	0.291
106	1584.2	665.2	0.103	0.209
108	1668.6	704.9	0.057	0.122
110	1756.4	746.6	0.013	0.029
110.6255	1784.5	760	0.00	0.00

Step4: from data in table (4.1) plot X_B versus y_B figure (4.3)

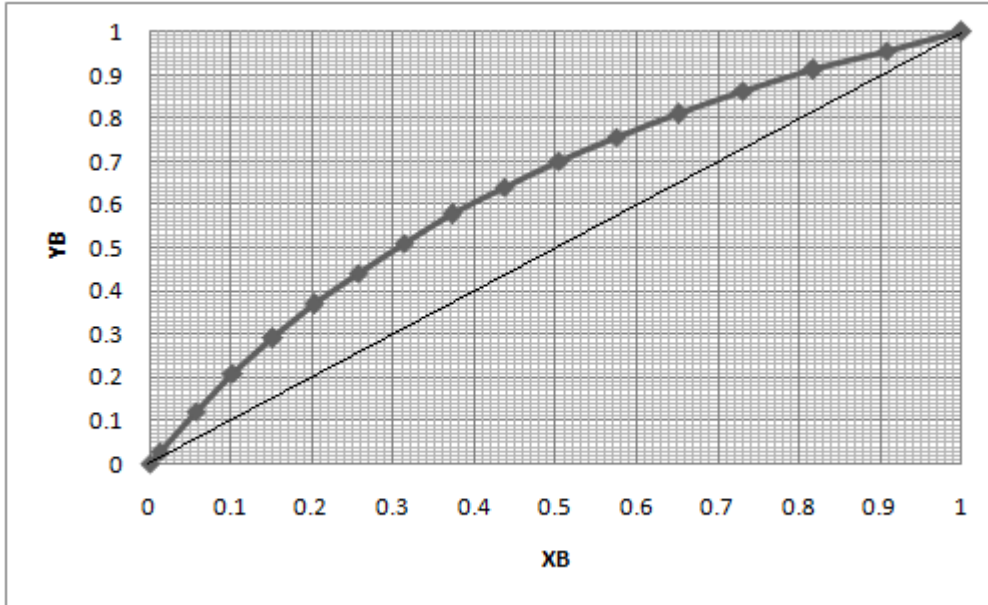


Figure (4.1) equilibrium curve (X_B versus y_B)

4.5. Physical properties:

Physical Data obtained is from various sources as note:

➤ **Molecular weight :**

For benzene 76.114 kg/kmol

For toluene 92.141 kg/kmol

➤ **Viscosity**

$\log(\mu) = \text{vis}A \left[\frac{1}{T} - \frac{1}{\text{vis}B} \right]$ with T in K and μ in SI unit

For Benzene visA = 545.64 visB=265.34 $\mu_{ave} = 0.27 \text{ cp}$

For toluene visA= 467.64 visB= 255.24 $\mu_{ave} = 0.28 \text{ cp}$

Data from [6,p(947,959,960)].

➤ **Density in liquid**

Density calculated from two constants of Racket equation $\rho = A \cdot B^{-\left[1 - \frac{T}{T_c}\right]^{\frac{2}{7}}}$ [(2) p(42)].

For benzene A=0.3641 B=0.269 $T_c = 562 K$

For toluene A=0.2906 B=0.265 $T_c = 593.1 k$

From this data we can calculate the density does not vary much over the entire column

$$\rho_l = 800 \text{ kg/m}^3$$

Data from [11].

➤ **Density in the vapor**

In the absence of data we can use the ideal gas law

The top of column $\rho_{vtop} = 2.716 \text{ kg/m}^3$

The bottom of the column $\rho_{vbottm} = 2.905 \text{ kg/m}^3$

➤ **Enthalpies:**

For benzene: $cp_{top,liquid} = 33 \frac{\text{cal}}{\text{mol}^\circ\text{C}} \times \frac{4.18\text{J}}{\text{cal}} = 138 \frac{\text{J}}{\text{mol}^\circ\text{C}}$

$$cp_{bottom,gas} = 23 \frac{\text{cal}}{\text{mol}^\circ\text{C}} \times \frac{4.18\text{J}}{\text{cal}} = 96.14 \frac{\text{J}}{\text{mol}^\circ\text{C}}$$

For toluene: $cp_{top,gas} = 33 \frac{\text{cal}}{\text{mol}^\circ\text{C}} \times \frac{4.18\text{J}}{\text{cal}} = 138 \frac{\text{J}}{\text{mol}^\circ\text{C}}$

$$cp_{bottom,liquid} = 40 \frac{\text{cal}}{\text{mol}^\circ\text{C}} \times \frac{4.18\text{J}}{\text{cal}} = 167.2 \frac{\text{J}}{\text{mol}^\circ\text{C}}$$

Data from [11].

➤ **Heat of vaporization:**

For Benzene $\Delta H_{vap,81^\circ\text{C}} = 7360 \text{ cal/mol} \times 4.18 \frac{\text{J}}{\text{cal}} = 30.67 \text{ KJ/mol}$

$$\Delta H_{vap,110^\circ\text{C}} = 29.02 \text{ kj/kmol}$$

For toluene $\Delta H_{vap,81^{\circ}\text{C}} = 7960 \text{ cal/mol} \times 4.18 \frac{\text{J}}{\text{cal}} = 33.272 \text{ KJ/mol}$

$$\Delta H_{vap,110^{\circ}\text{C}} = 33.22 \text{ kj/kmol}$$

Data from [11].

➤ **Surface tensions:**

For Benzene and toluene the surface tensions are approximately = 20 mj/m^2

Data from[12].

➤ **Heat capacity of water**

$$cp_{water} = 4.190 \frac{\text{kj}}{\text{kg.k}}$$

Data from [11].

4.6. Design calculations:

4.6.1. Flow sheet:

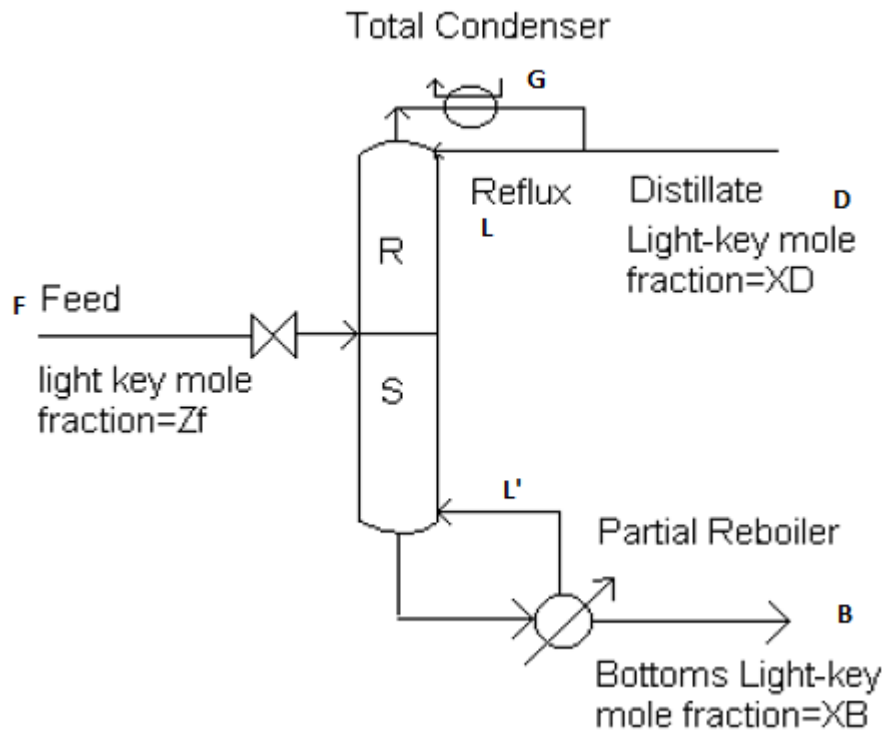


Figure (4.2) a typical distillation column

➤ **Overall material balance:**

$$F=D+W$$

$$3000=D+W \quad (4.1)$$

$$D=3000 -W \quad (4.2)$$

➤ **Component balance for benzene:**

$$0.6*3000=0.95D+0.05W \quad (4.3)$$

Substitute (4.2) in (4.3):

$$0.6*3000=0.95(3000 -W)+0.05W \quad (4.4)$$

$$W=1166.67 \text{ Kmol/hr} , D=1833.33 \text{ Kmol/hr.}$$

These flow rates are fixed whereas the flow rates inside the column are dependent on the reflux ratio.

4.7. Minimum reflux ratio R_{min} :-

The q-line in the McCabe-Thiele diagram is used to calculate our minimum reflux ratio. The feed enters at its bubble point since the q-line is vertical. The point (intersection between q-line and equilibrium curve), and the point (intersection X_d with 45 line), the slope of line between these two points equal $R_{min}/(R_{min} + 1)$, and the intersection of this line on the Y-axis is $X_d/(R_{min} + 1)$.

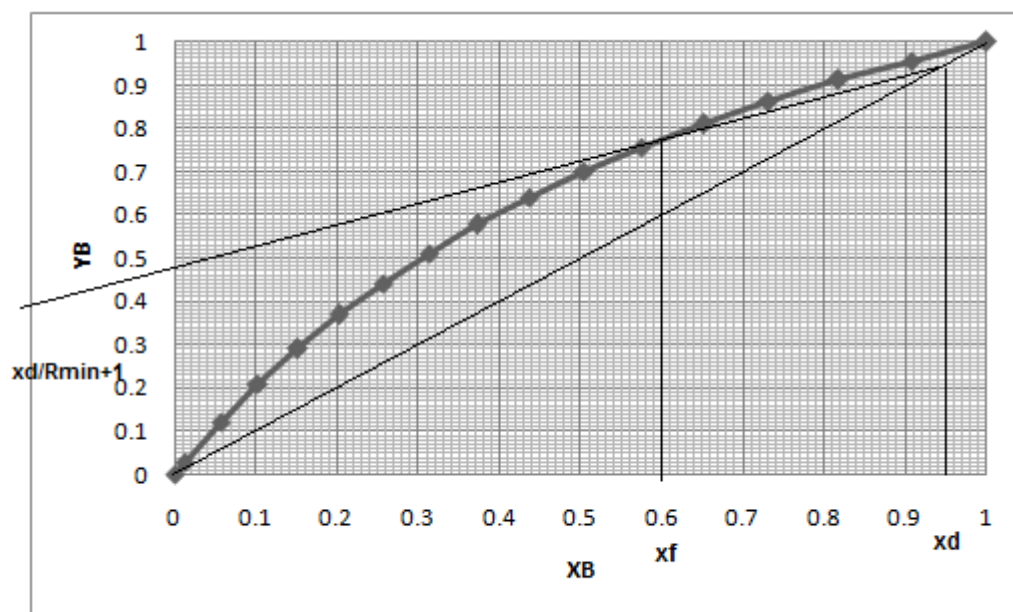


Figure (4.3) Minimum reflux ratio

From figure 4.2 the intersection on Y axis $X_d/(R_{min} + 1) = 0.48$

- $\frac{0.95}{(R_{min}+1)} = 0.48$
- $R_{min} = 0.979$

4.8. Minimum number of stage:-

The slope of rectifying section equal unity ; the operating line must coincide with the diagonal throughout the column.

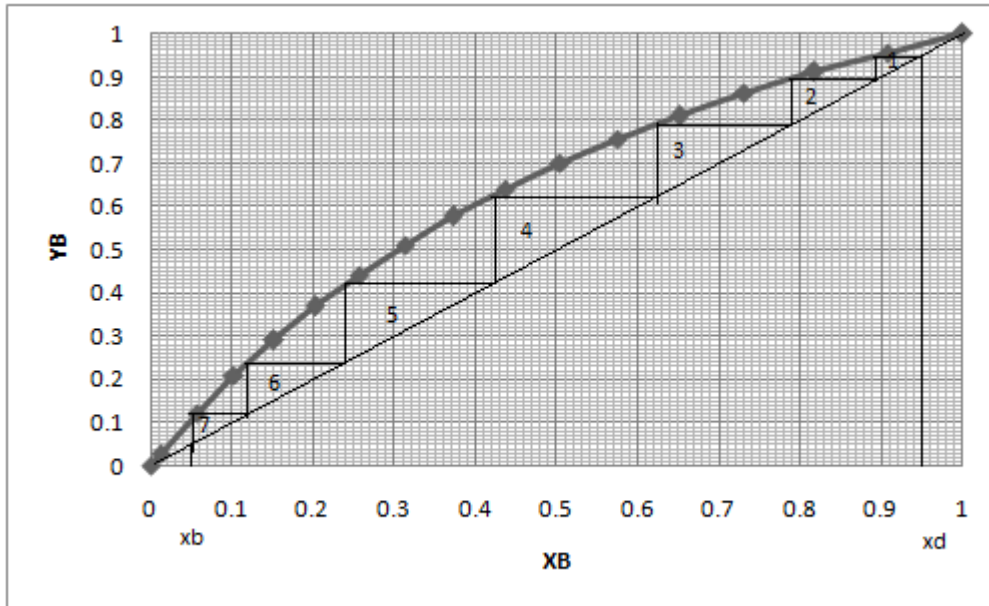


Figure (4.4) Minimum number of stage

From the McCabe-Thiele diagram $N_{min} = 7$ stages

4.9. Actual reflux ratio:-

$$R = 2R_{min} = 2 \times 0.979 = 1.958.$$

From R we determine The flow rates of vapor and liquid through the column:

- The Liquid flow rate the top section $L = RD$
 $= 1.958 \times 1833.33 = 3589.66 \text{ kmol/hr}$
- Liquid flow rate from the bottom section $\dot{L} = L + F = RD + F = 1.958 \times 1833.33 + 3000 = 6589.66 \text{ kmol/hr}$.
- Vapor flow rate in the column $G = L + D = RD + D = (R+1) D = (1.958+1) \times 1833.33 = 5422.99 \text{ kmol/hr}$.
- Vapor mass flow rate

$$V_w = (G) * MW_{ave}$$

$$MW_{ave} = 0.6 \times 78.114 + 0.4 \times 92.141 = 89.12 \text{ kg/kmol}.$$

$$V_w = 5422.99 \times 89.12 = 483296.86 \text{ kg/hr} = 134.24 \text{ kg/sec}.$$

- Liquid mass flow rate

$$L_w = (\dot{L}) \times MW_{ave} = 6589.66 \times 89.12 = 587270.49 \text{ kg/hr} = 163 \text{ kg/sec.}$$

- Vapor volumetric flow rate Q_v :

$$Q_v = V_w / \rho_v = 134.24 / 2.716 = 49.42 \text{ m}^3/\text{s}$$

4.10. Theoretical number of trays:

By draw the steps between operating lines and the equilibrium line and count them:

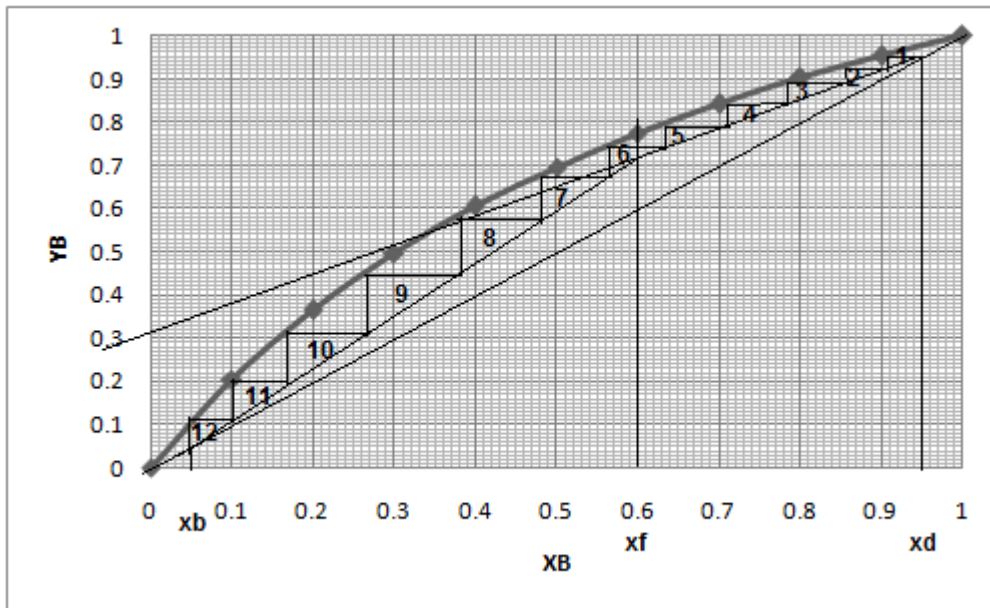


Figure (4.5) theoretical number of stage N_{theo}

From figure e(4.4) $N_{theo} = 12 \text{ stages}$.

- **The Actual number of trays** = $\frac{N_{theo}}{\epsilon_o}$

Where:

$$\epsilon_o \equiv \text{efficiency of tray, assume } 75\%$$

Actual number of tray= $(12/0.75) = 16$,we used in our calculation partial reboiler then the actual number of stage = $16-1=15$ stages .

4.11. Determining the location of feed:-

The location of the feed is determined Using eq (3.17):

$$\log \left[\frac{N_D}{N_B} \right] = 0.2061 \left[\left(\frac{1166.67}{1833.33} \right) \left(\frac{0.6}{0.4} \right) \left(\frac{0.05}{0.05} \right)^2 \right]$$

$$\left[\frac{N_D}{N_B} \right] = 1.573$$

- ✓ $N_D + N_B = 15$
- ✓ $N_D = 15 - N_B$
- ✓ $N_D = 1.573 N_B$
- ✓ $N_B = 6$

Feed location from bottom (6-1) = 5th

3.12. Condenser and reboiler duties:

The condenser at the top of the column has to only remove the heat of condensation of the gaseous stream. we can determined the condenser duty from energy balance eq(3.6):

$(L_{n+1} + D)H_{n+1} = L_{n+1} h_{n+1} + DH_d + q_c$ and substitute $D = \left(\frac{z_f - X_B}{X_d - X_B} \right) F$ in it and divide by D . after solve for q_c we obtained the following equation:

$$q_c = \left(1 + \frac{L_c}{D} \right) \left(\frac{z_f - X_B}{X_d - X_B} \right) F (H_d - h_{n+1})$$

$h_{n+1} \equiv$ the average enthalpy of gaseous stream from the stage at tope column .

$H_d \equiv$ the enthalpy of distillate product and equal to zero.

we take the temperature at top stag is 80°C .we have two component benzene with mole fraction 0.95 and toluene with 0.05 mole fraction . the enthalpy of benzene at 80°C is the latent heat of vaporization for benzene=7360 cal/mol. The enthalpy of toluene must be calculated as follow (the boiling point of toluene is 110.6°C):

$$H_{Tolat80^\circ C} = 7960 + (40 - 30)(110.6 - 80) = 8174 cal/mol$$

The average enthalpy $H_{averat80^\circ C} = 0.95 \times 7360 + 0.05 \times 8174 = 7400.7 cal/mol$

Substitute above values in eq(3.6):

$$q_c = (1 + 1.958) \left(\frac{0.6-0.05}{0.95-0.05} \right) \times 3000(0 - 7400.7) = 40.1339961 \times 10^6 \text{ cal/hr} = 46.66 \text{ KJ/s.}$$

The Reboiler duty: To determine the reboiler duty from energy balance around bottom section we obtained the following equation:

$$\text{Reboiler duty} = \dot{L} \times \Delta H_{vap,ave,110^\circ\text{C}} + (\dot{L})(C_{p,ave,bottom,110^\circ\text{C}})(\Delta T)$$

Where:

$\dot{L} \equiv$ liquid stream from bottom in Kmol/hr.

$C_{p,ave,bottom,110^\circ\text{C}} \equiv$ average specific heat capacity

$\Delta H_{vap,ave,110^\circ\text{C}} \equiv$ molar weighted heat of vaporization at 81°C.

$\Delta T \equiv$ different temperature in °C .

The enthalpy of benzene at 110°C = 7360 + 23(110.6 - 80) = 8063.8 cal/mol.

$\Delta H_{vap,ave,110^\circ\text{C}} = 0.95 \times 8063.8 + 0.05 \times 7960 = 8058.61 \text{ cal/mol.}$

$$C_{p,ave,bottom,110^\circ\text{C}} = 0.05 \times 23 + 40 \times 0.95 = 39.15 \frac{\text{cal}}{\text{mol}^\circ\text{C}}$$

$$\dot{L} = RD + F = 2.042 \times 1833.33 + 3000 = 6734659.86 \frac{\text{mol}}{\text{hr}}$$

$$\text{Reboiler duty} = 6734659.86 \frac{\text{mol}}{\text{hr}} \times 8058.61 \frac{\text{cal}}{\text{mol}} + 39.15 \frac{\text{cal}}{\text{mol}^\circ\text{C}} \times (110.6 - 80)^\circ\text{C} =$$

$$6.242 \times 10^{10} \text{ cal/hr} = 72480 \text{ kJ/s.}$$

4.13. Calculation the column diameter:

The vapor velocity of the column must be below the flooding velocity to avoid flooding occurred.

➤ Flow parameter from eq.(3.19)

$$F_{lv} = \frac{163}{134.24} * \left(\frac{2.716}{800} \right)^{0.5} = 0.071$$

➤ Flooding velocity u_f from eq.(3.18)

Firstly we need to determine the capacity parameter K_1 by Assuming the tray spacing =0.5m. from fig (3.9) $K_1 = 0.09$

$$u_f = 0.09 * \left(\frac{800-2.716}{2.716}\right)^{0.5} = 1.54 \text{ m/s.}$$

In Our calculation we will design for operation at 85% flooding velocity

$$\text{Actual vapor velocity} = 0.85 \times u_f = 1.309 \text{ m/s}$$

➤ Net column area A_n from eq (3.20)

$$A_n = \frac{51.03}{1.309} = 38.98 \text{ m}^2$$

Assuming for all our calculation that one downcomer =12% so $A_d = 0.12A_c$

to calculate column diameter we must calculate cross-section area, from eq(3.22)

$$A_c = 38.98 + 0.12A_c$$

$$A_c = 44.29 \text{ m}^2$$

The cross section area $A_c = \left(D^2 * \frac{\pi}{4}\right)$ this implies to:

$$D=7.5 \text{ m}$$

4.14. Plate layout:

$$\checkmark A_c = 38.98 \text{ m}^2$$

$$\checkmark A_d = 0.12A_c = 4.677 \text{ m}^2$$

$$\checkmark \text{ From eq (3.28) } A_a = 38.98 - (2 \times 4.677) = 29.62 \text{ m}^2$$

$$\checkmark \text{ The hole area } A_h = 0.1A_a = 2.962 \text{ m}^2$$

$$\checkmark \text{ Weir length from fig (3.12) } \frac{A_d}{A_c} \times 100 = 11.99 \text{ and } l_w/D_c = 0.76 \text{ so}$$

$$l_w = 5.7 \text{ m.}$$

Recommend initial guesses are: $d_h = 5 \text{ mm}$, Weir height $h_w = 50 \text{ mm}$, plate thickness $\tau = 5 \text{ mm}$

4.15. Check weeping:

The minimum design vapor velocity given by eq (3.25)

Where $h_{ow} = 750 \left(\frac{163}{800 \times 5.7} \right)^{2/3} = 17.1 \text{ mm liquid}$

Maximum liquid rate = 163 kg/sec

Minimum liquid rate = 0.7 maximum liquid rate = 114.1 kg/sec

At minimum liquid rate: $h_{ow \text{ min}} = 750 \left(\frac{114.1}{800 \times 5.7} \right)^{2/3} = 13.5 \text{ mm liquid}$

$$h_{ow} + h_w = 63.5 \text{ mm}$$

From fig (3.10) $K_2 = 30.3$

So using eq(3.25) the $\check{u}_{min} = \frac{[30.3 - 0.9(25.4 - 5)]}{2.716^{0.5}} = 7.24 \text{ m/s}$.

- Minimum operating point will be above the weep point.
- The maximum vapor velocity through the holes $\check{u}_{max} = \frac{51.03}{2.87} = 17.78 \text{ m/s}$.
- This actual vapor velocity = $17.78 \times 0.7 = 12.4 \text{ m/s}$ is above the weep point.

4.16. Plate Pressure Drop (P.D):

➤ **Dry pressure drop h_d :**

For flow through the orifices we have eq (3.30) where C_o is determined from fig (3.14)

by take ($A_p = A_a$), $\frac{A_h}{A_p} = 0.1$, $\tau/d_h = 1$ thus $C_o = 0.84$, $h_d = 51 \left[\frac{17.78^2}{0.84} \right] \frac{2.716}{800} = 77.57 \text{ mm}$

liquid

➤ **Residual head:**

From eq (3.31) the residual head $h_r = 15.62 \text{ mm liquid}$.

➤ **Total pressure drop:**

From eq (3.32) $h_t = 50 + 17.1 + 77.5 + 15.62 = 160.22 \text{ mm liquid}$

4.17. Downcomer liquid back up:

The height of apron by applied in Eq (3.36) $h_{ap} = 40 \text{ mm}$, and from (3.35)

$$A_{ap} = 0.04 \times 5.7 = 0.228 \text{ m}^2$$

From eq (3.34) $h_{dc} = 166 \left(\frac{163}{800 \times 0.228} \right)^2 = 132.5 \text{ mm}$

then, the back up in downcomer by applied in(3.33) = $160.22 + 132.5 + 50 + 17.1 = 0.241m = 0.241m < 0.5(\tau + h_w)$ so the tray space is acceptable.

4.18. Downcomer residence time:(t_r)

It was calculated from Eq (3.37) = $\frac{4.677 \times 0.3598 \times 800}{163} = 8.25 \text{ s}$

4.19. Check entrainment:

$F_{lv} = 0.071$, and percent flooding =85%

From fig (3.11) the fractional entrainment $\psi = 0.05$.

Well below the upper limit of (ψ) which is 0.1. Below this the effect of entrainment on efficiency is small.

4.20. Number of holes:

Area of 1 hole = $D_h^2 * \pi/4 = (3.14/4) \times 5^2 = 1.96 \times 10^{-5} m^2$

The number of holes = $(2.962 m^2 / 1.96 \times 10^{-5} m^2) = 150930$ holes.

4.21. Height of column

The Height of column from eq(3.40) $H_c = (N_{act}-1) H_s + \Delta H + \text{plate thickness} = (16-1)0.5 + 1 + 0.005 = 9m$.

Chapter 5

Materials of construction

Chapter5

Materials of construction

In this chapter suitable material is selected for the designed distillation column, according to the following standards:

- Ability to resist corrosion.
- Have sufficient strength and be easily worked.
- Lowest cost.
- Product contamination and process safety.

5.1. Materials properties

5.1.1. Strength:

It is the maximum stress that the material will withstand, measured by a standard tensile test. The design stress for a material, the value used in any design calculations, is based on the tensile strength

5.1.2. Hardness:

It is an indication of a material's ability to resist wear, This will be an important property if the equipment is being designed to handle abrasive solids, or liquids containing suspended solids which are likely to cause erosion.

5.2. Effect of temperature on the mechanical properties:

Equipment is being designed to operate at high temperatures, materials that retain their strength must be selected.

5.3. Corrosion resistance:

The corrosion divided into five types there are:

1. uniform corrosion:

it is occurs over the majority of the surface of a metal It is predictable, make the material thick enough to function for the lifetime of the component.

In corrosion testing, the corrosion rate is measured by the reduction in weight of a specimen of known area over a fixed period of time.

2. Galvanic corrosion:

It occurs when two different metals are placed in contact with each other and is caused by the greater willingness of one to give up electrons than the other.

3. Intergranular corrosion.

It is the preferential corrosion of material at the grain (crystal) boundaries.

It is caused by the physical and chemical differences between the centre's and edges of the grain.

4. Stress corrosion:

Corrosion rate and the form of attack can be changed if the material is under stress within normal design stress values, the rate of attack will not change significantly. Avoiding stress corrosion can be made by selecting materials that are not susceptible in the specific corrosion environment; or, less certainly, by stress relieving by annealing after fabrication and welding.

5. Erosion-corrosion:

is used to describe the increased rate of attack caused by a combination of erosion and corrosion, erosion will tend to remove the products of corrosion and any protective film, and the rate of attack will be markedly increased. If a fluid stream contains suspended particles, or where there is high velocity.

5.4. Selection for corrosion resistance:

Many factors must be considered to select the correct material of construction such as:

- Pressure.
- pH.
- Stream velocity.
- Temperature.

5.5. Material costs:

The cost of materials depending on its properties, when the materials have best properties the cost of it is high. The table (5.1) below shows the price of some metals in \$/tonne.

Table 5.1. The price of some metals

Metal	\$/tone
Carbon steel	300
Low alloy steels	400-700
Copper	800
Aluminum	900
Nickel	3000
Titanium	20,000

5.6. Contamination:

When selecting suitable materials, we prevented of the contamination of a process stream, or a product. With processes that use catalysts, care must be taken to select materials that will not cause contamination and poisoning of the catalyst.

5.7. Carbon steel:

It is most engineering materials use in the construction of equipments, It is cheap, available in a wide range of standard forms and sizes, and can be easily worked and welded. But it not resistant to corrosion, except in certain specific environments.

It is suitable for use with most organic solvents, except chlorinated solvents

Chapter 6

Discussion and results

Chapter6

Discussion and results

This chapter represents a summary of the results and calculations made in this project , It has been found that for the distillation of a binary mixture that contains 60% benzene and 40% toluene a distillation column of 9 meters long with a diameter of 7.5 meters is needed.A set of 15 sieve trays can give an efficiency of 75% other specifications are discussed below and summarized in tables 6.1 and 6.2.

6.1. Discussion of results:

- the velocity of vapor through the holes is 1.309m/s which is below flooding velocity1.54m/s.That means the flooding can not occur.
- the fractional entrainment $\psi = 0.05$.Well below the upper limit of (ψ) which is 0.1. Below this the effect of entrainment on efficiency is small.and the over all column efficiency is estimated to be 75%.
- The total pressure drop is 160.22 mm liquid which is Not excessive.
- The downcomer liquid backup (0.241m) that is the less than the half of plate spacing 0.5 m and weir high 0.05m added that indicate the flooding can not occur and tray spacing is acceptable.
- The minimum design vapor velocity 7.24 m/s and the actual vapor velocity 12.4m/s is above the weep point.
- There is residence time of 8.25s in the downcomer which is efficient time to allow the vapor and liquid disengage.

The following table (6.1) and (6.2) represents the summary of calculations that achieved in chapter4:

6.2. Column parameters:

Table 6.1 the column parameters

Parameter	Size
Column diameter	7.5 m
Column height	9 m
Condenser duty	46.66 <i>kJ/s</i>
Reboiler duty	72480 <i>kJ/s</i>
Number of actual trays	15
Number of theoretical tray	12
Feed entry tray	5
Pressure drop over the column	160.22 mm liquid
Percentage flooding	85%

6.3. Tray parameters:

Table 6.2 the tray parameters

Parameter	Size
Material of construction	Carbon steel
Column area	44.29m ²
Downcomer area	4.677m ²
Active area	29.62m ²
Total hole area	2.962m ²
Hole diameter	5 mm
Number of hole	150930
Hole pitch	triangular
Plate spacing	0.5 m
Plate thickness	5 mm

Recommendation:

- It is recommended to perform an effective control study to construct effective system controlling the critical parameter.
- Using software to make simulation for distillation was designed.
- It is recommended to make control for important parameter such as flow rate, pressure and temperature.

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