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Design of a distillation column for palm oil Cracking

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Abstract
The design of a Distillation column is necessary for cracking of palm oil to produce oleochemicals. The proposed distillation column is to crack 500Kg per day of palm oil aimed at producing Fatty acid ethyl ester (oleochemicals). This figure was chosen as the basis of the design. Also the energy and material balance of the distillation column was calculated. Finally the equipment design for the distillation column was carried out. The number of plates required is 17 plates with column diameter and height of 0.1m and 11.55m respectively. The study shows the importance of a distillation column in cracking of palm oil which buttresses the point that complete conversion of triglyceride to oleochemicals can be achieved with the aid of a fractionating column.

Keywords: Distillation column, material balance, energy balance, palm oil cracking, oleochemicals, equipment design.

1. Introduction
Oleochemicals are chemicals derived from oils and fats. They are analogous to petrochemicals which are chemicals derived from petroleum [1, 2, 3, 4]. Oleochemicals or derivatives based on C_{12}-C_{14} and C_{16}-C_{18} chain lengths have a variety of uses. The hydrolysis or alcoholysis of oils and fats formed the basis of the Oleochemicals industry [5, 6]. The five basic Oleochemicals are Fatty acids, Fatty Methyl esters and Ethyl esters, fatty alcohol, fatty nitrogen compounds and glycerol. The process of derivation or obtaining these chemical intermediates from palm oil requires high temperature pyrolysis, alcoholysis, gasification or destructive distillation in the presence of a catalyst [7, 8, 9].

Nomenclatures
V_O	Volumetric flow rate, m^3/hr
X	Conversion
M_{AO}	Initial mass flow rate, kg/hr
P	Density, kg/m^3
K	Rate constant, g mol/min
T	Temperature, K
P	Power, kW
V	Reactor volume, m^3
H	Height, m
L	Baffle spacing, m
d_O	Baffle orifice diameter, m

2. Background to the study
The reactor is the starting point of the cracking process were the raw material which is the palm oil is introduced at an ambient temperature and is heated to a certain temperature made possible by the aid of a heat exchanger device embedded in the reactor thereafter the content of the reactor is transferred to the distillation column where the different components in terms of the oleochemicals are separated by the rectification and stripping sections of the column based on their different boiling points. Material Balance is carried out which summarizes the sum of masses in the reactor and distillation column in conformity with the law of conservation of matter. The Energy balance shows the heat effects of chemical reaction and physical transformation which occurs in the equipment. The equipment design shows the various design specifications and conditions for the design of the distillation column in conformity with the standard design procedures and for the distillation column to function effectively. The basic procedure that will lead to the yield of oleochemicals consists of pre-heating the palm oil (triglyceride) with the aid of a heat exchanger, and then passed to the pump, and from the pump

It is charged into the reactor which operates at a chosen temperature of 374 ^{O}C and atmospheric pressure, their after is transferred to the distillation column for separation of the different oleochemicals.

3. Design Method for Distillation Column

The geometric configuration is a proposed approach to design a Distillation (fractionating) column by maintaining dynamic stability using various constants, McCabe-Thiele method of continuous distillation was used to achieve this purpose.

3.1 Equation of the reaction and mechanism

Equation for the reaction:

Triglycerides ⟶ Heat Fatty Acids + Fatty acid Methyl Ester + Glycerol [10]

Triglycerides ⟶ Heat Ethyl propanate + Stearic acid + Linoleic acid + Ketene (octadecatrien-1-ol)

Mechanism

CH_2OCOR^{I} CH_3

I I

$CHOCOCH_2R^{II}$ ⟶ $CH_2 + R^{I}COOH + R^{II}COOH + R^{III}CH=CHO$

I I

CH_2OCOR^{III} $COOC_2H_5$

Triglyceride Ethyl Propanoate Stearic acid Linoleic acid Ketene (octadecatrien-1-ol)

Figure 1: Reaction equation for cracking of vegetable oil [11].

$R^{I} = (CH_2)_{16}CH_3$,

$R^{II} = (CH_2)_6CH=CH(CH_2)_7CH_3$

$R^{III} = (CH_2)_7CH=CH-CH_2-CH=(CH_2)_4CH_3$

Assumptions

1. 95% of FAEE produced at reactor is recovered as distillate.
2. For practical purpose, steady state condition is assumed
3. Batch operation
4. A day is assumed to be 8 hours
5. 85% conversion

Basis: 1 day operation

Plant capacity = 500kg/day

3.2 Calculation parameters used

Physical/Chemical Properties of Equation of Reaction:

Density

Triglyceride- 0.928g/cm^3

Ethyl propanoate- 0.8844g/cm^3

Stearic Acid- 0.847g/cm^3 at 70^{o}C, 1.017g/cm^3 at 17 ^{o}C

Ketene (octadecatrien-1-ol):1.08g/cm^3

Linoleic Acid- 0.90g/cm^3

Vapour Pressure

Triglyceride- Between 25-760mmHg

Ethyl propanoate- 35.90mmHg, 5.32kpa/27 ^{o}C, 44mmHg at 27.2 ^{o}C

Stearic Acid- 10mmHg at 21 ^{o}C, 1mmHg at 344.7^{o}F, 5mmHg at 408.2^{o}F

Ketene (Octadecatrien-1-ol): Greater than 1atm, 1.4*10^{3}kPa at 25^{o}C

Linoleic Acid- 3.5mmHg at 25 ^{o}C, 1.0mmHg at 180 ^{o}C, 1.0mmHg at 176.5^{o}F

Viscosity

Triglyceride- 15.9cSt at 40 ^{o}C

Also Palm Oil (Triglyceride)

Kinematic Viscosity is 39.6mm^2/s at 38 ^{o}C,

Density is 0.9180g/cm^3

Therefore Viscosity = Kinematic Visco. * Density = 39.6 * 0.9180 = 36.3528

At room temp. Viscosity is 8.46Pa.s

Ethyl propanoate- 0.426cP at 25^{o}C

Stearic Acid- 176.50cP (160 ^{o}C), 9.04cP (75 ^{o}C), 7.79mPa.s (80 ^{o}C), 6.29m Pa.s (90 ^{o}C)

Linoleic Acid- 29.0cP (25 ^{o}C), 17.7cP (38 ^{o}C), 13.1cP (50 ^{o}C), 9.41cP (60 ^{o}C), 3.41cP (110 ^{o}C)

Rate Constant Values

K values of the cracking of palm oil in autoclave Batch reactor

Temperature	Rate constant
623	0.0053gmol/min
648	0.0021gmol/min
673	0.0060gmol/min

4. Results and Discussion

4.1 Material balance:

Plant capacity = 500kg/day

$$\text{Capacity} = \frac{500}{8} = 62.5\text{kg/hr}$$

Considering 20% Safety factor

$$\text{Plant through put} = \frac{20}{100} \times 62.5 = 75\text{kg/hr}$$

Note: This safety factor takes care of material loss due to over pressure and other losses

From assumption, 95% of FAEE expected to be produced at reactor is recovered as distillate

$$\text{FAEE is } S_2 = \frac{75}{95} \times \frac{100}{1} = 78.95\text{kg/hr}$$

$$\text{Kmol of FAEE produced from reactor} = \text{mass} = \frac{78.95}{\text{molar mass } 102.13} = 0.733$$

Kmol of FAEE produced=0.773kmol/hr

$$\text{Conversion} = \frac{\text{kmoles of palm oil consumed}}{\text{kmoles of palm oil fed}} \times \frac{100}{1}$$

Similarly,

$$\text{Reactor Yield} = \frac{\text{kmoles of FAEE produced}}{\text{kmoles of palm oil fed}} \times \frac{100}{1}$$

SC = stoichiometric factor = ratio of stoichiometric moles of product to reactant = 1

Stoichiometric Equation of the Reaction:

Triglycerides => Ethyl propanoate (FAEE) + Stearic acid + Linoleic acid + ketenes

$$0.95 = \frac{0.773}{\text{Kmole of palm oil converted}} \times \frac{1}{1}$$

$$\text{Kmoles of palm oil converted} = \frac{0.773}{0.95} = 0.814\text{kmol/hr}$$

$$\text{Triglyceride fed} = \frac{0.814}{0.85} = 0.958 \text{ kmoles}$$

Kmoles of Stearic acid produced = 0.773, Kmoles of Linoleic acid produced = 0.773

Kmoles of ketene produced = 0.773,

Kmoles of triglyceride unconverted = 0.958-0.814 = 0.114

Mass of triglyceride fed to the reactor

Kmoles fed = 0.958, Molar mass = 846.26

Mass of Triglyceride fed = 0.958 x 846.26 = 810.72kg/hr

Mass of triglyceride unconverted = (0.958-0.814) x 846.26 = 121.86kg/hr

Calculation of Mass of Ketene (Octadecatrien-1-ol) Produced

From the stoichiometric equation, 1 kmole of ketene is produced along with 1 kmole of FAEE, 1 kmole of Stearic acid and 1 kmole of Linoleic acid.

Hence, total kmoles of ketene produced = 0.773

Molar mass of ketene = 223.42kg/kmol, Mass of ketene = 0.773 x 223.42= 172.72kg/hr

Calculation of Mass of Stearic Acid Produced

From stoichiometric equation, 1 kmole of stearic acid was produced.

Hence total kmol of stearic acid produced = 0.773

Molar mass of stearic acid = 284.48kg/kmol, Mass of Stearic acid = 0.773 x 284.48 = 219.9kg/hr

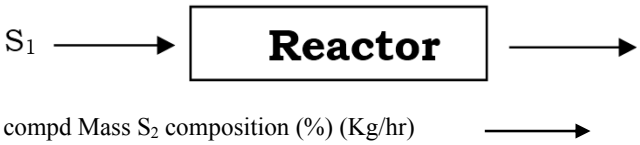
Calculation of Mass of Linoleic Acid Produced

From the stoichiometric equation, 1 kmole of Linoleic acid produced.

Hence total kmoles of linoleic acid produced = 0.773

Molar mass = 280.45kg/kmol., Mass of Linoleic acid = 280.45 x 0.773 = 216.79kg/hr

Material Balance Summary



Inlet Stream

Triglyceride (Mass: 810.72kg/hr, Composition: 100%)

Outlet Streams

Compound	Mass (kg/hr)	Composition (%)
Ethyl Propanoate (FAEE)	78.95	9.74
Stearic acid	219.9	27.14
Linoleic acid	216.79	26.76
Ketenes	172.7	21.32
Triglyceride Unconverted	121.86	15.04
Total	810.2	

Table 1: Summary of material balance on reactor

	Reactor		Quantity	
Stream	**Component**	**Mol. wt Kg/kmol**	**Kmole/hr**	**Kg/ hr**
S_1 (Input)	Triglyceride	846.26	0.958	810.72
		ΣS_1		810.72
S_2 (Output)	FAEE (Ethyl propanoate)	102.13	0.773	78.95
	Stearic acid	284.48	0.773	219.9
	Linoleic acid	280.45	0.733	216.79
	Ketenes	223.42	0.773	172.7
	Triglyceride Unconverted	846.26	0.144	121.86
		ΣS_2		810.2

Input, ΣS_1 = 810.72 kg/hr, Output, ΣS_2 = 810. 2 kg/hr

For Fractionating Column

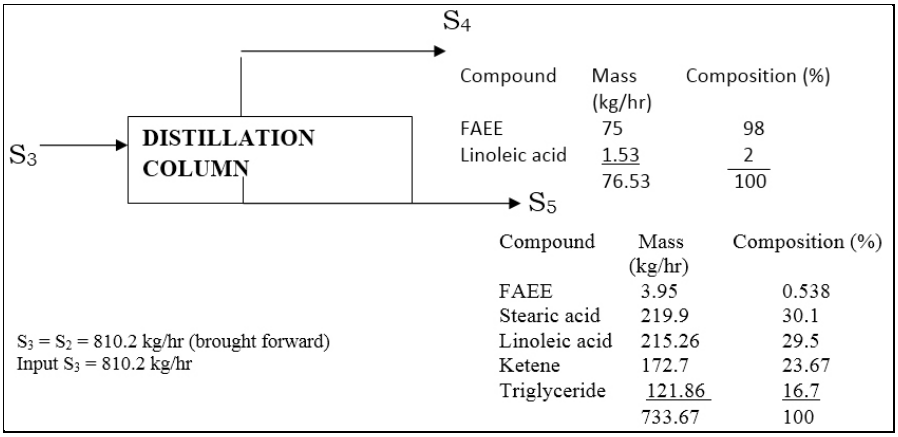
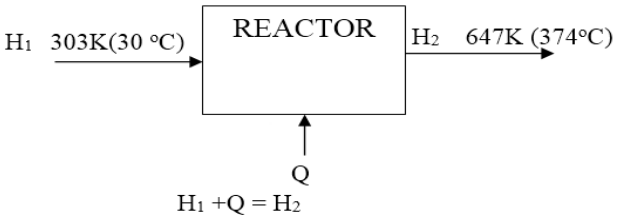


Table 2: Summary of material balance on distillation column

		Mol. Wt	Quantity	
Stream	**Component**	**Kg/kmol**	**Kmole**	**Kg/hr**
S_4	FAEE	102.13	0.734	75
	Linoleic acid	280.45	0.00546	1.53
				76.53
S_5	FAEE	102.13	0.0387	3.95
	Stearic acid	284.48	0.773	219.9
	Linoleic acid	280.45	0.768	215.26
	Ketene	223.42	0.773	172.7
	Triglycerides	846.26	0.144	121.86
				733.67
Output = $\Sigma S_4 + \Sigma S_5$ = 810.2 kg/hr				

4.2 Energy Balance



Assumptions

1. Steady State Operation
2. Datum Temperature

Hence, the input to the reactor

Triglyceride = 810.72kg

$H_1 = MCpT_1$, = 810.72 x 2425.78 x 303 = 595888.39KJ/hr = 595.888 MJ/hr

Effluents, from the reactor

Ethyl propanoate = 78.95kg/hr, Stearic acid = 219.9kg/hr, Linoleic acid = 216.79 kg/hr

Ketene = 172.7kg/hr, Triglyceride uncovered = 121.86 kg/hr

$Q = H_{out} - H_{in}$

Heat output, $H_{out} = \Sigma mc_pT = (mc_pT) + (mc_pT) + (mc_pT) + (mc_pT) + (mc_pT)$ + Ethyl propanoate + stearic acid + Linoleic acid + Ketene Unconverted Triglyceride

a. Head output for Ethyl propanoate = mc_pT = 78.95 x 1.920 x 647 = 98074.848 = 98.074KJ/hr
b. Heat output for Stearic acid = mc_pT = 219.9 x 2359.42 x 647 = 335687.183KJ/hr
c. Heat output for Linoleic acid = mc_pT = 216.79 x 2884.27 x 647 = 404556.738KJ/hr
d. Heat output for Ketene = mc_pT, H_{out} = 172.7 x 3.568 x 647 = 398.677KJ/hr
e. Heat output for unconverted Triglyceride = mc_pT = 121.86 x 2425.78 x 647 = 191256.791KJ/hr

Total Heat output = 931.997MJ/hr

$Q = H_2 - H_1$ = 931.997 - 595.888 = 336.109 MJ/hr

Heat added to the reaction for cracking of the palm oil = 336.109MJ/hr = 336109KJ/hr

$$\text{Power} = \frac{336109 \times 10^3}{60 \times 60} = 93363.6\text{W} = 93.3636\text{ KW}$$

Energy Balance Summary

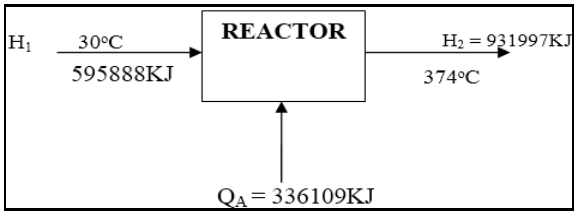
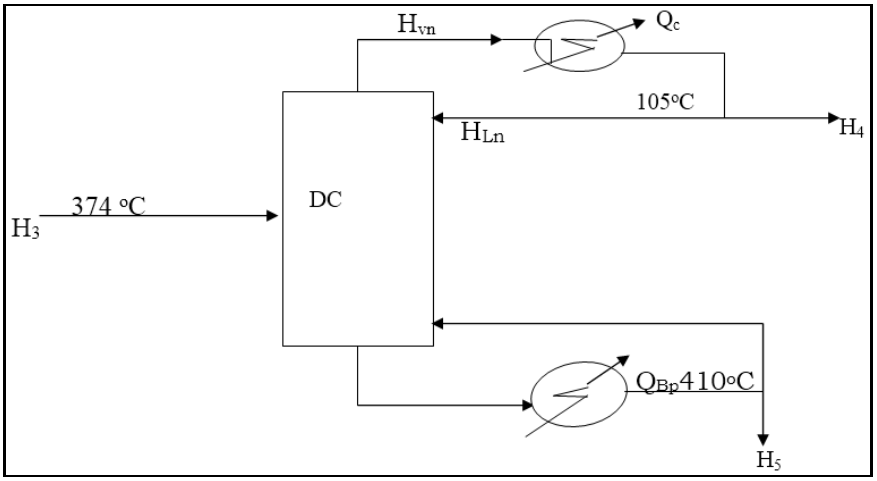


Table 3: Summary of energy balance on reactor

Input	Output
H_1: 595888	H_2: 931997
Q: 336109	-
931997	931997

Energy around Distillation (Fractionating) Column



For effective separation, care is taken in fixing the temperature at the bottom and top of the column.

Determination of top and Bottom product temperatures

Table 4: Boiling points of some fatty acids and ester

Compound	Boiling Point
Ethyl propionate (propanoate)	98.9°C
Linoleic acid	230°C
Stearic acid	383°C
Ketene (octadecatrien-1-ol)	374.227°C
Triglyceride	818.7°C

Boiling point of mixture, $B_{Pm} = \Sigma X_m B_{Pc}$

Where B_{Pm} = boiling point of mixture, X_m = mass fraction of each component

B_{Pc} = boiling point of each component

Boiling point of bottom product mixture, B_{Pmb}

B_{Pmb} = 0.301(383) + (0.295 x 230) + (0.237 x 374.23) + (0.167 x 818.7) = 408.55 °C

Boiling point of top product mixture, B_{Pmt}

B_{Pmt} = (0.98 x 98.9) + (0.02 x 230) = 101.52 °C

Specific heat capacity of component using several group contributions methods by Riham and Doraiswamy (1965) [12] for organic compounds Coulson and Richardson vol.6

$Cp = a + bT + CT^2 + dT^3$

Enthalpy of the feed, $H_3 = H_2$ = 931997. 463kJ/hr

Enthalpy of the distillate $H_4 = \sum MCpdT = \Delta \sum XdCp\Delta T$

Temperature of 105 °C

F = D + B, F = 810.2kg/hr, D =76.53kg/hr, B=729.72kg/h

Table 5: Summary of specific heat capacity calculation for top products

Compound	m_i (kg/hr)	$X_d = m_i/m_T$	Cp (ks/kgk)	ΔT(k)	$x_d C_p \Delta T$ KJ/kg.hr
Ethylpropanoate	75	0.98	1.487	75	8197.088
Linoleic acid	1.53	0.02	1.779	75	4.083
					8201.171

Top column Temperature

The temperature at the top of the column is to be above the top products boiling point B_{pm} = 101.52 ^oC but below the entry point of the feed, which is 374 ^oC, so far effective separation, 105 ^oC top column/ product

Temperature is chosen

H_4 = 76.53 x 8201.171 = 627635.62kJ = 627.635MJ/hr

Enthalpy of liquid reflux, if reflux ratio of 2 is assumed

$R = L_n/D$, $L_n = RD$, $L_n = 2D = 2 \times 76.53 = 153.06$kg/hr

$V_n = L_n + D = 153.06 + 76.53$: $V_n = 229.59$kg/hr

Hence, $H_{LN} = L_n \sum X_d C_p \Delta T = 153.06 \times 1.492 \times (410 - 105) = 69651.4896$kJ/hr

$H_{vn} = \sum mcp \Delta T + V_n \lambda_{av}$

Where λav = average latent heat of vaporization = $\sum xf \lambda_i$

Table 6: Summary of latent heat of vaporization calculation of bottom product

Compound	X_f	λ_i (KJ/kg)	$X_f \lambda_I$ (kJ/kg)
Ethyl propanoate	0.0974	96.837	9.4319
Linoleic acid	0.2676	82.011	21.946
Stearic acid	0.2714	134.63	36.54
Ketene	0.2132	164.5	35.07
Triglyceride	0.1504	96.74	14.55
			117.5309

From Coulson and Richardson volume 6

λ_v = constant

$\overline{T_b}$

λ_v = Latent heat of vaporization, kJ/kmol, T_b = normal boiling point, K

For organic liquids the constant can be taken as 100

$Hv_n = 8568.5 + 117.5309 = 8568.031$KJ/hr

$Hv_n = Q_C + H_4 + H_{LN}$, where $Q_C = Hv_n - H_4 - H_{Ln}$

$= 8686.031 - 627635.62 - 69651.4896 = -688719.07$KJ/hr, $Q_C = -688.719$ MJ/hr

Enthalpy of bottom, $H_5 = B\sum X_b C_p \Delta T$

T = 410 ^oC (683k)

Table 7: Summary of specific heat capacity calculation for bottom products

Compound	m_i (kg/hr)	x_b	Cp (kT/Kg)	ΔT (k)	$\sum X_b Cp \Delta T$ (Ks/kg)
Stearic acid	219.9	0.301	2.935	380	73821.6
Linoleic acid	215.26	0.295	2.747	380	66286.88
Ketene	172.7	0.2367	3.53	380	54833.87
Triglyceride	121.86	0.167	2.45	380	18946.27
					213888.62

$H_5 = 729.72 \times 213888.62 = 156078808.8$kJ/hr

Taking overall Energy Balance

$H_3 + Q_B = H_4 + H_5 + Q_C$, $Q_B = H_4 + H_5 + Q_C - H_3$

$= 627635.62 + 156078808.8 - 688719.07 - 931997.463$

Q_B =15646316KJ/h, Q_B =156463.156MJ/hr

Fractionating Column

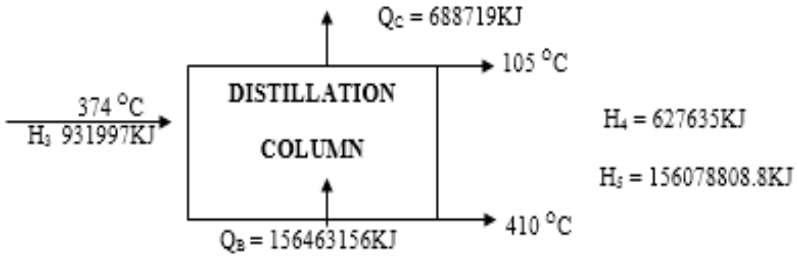


Table 8: Summary of energy balance across distillation column

	Input stream (KJ/hr)	Output stream (KJ/hr)
$H_3=H_2$	931997	-
H_4	-	627635
H_5	-	156078808.8
Q_C	-	688719
Q_B	156463156	-
	157395162	157395162.8

4.3 Equipment design calculations for the reactor

A) Detailed Design of Fractionating Column

From Material Balance

Table 9: Summary of overall material balance

	Feed			Top	Bottom	
Component	**Mass (kg)**	**Composition (%)**	**Mass (kg)**	**Composition (%)**	**Mass (kg)**	**Composition (%)**
FAEE	78.95	9.74	75.0	98.0	3.95	0.538
Ketenes	172.7	21.32	-	-	172.7	23.67
Stearic acid	219.9	27.14	-	-	219.9	30.1
Linoleic acid	216.79	26.76	1.53	2	215.26	29.5
Triglyceride	121.86	15.04	-	-	121.86	16.7
	810.2	100	76.53	100	733.67	100

NB: Top is assumed to be the last plate in the column while bottom is the first plate.

Overall material Balance

F = D + B

F = Mass of feed mixture, D = mass of top product, B = mass of bottom product

FAEE is assumed to be the light key while linoleic acid is high key

Taking the composition material balance over FAEE, $Fx_{f,\ A} = Dx_{d,\ A} + Bx_{b,\ A}$

Where $X_{F,\ A}$ = mass fraction of FAEE at the feed

$X_{d,\ A}$ = mass fraction of FAEE at the top, $X_{b,\ A}$ = mass fraction of FAEE at the bottom

Number of stages

Calculating the minimum number of stages using Fenske Equation

$$\left[\frac{N_m = \log (X_A/X_B)_D - (X_B/X_A)_B}{\text{Log } X_{AB}} \right]$$

X_{AB} = relative average volalility of FAEE – linoleic acid

X_B = mass fraction of linoleic acid, X_A = mass fraction of FAEE

$X_{A,b} = 0.00538$, $X_{B,b} = 0.295$

Relative volatility, $X = \frac{\text{Vapour pressure of FAEE}}{\text{Vapour pressure of linoleic acid}}$

N_m = 11 plates (stages)

Reflux Ratio

From separation process principles by Seader and Ernest J. Henley (2006) [13], we assumed internal reflux

$$\left[R_m = 1 \frac{(X_{d,,A}/X_{f,A}) - X_{AB}(X_{d,B}/X_{f,B})}{(X_{AB} - 1)} \right]$$

$X_{f,A}$ = mass fraction of FAEE in the feed mixture = 0.0974

$X_{f,B}$ = mass fraction of linoleic acid in the feed mixture = 0.2676

$X_{d,A}$ = mass fraction FAEE at the top = 0.98

$X_{d,B}$ = mass fraction of linoleic acid at the top = 0.02

$R_m = 8.76$, Assuming $R = 1.5\ R_m = 1.5 \times 8.76 = 13.14$

Actual Number of stages/ plates

From Gilliland correlation

$(R - R_m) / (R + 1) = (13.14 - 8.76) / (8.76 + 1) = 0.449$

Using the graph showing the empirical relation between reflux ratio R and number of plates n, in which only the minimum reflux ratio and the number of plates at total reflux n_m are equal. Fig. 11.42, Coulson and Richardson's vol.2

However, tracing 0.449 to number of plate's axis, we have 0.3

i.e. $N\text{-}N_m / N + 2 = 0.3$

N = 17 plates (Actual number of plates = 17 plates)

Feed point

To get the feed point, we use kirkbride's empirical equation [1]

$\text{Log } (Nr/Ns) = 0.206 \log [(B/D) * (X_{f,B}/X_{f,A}) * (X_{b,A}/X_{d,B})^2]$

This equation is used considering the bottom product to be at the bottom and top product at the last plate

B = mass of bottom product = 733.67 kg/hr

D = mass of top product = 76.53 kg/hr

$X_{f,B}$ = mass fraction of linoleic acid in the feed mixture = 0.2676

$X_{f,\ A}$ = mass fraction of FAEE in the feed mixture = 0.0974

$X_{b,\ A}$ = mass fraction of FAEE in the bottom product = 0.00538

$X_{d,\ B}$ = mass fraction of linoleic acid in the top product = 0.02

$\text{Log } (N_r/N_s) = N_r = 1.14N_s$

The number of stages excluding the reboiler (bottom)

$N_r + N_s = 16$, $1.14N_s + N_s = 16$

$N_s = 8$ plates

Feed point is at the 8 plates from the bottom

Column Diameter

The principal factor that determines the column diameter is the vapour flow- rate. The vapour velocity must be below that which would cause excessive liquid entrainment or a high pressure drop. The equation given below, this is^based on the well-known Sounders and Brown equations. Lowenstein (1961) [14], can be used to estimate the maximum allowable superficial vapour velocity and hence the column area and diameter.

$U_v = (-0.171L_t^2 + 0.27L_t - 0.047)(\rho_L - \rho_v / \rho_v)^{1/2}$

Where U_v^ = Maximum allowable velocity vapour based on the gross column cross-sectional area, m/s, L_t = plate spacing, in (range 0.15- 0.6) m

Assuming 0.5m spacing

ρ_v = Density of FAEE = 0.884g/cm^3 = 884.4kg/m^3

ρ_L = Density of linoleic acid = 0.90g/cm^3 = 900kg/m^3

$U_v = (-0171 \times 0.5^2 + 0.27 \times 0.5 - 0.047)(900\text{-}884.4/900)^{1/2}$

U_v^ = 0.04525 x (0.132)

U_v^ = 0.00597m/s

Diameter of the column, $D_L = D^{\wedge}_C = 0.095m = 0.1m$

Area = $\frac{\pi d^2}{4} = \frac{3.14^2 \times 0.1^2}{4} = 0.0078m^2$ $\sqrt{\frac{4V_{w,}}{\pi\rho_v U_v}}$ ^ U_v

Column Area, $A_c = 0.00786m^2$

Determination of Column Height

Height of the column, H_c (N_{real}-1) x I_p x Disengagement height

Where H_C = Column height

I_p = plates or stage spacing

From McCabe and Smith the top & bottom disengagement height for distillation column are 4ft at top and 6ft at bottom.
If plates spacing is 0.5, 1foot = 0.3048 metres
H_c = (17-1) x 0.5 + 10 (0.3048)
H_c = 11.55m

Provisional Plate Design for Sieve Tray
Column diameter, D_c = 0.1m
Column Area, A_c = 0.00786m^2
Downcomer area = 12% of column area (Coulson & Richardson Vol.6)
A_d = 0.12 + 0.00786 = 0.000943m^2
Net area, $A_n = A_c - A_d$= 0.00786 - 0.000943 = 0.00692m^2
Active area, $A_a = A_c - 2A_d$ = 0.0078 - 2(0.000943) = 0.00597m^2, A_a = 0.00597m^2
From Coulson and Richardson's Vol.6
Assume the Hole area, A_h as 10% of active area, A_a at first trial
A_h = 0.1 x 0.00597 = 0. 000597m^2

Weir Length
(From figure 11.31 Coulson & Richardson)
l_w = 0.76 x 0.1 = 0.076m

Weir Height
The height of the weir determines the volume of liquid on the plates and is an important factor in determining the plate efficiency.
A high weir will increase the plate efficiency but at the expense of a higher plate drop. For column operating above atmospheric pressure, the weir height will normally be between 40mm to 90mm; 40mm to 50 is recommended [1].
Weir height = 50mm**,** Hole diameter = 5mm**,** Plate thickness = 5mm

Check Weeping
D = 76.53kg/ (8 x 60 x 60) = 0.00266kg/s
$V_n = D(1+R)$ = 0.00266 (1+13.14) = 0.0376kg/s
$L_n = DR$ = 0.00266 x 13.14 = 0.035kg/s
F = 0.028kg/s
$L_m = L_n + F$ = 0.035 + 0.028 = 0.063kg/s
$V_n = L_m - B$ = 0.063 – 0.0255 = 0.0375kg/s
L_n = Liquid flow rate above feed composition, V_n = Vapour flow-rate above feed composition
L_m = Liquid flow rate below feed composition, V_m = Vapour flow-rate below feed composition
Maximum liquid rate = 0.063kg/s
Minimum liquid rate at 70% turn down = 0.7 x 0.063 = 0.0441kg/s
Maximum height of the liquid crest over the weir can be estimated using the Francis weir formula, for a segmented dowcomer. This can be written as:

$$h_{ow} = 750 (L_m/\rho_L I_w)^{2/3}$$

Where I_w = weir length, m, h_{ow} = weir crest, mm liquid
L_m = liquid flow rate (kg/s), ρ = liquid density = 900kg/m^3

$$\text{Minimum, } h_{ow} = 750\left[\frac{0.0441^{2/3}}{900 \text{ X } 0.076} = 5.6\text{mm}\right]$$

$$\text{Maximum, } h_w = 750\left[\frac{0.063^{2/3}}{900 \text{ X } 4 \text{ x } 0.076} = 7.10\text{mm}\right]$$

At minimum rate $h_w + h_{ow}$ = 7.10 +5.6 = 12.70mm liquid
From fig 11.30 Coulson and Richardson vol. 6 weep correlation
k_2 = 27.0

$$U_n\text{^(min)} = \left[\frac{k_2 - 0.90(25.4 - d_h)}{(\rho_v)^{1/2}}\right]$$

Where U_n = minimum vapour velocity through the holes based on the hole area m/s d_h = hole diameter, K_2 = constant

$$U\text{^}_n = 27 - \frac{0.90(25.4 - 5)}{(884.4)^{1/2}} = 0.291\text{m/s}$$

$$\text{Actual minimum vapour velocity} = \frac{\text{minimum vapour rate}}{A_h}$$

Maximum volumetric flow-rate base = V_m/ρ_L = 0.0375/900 = 4.17 x 10^{-5} m/s

$$\text{Actual minimum vapour velocity} = \frac{0.7 \text{ x } 4.17 \text{ x} 10^{-5}}{0.000597} = 0.049\text{m/s}$$

So minimum operating rate will be well above weep point

Plate Pressure Drop, Dp_t
Dry Plate drop
Maximum vapour velocity through holes, U_h = 4.17x10^{-5}/0.0005 = 0.0698m/s
From fig 11.34, Coulson and Richardson's vol.6 for plate thickness/ hole diameter = 1 and $(A_h/A_p) = (A_h/A_a)$ = 0.1
Orifice coefficient, C_O = 0.84

$$h_d = 51 (u_h/C_O)^2 (\rho v/\rho_L) = 51 \text{ x } (0.0698/0.84)^2 \text{ x } (884.4/900)$$
$$= 0.346\text{mm liquid}$$

$$\text{Residual head, } h_r = \frac{12.5 \text{ x } 10^3}{\rho_L} = \frac{12.5 \text{ x } 10^3}{900} = 13.89\text{mm liquid}$$

Total plate pressure drop, $h_t = h_d + (h_w + h_{ow}) + h_r$ = 0.346 + (5.6 + 7.10) + 13.89
h_t = 26.936mmliquid
$\Delta P_t = \rho g h_t$ = 237.8Pa, Δp_t = 0.238 kPa

Downcomer Liquid Back-Up
The downcomer area and plate spacing must be such that the level of the liquid and froth in the downcomer is well below the top of the outlet weir on the plate. If the level rises above the outlet weir the column will flood. The back-up of liquid in the downcomer caused by the pressure drop over the plate (the downcomer in effect forms one leg off a U-tube and the resistance to flow in the downcomer itself)

Downcomer Pressure Loss
Take $h_{ap} = h_w$ – (5 to 10mm)
Where h_{ap} = height of the bottom edge of the apron above the plate. This height is normally set at 5 to 10mm below the outlet weir height
$h_{ap} = h_w - 10$ = 50 -10 = 40mm
Area under, $A_{ap} = h_{ap} I_w$
Where I_w = weir length = 0.076m
A_{ap} = 0.76 x 40 x10^{-3} = 0.00304m^2
Since A_{ap} = 0.00304m^2 is less than A_d in equation below
$h_{dc} = 166 (L_{md}/\rho_L A_m)^2$
Where h_{dc} = head loss in the downcomer, mm
L_{md} = liquid flow rate in downcomer (below the feed composition), kg/s
A_m = either the downcomer area, A_d or the clearance area under the downcomer A_{ap} whichever is the smallest, m^2(value calculated already from provisional plate design of sieve tray)
L_{md} = 0.063kg/s (value gotten from check weeping)
A_m = 0.000943m^2 (value gotten from provisional plate design)
$h_{dc} = 166[0.063/(900 \text{ x } 0.000943)]^2$ = 0.915mm = approximately 1mm

Back-up in downcomer
$h_b = (h_w + h_{ow}) + h_t + h_{dc} = (50 + 7.10) + 26.94 + 1$
$h_b = 8.504mm = 0.08504$
$0.08504 < ½$ (plate spacing + weir height)
So plates spacing is acceptable
Check residence time

$$t_r = \frac{A_{dc} \times h_{bc} \times \rho_L}{L_{md}} = \frac{0.000943 \times 0.08504 \times 900}{0.063} = 1.15s$$

The residence time in Downcomer, $t_r = 1.15s$

Perforated Area
The area available for perforation will be reduced by the obstruction caused by the structural members (the support rings and beams), and by the use of calming zones.
Calming zones are unperforated strips of plates at the inlet and outlet sides of the plate. The width of each zone is usually made the same; recommended values are below 1.5m diameter, 75mm above, 100mm.
The width of the support rings for sectional plates will normally be 50 to 75mm; the support ring should not extend into the downcomer area. A strip of unperforated plate will be 6ft round the edge of cartridge type trays to stiffen the plate.
The unperforated area can be calculated from the plate geometry. The relationship between the weir chord length, chord height and the angle subtended by the chord is given in fig 11.32, Coulson and Richardson Vol.6 from
Figure 11.32, at $I_w/D_c = 0.076/0.1=0.76$, $\theta_c = 99\ ^0C$
Angle subtended by the edge of the plate = $180-99 = 81^o$
Mean length, unperforated edge strips = $(0.1\text{-}50 \times 10^3) \times (81/180) = 0.0707$
Area of unperforated edge strips = $50 \times 10^{-3} \times 0.0707 = 0.00354m^2$
Mean of Calming zone = weir length + width of unpreforated strip = $0.076 + 50x10^{-3} = 0.126m$
Area of calming zone = $2\ (0.126 \times 50 \times 10^{-3}) = 0.0126m^2$
Total mean for perforation, $A_p = 0.00397 – 0.00334 = 0.00063m^2$

Hole Pitch
The hole pitch (distance between the hole centre) I_p should not be less than 2.0 hole diameter and the normal range will be 2.5 to 4.0 diameter within the range the pitch can be selected to give the number of active holes required for the total hole area specified.
Ratio of hole area to perforated area = $(A_h/A_p) = (0.00597/0.00597) = 0.1$
From fig 11.33, Coulson and Richardson Vol. 6 $L_p/ d_h = 2.75$
This is satisfactory, within 2.5 to 4.0

$L_p = 2.75d_h = 2.75 \times 5 = 13.75mm$
Hole pitch = 13.75mm

Hole pitch arrangemented selected = Equilateral Triangular
Downcomer selected = segmental or chord
This is due to its simplicity and cheapness. The apron is usually vertical, but may be sloped to increase the plate area available for perforation.

Number of Holes on the Plate
Number of holes = hole area/area of one hole
Area of hole $= \pi d_h^2 = \frac{3.142}{4} \times \frac{(5x10^{-3})^2}{4} = 1.964x10m^2$

Number of holes $= 0.000597/1.964x10^{-5} = 30.4$ holes = **31holes**

Fractionating Column
The column consists of a cylindrical structure divided into sections by series of perforated trays or plate (sieve plate), which permit the upward flow of vapour. The liquid reflux flows across each tray, over a weir and a down comer to the tray. The vapour rising from the top tray passes to a condenser and then through an accumulator or reflux drum and a reflux divided, where part is withdrawn as the overhead product and the remainder is returned to the tray as reflux.

Table 10: Summary of Fractionating Column Design

Minimum no. of plates:	11 plates
Actual no. of plates:	17 plates
Feed point:	8 plates
Column diameter:	0.1m
Area of column:	$0.00786m^2$
Column Height:	11.55m
Weir Height:	50mm (0.5m)
Hole diameter:	0.05m
No. of holes per plate:	31 holes
Plate thickness:	0.05m
Hole pitch:	13.75mm
Hole pitch arrangement:	Equilateral triangle
Column wall thickness:	4.0463mm
Type of plate:	Sieve plate

Conclusion
The design for a distillation column follows an appropriate methodology based on the McCabe-Thiele method of column design. Other criteria employed to take into account are the type of plate and number of plates. The number of plates required is 17 plates with column diameter and height of 0.1m and 11.55m respectively. The study shows the importance of a distillation column in cracking of palm oil which buttresses the point that complete conversion of triglyceride to oleochemicals can be achieved with the aid of a fractionating column.

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