Sizing of Major Equipment

Sizing of Storage Tank

Assumption:

1) The Storage vessel can store 47% NaOH Solution for 18 days raw material in a basis of 5 batch / day.
2) The vessel should not be filled beyond 75% of the total volume.
3) The vessel is operating under atmospheric condition.

Solution:

Basis: 638 Kg Solution NaOH / hr (47 wt %) is required per batch of soap production. Per batch requires 4 hr to complete the saponification reaction process.

So, In one day, We assumed only 5 batch operations can be done.

Density of 47% NaOH ($\rho_{47\%\ NaOH}$) = 1497 Kg/M^3 (Avg Temp = 25°C in Chittagong)

Volume of Required Storage Tank(V) = 638 Kg/hr $\times$ 4hr/batch$\times$5Batch/day$\times$(18day)$\times$

$$(1M^3/1497Kg) \times (1/0.75) = 204\ M^3$$

$$= 204\ M^3 \times[(3.2808ft)^3/(1\ M)^3]\times(7.4808gallon/1ft^3)\times(1bbl/42gallon)$$

$\therefore$ **V =** 1283bbl

Diameter suitable for this volume is given from Chart(Ref1)

ID = 21.5 ft = 6.55M

$$V = \pi/4\times(ID)^2 \times h$$

$$204\ M^3 = (\pi/4)\times(6.55)^2 \times h$$

$$\therefore \quad h = 6.05\ M = 20\ ft.$$

Sizing of pump p . 101

Caustic soda feed pump:

Pump specification Data:

Flow rate, G= 638 kg/ h

Density, p= 1.47 g/$_{em}^{3}$= 1.47×1000 kg/ $_{m}^{3}$

Viscosity,µ = 1.5×w^{3} kg/ m.s

Temperature, T= 30.c

Sizing of pump:

Basis: 1 kg of flowing fluid

Total Mechanical energy (ME) Balance bet^{n} 1 all 2

$$w_o = (z_2 - z_1) \, {}^{g}\!/_{g_c} + \frac{v_2^{\,2}}{2\times g_c} - \frac{v_1^{\,2}}{2\times g_c} + \frac{p_2}{r_2} - \frac{p_1}{r_2} + \left[\sum f + (wpsh)\, {}^{g}\!/_{g_c}\right]$$

Here V_1 = V2 [incompressible fluid]

So the ME balance becomes,

$$w_o = (\Delta Z)\, {}^{g}\!/_{g_c} + \frac{\Delta p}{p} + (NPSH)\, {}^{g}\!/_{g_c} + \sum F$$

(i) Head developed Due to Elevation:

$$(\Delta Z)\, {}^{g}\!/_{g_c} = 5M\left(9.8\frac{N}{g}\right) = 49 Nm/kg$$

(ii) Pressure Head.

Here DP = 18.915-5

= 13.912 psia

= $95.85\times10^{3}\ {}^{N}/_{m2}$

p= 1470 kg/m^3

$$\frac{\Delta P}{r} = \frac{95.85\times10^{3}}{m^{2}}\left|\frac{m^{3}}{1470kg}\right.$$

= 65.20 N^m/kg

(ll) Head Due to friction lassess ($\sum$ F)

(a) Loss in pipe

(b) Piping specification

(c) Diameter of the pipe, D_p= 3 inch

Choice of pipe:

Nominal pipe size IPS= 3 in.

OD= 3.5

Schedule no.= 40

ID= 3.068 in = 77.92×10^{-3}m

Weight = 7.58 (Ibm steel/liner feed)

pipe cross section area, A=$\pi/4$x$(ID)^2$

=$\pi/4$x$(77x10^{-3})$

=$4.769x10^{-3}\ m^2$

Normal and design fluid velocities:

Normal fluid velocity u=$\frac{\Phi}{A}$

Here Φ =G/ρ

$$= \frac{638\text{kg}}{\text{h}}\bigg|\frac{m^3}{1470kg}\bigg|\frac{1h}{3600s}$$

= 1.25×10^{-4} m^3/s

$$u = \frac{\Phi}{A} = \frac{1.205\times10_{-4}}{4.769\times10^{-b}} = 0.252 \quad \text{m/s}$$

As a design practice, 20% above the normal flow is considered.

$$,u_{man} = (1.2)^2 u$$

[Reference: Coulson ;vol^m- 6]

= 0.364 m/s

Friction loss in pipe line

Reynolds No. $\text{Re} = \frac{Duman^P}{u}$

$$= \frac{77.92\times10^{-3}\times0364\times1470}{1.5\times10^{-3}}$$

=27795.62

Absolute roughness for steel pipe = 0.00015 ft

$$\text{Relative roughness} = \frac{e}{\text{D}} = \frac{.00015\times12}{9.068}$$

= 5.86×10^{-4}

From funning friction factor chart [Peter & Timmerhaus p.482 Pig 14-1

f= 0.006

Lenght of the pipe line= 40 m

$$Frictioallos\sin\ pipe, F_1 = \frac{2f\ln on^2 L}{g_c D}$$

$$F_1 = \frac{2\times.006\times(.364)^2\times 40}{1\times 77.92\times 10^{-3}}$$

=0.816 n-m/kg.

(b) Loss in fitting and valves.

There are two gate valves and four (4) elbows, std radius

From peter & Timerhause (p.484 Table 1)

Equivalent length of gate valve= 7 D= $7\times77.92\times W^{-3}$

= 0.545 m

Elbow eguivqlent length = 32 D

= $32\times77.92\times10^{-3}$

=2.5m.

Total line eduivalent lenght

L_e= $2\times.545+4\times2.5$

= 11.063m.

Loss in fitting and valve, $F_2 = \frac{2fu_m{}^2 L_e}{g_c D}$

$$F_2 = \frac{2\times.006\times(.364)^2\times 11.063}{77.92\times 10^{-3}}$$

= 0.225 N-m/kg

Total frictional losses,

$\sum F$ = F_1+F_2

= .816+.225 N-m/kg

= 1.041 N-m/kg

(iv) NPSH calculation:

$$\text{Volumeric flow rate} = \frac{638kg}{m} \times \frac{m^3}{1470kg}$$

$$= .434\ m^3/_h$$

As a general guide, the NPSH should be above 3 m for pump capacities upto fat flow rate of 100 m^3/h

NPSH = 3 m

(NPSH) xg/g_c= 3×9.81=29.4N-m/Kg

eq^n (i) ME Necessary for pump.

$$w_o = (\Delta z)\, {}^{g}\!/_{g_c} + \left(\sum F + NPSH\, {}^{g}\!/_{g_c} + \frac{\Delta p}{r}\right)(1.2)^2$$

[20% above normal flow is taken]

$$w_o = 49 + [65.20 + 1.041 + 29.4](1.2)^2$$

= 186.72N-m/kg.

Power required to drive the pump, p= G w_o

$$\frac{638kg}{h}\left|\frac{1h}{3600s}\right. \times 186.72\frac{N-M}{kg}$$

= 133.09 watts

$$= \frac{133.09}{746} hp$$

= 0.18 hp

Assuming 70% efficiency of the pump i.e η= 0.7

$\therefore$ Total power required=P/η=0.18/0.7

=0.25 h_p.

Here scale factor is five

so, it required=5x0.25

=2.25 h_p

It may be used 3 horse power pump used

SIZING OF KETTELE TYPE PAN :

Assumptions:-

$\Rightarrow$The pan should gave the capacity to store 47% NaOH solution & bleach tallow For one and half hr. of full load production.

⇒Four pan should have two standby & two continuous.

⇒the pan should not be filled beyond 60% of the total height.

⇒It is Cubic shape tank.

⇒The ratio of pan height, width & length is 1:1:1.

⇒It is atmospheric storage tank

Solution:

The flow rate of 47% NaOH solution is 638 kg/hr. & tallow flow 1000kg/hr.

Our material balance basis 1000kg tallow/hr but factory (Liver brother) produce(40 ton/day Toilet+70 ton/day Londy+40 ton/day Life boy)=150 ton/day

In material balance 1000 Kg tallow/hr produce 1250 Kg/hr soap.

So, Scale factor=

=5

∴ flow rate of 47% NaOH sol. = $638 \times 5 = 3190$ Kg/hr flow rate of tallow = $1000 \times 5 = 50000$ Kg/hr

We know, Density of NaOH sol.=1.47 gm/cm^3

Density of tallow=.82 gm/ cm^3

$\therefore$ Volume required for pan=Vol. Of NaOH sol.+ Vol. Of tallow

$$= \frac{3190 \text{ kg}}{\text{hr}} \left| \frac{\text{m}^3}{1.47 \times 1000 \text{ kg}} \right| \frac{1.5 \text{ hr}}{\text{Batch}} + \frac{5000 \text{ kg}}{\text{hr}} \left| \frac{\text{m}^3}{820\text{kg}} \right| \frac{1.5 \text{ hr}}{\text{Batch}}$$

= 3.255+9.146 m^3/Batch

=12.40 m^3/Batch

In this section is batch wise but next section is continuous. So 3.5 times capacity are designed in each batch because pan preparation (Cleaning+ washing + filling + cooling+ safety times)

$\therefore$ Actual Volume required=12.40×3.5 m^3

$$=43.4 \text{ m}^3$$

Let length of the pan=l m^3

$\therefore$ Volume of pan=$l \times l \times l = l^3$ m^3

$\therefore$ $l^3 \times (1-.4) = 43.4$ (Because 40% allowance)

$l = 4.16$ m

So , height= width = length=4.16 m

Sizing of heating medium

Here, Direct steam injection in the bottom section at 3 bar saturated steam . Steam flowing nominal pipe diameter 2 inch. OD.

For vessel side,

Vessel length, height & width same and it =4.16 m

So, Tube side

Tube OD=2.38 in.

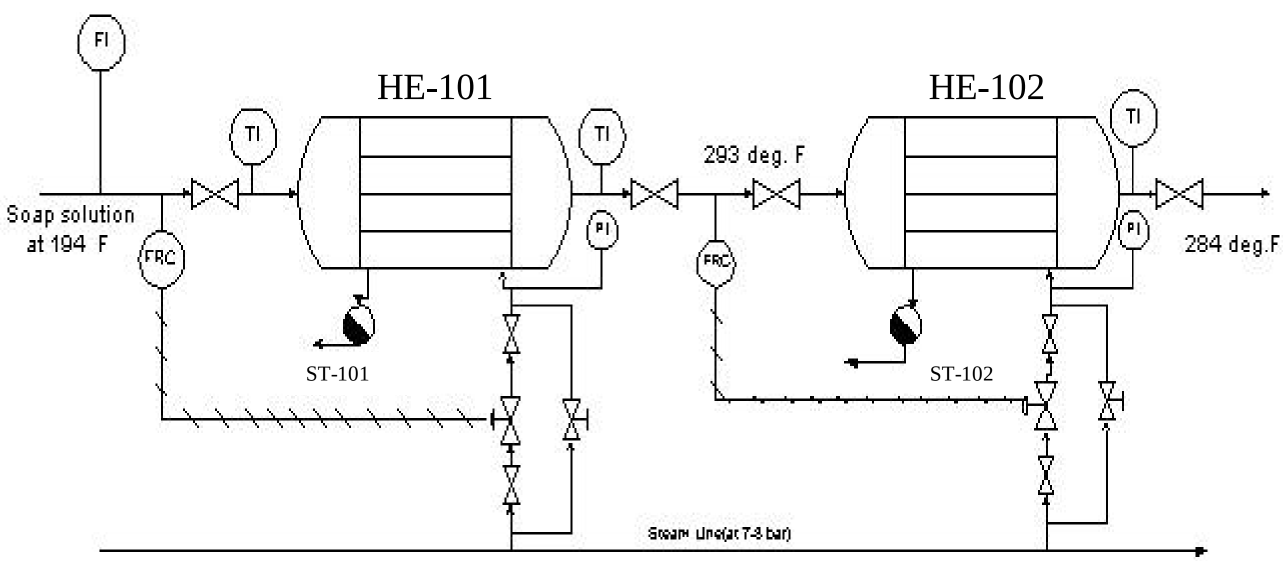
Schedule No. =40

ID =2.067 in.

Tube length $\cong$ vessel length =4.16 m

Surface area =

Sizing of (1-1) Shell and Tube Heat Exchanger:



P & I Diagram of Heat Exchanger Unit

40-ton/day soap produced at 40% capacity.

at 100% capacity 100 ton soap/day must be produced.

$$\frac{100 tonsoap}{day} \times \frac{1000kg}{1ton} \times \frac{1day}{24hr} = 4166.66\frac{kg}{hr}$$

Multiplying factor= $\frac{4166.66}{983} = 4.23387$

Soap solution enter into the heat exchanger=4.23387x1637 kg/hr

=6935.07 kg/hr

≅7000 kg/hr

Specific heat of soap solution,

C_p=2.18108x10^{-3} Btu/gm-^{0}c.

Rate of flow of Soap solution→

m=7000 kg/hr

=7x10^6gm/hr.

Δt =(115^{0}c-90^{0}c)

= 25^{0}c

So, heat load Q= m x C_p x Δt.

Q=381693.9 Btu/hr.

$\boxed{\text{Q=381693.9 Btu/hr.}}$

Approximate Overall heat transfer coefficient:

for heavy organic U_D= 6-60 Btu/hr-ft^2-^{0}F. (Kern, McGraw-HILL INTERNATION EDITION, Table-8; Page-840)

Let assume, U_D= 10 Btu/hr-ft^2-^{0}F.

LMTD Calculation:

Hot Fluid(steam)		Cold Fluid(soap sol^n)	Diff
338^0 F	HigherTemp	139	99 (Δt_2)
338^0 F	LowerTemp.	194	144 (Δt_1)
0	Diff.	45	45

$\therefore$ LMTD= $\left(\dfrac{\Delta t_2 - \Delta t_1}{\ln(\Delta t_2 / \Delta t_1)}\right)$

=120 ^{o} F.

Q= U_D x A x $\Delta T_{LMTD.}$

$\therefore$ A=$\left(\dfrac{Q}{U_D x A_t x \Delta T_{LMTD}}\right)$

=318.07825 $feet^2$.

Now tube size:

OD=3/4, 20 BWG.

$\therefore$ A/L=0.1963 $feet^2$/feet

Tube length=10 feet.

A=0.19693x10.

=1.962 $feet^2$ per pipe.

Total no. of tube is n.

$\therefore$ n= 318.078/1.963

=162

Nearest pipe number is = 185 (Ludwig, 2^{nd} Edition, Vol^m-3, Table 10-9B,page-32)

Shell dia= 18 in .

Squire pitch, pitch length=1 in.

So, corrected U_D= 381693.9/(120x1.963x185).

=8.3587. Btu/hr-feet^2-^{o}F

$$\boxed{\mathbf{U_D = 8.3587.\ Btu/hr\text{-}feet^2\text{-}^oF}}$$

Flow area per unit tube= $\frac{p}{4}\left(\frac{ID}{12}\right)^2$

$$=\frac{p}{4}\text{x}\left(\frac{0.68}{12}\right)^2$$

=2.521731x10^{-3} feet^2/tube.

$\therefore$ Total flow area = 2.52173x10^{-3}x185.

=0.46652 feet^2.

Now Sp. gravity of soap solution is =0.9192.

Density of soap solution is =0.9192x62.4

=57.358 Ib$_m$/feet$^{3.}$

Velocity of soap solution is= $\left|\frac{700kg}{hr}\right|$ x $\left|\frac{2.2Ibm}{1kg}\right|$ x $\left|\frac{1\,feet^3}{57.358Ibm}\right|$ x $\left|\frac{1}{0.46652\,feet^2}\right|$ x $\left|\frac{1hr}{3600s}\right|$

=0.159 feet/sec

Recommended max. Velocity is 2.5 feet/sec (Ludwig, 2nd Edition, Vol^m-3 Table10-23 page 85).

So, Velocity through the pipe is acceptable.

<u>Calculation of tube side heat transfer coefficient</u>:

Tube ID=0.68/12

$=.056667$ feet.

Mass velocity $=m/A$

$$=\left(\frac{15400}{0.46652}\right)$$

Gt =33010.37 Ibm/feet^2-hr.

$$Re_t=\frac{DG}{m}$$

$$=\frac{33010.37\times 0.056667}{7\times 2.42}$$

$=110.$

(Viscosities of Soap solution assume 7 Cp. comparing to coconut oil)

Conductivity of Soap Solution ,k=.0855 Btu/hr –(ft^2/feet)-^{0}F.(Assumption on Palmitic acid, kern; table 4,page-800)

L/D=10/.0566

$=176.4801$

and $Re_t=110$.

$J_h=2$ (Fig10-38 Ludwig, Vol-3,page-73)

$$J_h=\frac{hiD}{k}\ x\left(\frac{cm}{k}\right)^{-1/3}\ x\left(\frac{m}{mw}\right)^{-0.14}$$ (Assuming $\frac{m}{m_w}=1$)

$$2=h_{io}\ x\frac{0.0566636}{0.0855}\ x\left(\frac{.02933\times 7\times 2.42}{0.0855}\right)^{-1/3}$$

$h_{io}=9.250$ Btu/hr-feet^2-^{o}F

$\boxed{\mathbf{h_{io}=9.250\ Btu/hr\text{-}feet^2\text{-}^oF}}$

Tube side Pressure calculation:

Pressure drop due to change in flow direction→

$$\Delta P_r = \frac{4n \times v^2}{s \times 2g \times 144} \text{ Psi.}$$

$$= \frac{4 \times 1 \times .159^2}{.9192 \times 2 \times 32.17 \times 144}$$

$=2.37260\text{x}10^{-5}$ Psi

Pressure drop due to flow→

$$\Delta P_t = \frac{f \times Gt^2 L \times n}{5.22 \times 10^{10} \times Det \times S \times \Phi t}$$

$$\therefore \quad \Phi t = \left(\frac{m}{m_w}\right)^{0.14} = 1$$

at Re_t=110 f=0.005 Kern, McGraw-HILL INTERNATION EDITION, fig-26 Page-836

$$\Delta P_t = \frac{.005 \times 3301037^2 \times 10 \times 185}{5.22 \times 10^{10} \times 0.056636 \times 0.9192 \times 1}$$

=3.7073 Psf

=1.6064 Psi.

$\Delta P_T = \Delta P_t + \Delta P_r$

=1.606521 Psi.

Shell side calculation:

Stationary Tube-sheet Exchanger(Fixed tube-sheet)

Shell ID= 18 inches.

Baffle spacing =5 in.

Square pitch, P_T=1 in.

$C' = P_T - ID_t$

=1-.75

=0.25 in.

The shell side or bundle cross flow area,

$$a_s = ID x \frac{C \times B}{P_T \times 144} \text{ feet}^2$$

$$= 18x \frac{0.25 \times 5}{1 \times 144} \text{ feet}^2$$

$$= 0.1562 \text{ feet}^2$$

Saturated steam at 8 bar(338 ^oF),

Sp. enthalpy=880 btu/Ibm   (From steam table).

$$m_s = \frac{381693.9}{880}$$

$$= 433.743 \text{ Ibm/hr.}$$

$$\therefore G_s = \frac{433.743}{0.15625}$$

$$= 2775.95 \ \frac{Ibm}{feet^2.hr}$$

For square pitch ,

$$De = \frac{4 \times (P_T^2 - \Pi d_o^2 / 4)}{\Pi d_o}$$

$$= \frac{4 \times (1^2 - 3.14 \times 0.75^2 / 4)}{3.14 \times 0.75}$$

$$= 0.95 \text{ in.}$$

$$= .079 \text{ feet.}$$

Physical property of saturated steam at 8 bar (338 ^oF)→

Viscosity of steam, m_s=0.015 Cp   (From fig-15, kern , McGraw-HILL INTERNATION EDITION, page-825)

Conductivity of steam K=.0172 Btu/hr-(feet^2/feet)-^oF (from kern, McGraw-HILL INTERNATION EDITION, page-802)

Sp. heat of steam, Cs=0.45 Btu/Ibm-^oF (from fig.-3 kern page-805)

$Re_s = \frac{D_e G_s}{m_s}$

$Re_s = \frac{.079 \times 2775.95}{.015 \times 2.42}$

$= 6041.32$

So from fig.28 kern page 838,

$J_H = 40$

$$40 = \frac{h_o \times 0.079 \times \left(\frac{0.45 \times 0.015 \times 2.42}{0.172} \right)^{-1/3}}{0.0172}$$

$h_o = 149.7005$ Btu/hr-feet^2-^{o}F.

h_o=149.7005 Btu/hr-feet^2-^{o}F.

Shell Side Pressure Drop:

Shell side friction factor, $f_s = 0.0023$ (From fig.29 kern, McGraw-HILL INTERNATION EDITION, page839)

$D_s = \frac{18}{12}$

$= 1.5$ feet.

No of Cross $N+1 = 12x\frac{10}{5} = 24$

$Re_s = 6041.32$

$$\Delta P_s = \frac{f \times Gs^2 \times D_s (N+1)}{5.22 \times 10^{10} \times D_e \times S \times \Phi_s} \text{ psf}$$

$$= \frac{0.0023 \times 2775.95^2 \times 1.5 \times 24}{5.22 \times 10^{10} \times 0.079 \times 4.25 \times 10^{-3}}$$

$$= 0.3640 \text{psf} \frac{62.4}{144}$$

$$= 0.16618 \text{ psi.}$$

$\therefore \quad \Delta Ps < 2\Delta Psi$

Now,

$$U_C = \frac{h_{io} \times h_o}{h_{io} + h_o}$$

$$= \frac{9.25 \times 149.70}{9.25 + 149.7}$$

$= 8.7173$ Btu/hr-feet^2-^{o}F.

$$\boxed{\mathbf{U_C = 8.7173 \text{ Btu/hr-feet}^2\text{-}^oF.}}$$

$$R_d \text{ Cal} = \frac{U_C - U_D}{U_C \times U_D}$$

$$= \frac{8.71173 - 8.3587}{8.71173 \times 8.3587}$$

$$= 0.0048$$

R_D required= 0.001 (Kern , McGraw-HILL INTERNATION EDITION, page -840)

$\therefore$ R_D calculation> R_D required

So design is satisfied.

Hot Side (Steam)	Cold side (Soap solution)
Saturated steam at 8 bar G_s=2775.95 $\frac{Ibm}{feet^2.hr}$ h_o=149.7 $\frac{Btu}{hr.feet^2.^oF}$ Re_s=6041 ΔP_s=0.16618 psi.	Soap solution (40%) at194 ^oF G_t=33010.37469 $\frac{Ibm}{feet^2.hr}$ h_{io}=9.25 $\frac{Btu}{hr.feet^2.^oF}$ Re_t=110 ΔP=1.606521 Psi.

Shell Side	Tube Side
ID=18 inches Baffle spacing=5 inches Passes= 1	Tube ID=3/4 inches, 20 BWG. Tube no=185 Tube length=10 feet. Square pitch, pitch P_T=1 in.

Sizing of Packed Bed Distillation Column:

For Vapour-Liquid equilibrium curve (Glycerine-Water) at 5mm Hg is generated by the following procedure if not available in Industry :

The vapour Pressure of Glycerine at different Temperature can be found out by Antoine Equation but It is not given . So we have to generate from the following available data:

Pressure	1mm	5mm	10mm	20mm	40mm	60mm	100mm
Temp(°C)	125.5	153.8	167.2	182.2	198.0	208.0	220.1

The vapour Pressure of Glycerine is

T_1=153.8°C = 427K ; Vapour Pressure (P_1*) = 5mm Hg

T_2=162.2°C = 440.35K ; Vapour Pressure (P_2*) = 10mm Hg

So we start evaluating the A,B,C of Antoine equation as follows:

$\log_{10}P^{sat}$ (mm) = A-[B/(T(°C)+C)]

$\therefore$ $\log_{10}$ 5 = A-[B/(153.8°C+C)]

$\therefore$ $\log_{10}$ 10 = A-[B/(167.2°C+C)]

After Simplification, at T= 153.8°C

153.8 A – B – 0.6989C = 107.5 – A.C…………………..Eq^n (1)

Similarly at T = 198.0°C

198.0 A – B – 1.602C = 317.2 – A.C…………………… Eq^n (2)

At T = 220.0°C

220.0 A – B – 2C = 440.2 – A.C……………………….. Eq^n (3)

After Solving these equations We get ,

A= 11.643, B= 5378.68 , C= 337.67

Thus The Antoine equation for Glycerine is given by

$\therefore \log_{10}P^{sat}$ (mm) = 11.643 - [5378.68 /(T(°C)+337.67)] ………………….. Eq^n (4)

The Antoine equation for Water is already given

$\therefore \log_{10}P^{sat}$ (mm) = 7.96681 - [1668.21 /(T(°C)+228)] …………………….. Eq^n (5)

We know, The liquid starts boiling when surrounding environment pressure equals the partial pressure of component (Vapour pressure multiplied by its mole fraction)

So at 5mmHg Pressure of System pressure(absolute Pressure),

Pure Water boils at 1.15°C

And Pure Glycerine boils at 153.8°C

If Enthalpy Data is not available then it can be found by Clausius-Clapeyron equation by following:

$\ln P_1^* = -(\Delta Hv/RT_1) + B$…………………………………….. Eq^n (6)

$\ln P_2^* = -(\Delta Hv/RT_2) + B$…………………………………….. Eq^n (7)

Eq^n (4)- Eq^n (5), We get

$-(\Delta Hv/R) = [\{\ln(P_2^*/P_1^*).T_1T_2\}/(T_1-T_2)] = [\{\ln(10/5).427.(440.35)\}/(427-440.35)$

$= 9762.7$ K

$\Delta Hv = (9762.7K).R = (9762.7K).(8.314J/mol\text{-}K) = 81167.11 J/mol$

Thus, The Heat of Vapourisation of Glycerine at 5-10 mm Hg absolute Pressure is 81167.11 J/mol

Similarly, at 20-40 mm Hg Pressure (ΔHv) = 78250J/mol

We Know There is no assurance at the outset that at the stated conditions the system is actually in the two phase region . This should be determined before a flash calculation is attempted. A two phase system at a given temp and with given overall composition can exist over a range of pressure from the bubble point at P_b where V=0 and $\{Z_i=X_i\}$

To Dew point at P_d where V=1 and $\{Z_i=Y_i\}$If the given Pressure lies between the P_b and P_d then the system is indeed made up of two phases at the stated condition.We can find the Bubble Pressure by following formula,

$P_b = P = X_{Glycerine}P^*_{Glycerine} + X_{water}P^*_{water}$

Similarly, For Dew point Calculation, We use following formula,

$P_d = P = [1/\{ Y_{Glycerine}/P^*_{Glycerine} + Y_{water}/P^*_{water}\}]$

Now Our main purpose is to generate the Vapour-Liquid Equilibrium Curve for Glycerine –Water . This can be generated by the following:

Assumption(1):

$Z_{glycerine}$ = 0.60 mole Glycerine/mole ; Z_{water}= 0.4molewater/mole

System Pressure (P) = 5mm Hg ; Temperature(t) = 15°C

By putting The above data in equation (4) and (5) , we get

$P^*_{glycerine}$ = 0.000246mmHg ; P^*_{water} = 12.64mmHg

Then Putting the above obtained values in the following equation, we get

$P_b = P = X_{Glycerine}P^*_{Glycerine} + X_{water}P^*_{water}$ = 5.056mm Hg

Similarly, We calculate Dew point by following

$P_d = P = [1/\{ Y_{Glycerine}/P^*_{Glycerine} + Y_{water}/P^*_{water}\}]$ =……….

So, We always evaluate such temp at which the bubble pressure equals the system pressure so at that pressure, the system will start boiling . Thus, The System Pressure

P_b= 5mmHg lies below P_b=5.056mmHg . Thus, The system starts boiling at T=15◦C

Now, For the flash calculation,We will get by the following formula

$Y_{Glycerine} + Y_{water} = 1$

$\therefore [Z_{glycer}.(P^*_{glycer}/P)/\{1+V\{(P^*_{glycer}/P)-1\}] + [Z_{water}.(P^*_{water}/P)/\{1+V\{(P^*_{water}/P)-1\}] = 1$

$\therefore [0.6.(0.000246/5)/\{1+V\{(0.000246/5)-1\}] + [0.4.(12.64/5)/\{1+V\{(12.64/5)-1\}] = 1$

By trial & error, We get

V = 0.007 mol vapour ;

L + V = 1

$\therefore$L = 1- 0.007 = 0.993 mole liquid

$\therefore Y_{glycerine}=[Z_{glycerine}.(P^*_{glycerine}/P)/\{1+V\{(P^*_{glycerine}/P)-1\}]$

$=[0.6.(0.000246/5)/\{1+0.007\{(0.000246/5)-1\}]$

= 0.0000297 mole Glycerine /mole

$\therefore Y_{water}$ = 0.99997mole water/mole

Now, we will calculate the vapour composition assuming different feed composition but the same system pressure P = 5mm Hg So,

Assumption(2)

$Z_{glycerine}$ = 0.70 mole Glycerine/mole ; Z_{water}= 0.3mole water/mole

System Pressure (P) = 5mm Hg ; Temperature(t) = 20°C

By putting The above data in equation (4) and (5) , we get

$P^*_{glycerine}$ = 0.0004026mmHg ; P^*_{water} = 17.38mmHg

Then Putting the above obtained values in the following equation, we get

$P_b = P = X_{Glycerine}P^*_{Glycerine} + X_{water}P^*_{water}$ = 5.21mm Hg

Similarly, We calculate Dew point by following

$P_d = P = [1/\{ Y_{Glycerine}/P^*_{Glycerine} + Y_{water}/P^*_{water}\}]$ =………

So, The system will start boiling at T= 20°C around . Now, The flash calculation gives

The equilibrium vapour composition ,

$$Y_{Glycerine} + Y_{water} = 1$$

$\therefore [Z_{glycer}.(P^*_{glycer}/P)/\{1+V\{(P^*_{glycer}/P)-1\}] + [Z_{water}.(P^*_{water}/P)/\{1+V\{(P^*_{water}/P)-1\}] = 1$

$\therefore [0.7\times(0.0004026/5)/\{1+V\{(0.0004026/5)-1\}] + [0.3\times(17.38/5)/\{1+V\{(17.38/5)-1\}] = 1$

By trial & error, We get

V = 0.01 ; Σ = 1.017

V = 0.015 ; Σ = 1.0055

V = 0.015 mole ; L = 1- 0.015 = 0.985 mole

$\therefore Y_{glycerine} = [Z_{glycerine}.(P^*_{glycerine}/P)/\{1+V\{(P^*_{glycerine}/P)-1\}]$

$= [0.7\times(0.0004026/5)/\{1+0.015\{(0.0004026/5)-1\}]$

= 0.00005722 mole Glycerine /mole

$\therefore Y_{water}$ = 0.999942mole water/mole

Similarly By this Procedure, We will obtain Glycerine-Water equilibrium curve data at

P = 5 mmHg absolute Pressure

Feed Composition of more Volatile component (X_{water})mole	Vapour Composition of more volatile component (Y_{water})mole
0.56mole	0.99998mole
0.4	0.99997
0.3	0.999942
0.2	0.99988
0.1	0.999617
0.04	0.998
0.02	0.9936
0.01	0.9784

Packed Bed Distillation Column:

Before Initiating the calculation for the Design of Packed Distillation Tower following points should be always considerd:

1) Net free flow cross- sectional area of the support should be 65% (or Larger) of the tower area.
2) As a general rule ,Packing height per support plate should not exceed 12 ft for Raschig rings or 15-20 ft for most other packing shapes .

3) Good Design generally considers that the streams of liquid should enter onto the top of packing on 3 –6 inch square centers for small towers less than 36 inch in diameter, should number $(D/6)^2$ streams for 36 inch and larger , where 'D' is tower inside diameter in inches.
4) Liquid distribution should be after every 6 inch (one stream should be in every 6 inch height).
5) Raschig rings(ceramic) is generally used for the benefit of corrosion resistant to many reactants.
6) The liquid coming down should be redistributed after a bed depth of approx 3 tower diameter for raschig rings, 5-10 tower diameter for saddle packing.
7) As a general rule, Smaller the packing, more efficient contact and high will be pressure drop and vice-versa.

Packing Size(Nominal)	Tower Diameter
2"- 3"	36" or Larger
1"- 1.5"	18" – 24"

For Raschig rings :

Packing size: Tower diameter = 1:20 for good liquid-vapour contact

= 1: 8 for low pressure drop consideration

8) Towers with a 24 inch dia and smaller are most often used with packing rather than trays.
9) Minimum Liquid wetting rate for absorption

L_{min} (ft^3/hr-sq ft cross section)= (MWR)(a_t) = (0.85)(a_t)

Where,

a_t = Packing surface area per unit volume , (ft^2 / ft^3)

MWR = value of minimum wetting rate

10) Towers are usually designed to operate with gas-liquid rates in the loading region

Or within 60-80% of its lower point.

11) Preffered design range for pressure drop is 0.35 to max^m 1” of water/ft.

12) Towers are designed (flood point) usually 40- 60% of gas-liquid rate associated with flooding point.

13) Packing factor is given by

$$F = a / \varepsilon^3$$

Where,

a_t = Sp. Surface of packing (ft^2 / ft^3)

a = Effective inter facial area for contacting (ft^2 / ft^3)

ε = void fraction

Also, Wet packed unshacked Tower:

$$\varepsilon = 1.029\text{-}0.591\varphi \text{ ; Where } \varphi = [\ 1\text{- } (\ d_i/d_o)^2]/\ [1d_o{}^2]^{0.017}$$

Not valid if $\varphi < 0.20$ or for extra thick walls or solids

1 = Ring height (inch)

d_o = outside diameter of ring (inch)

d_i = inside diameter of ring (inch)

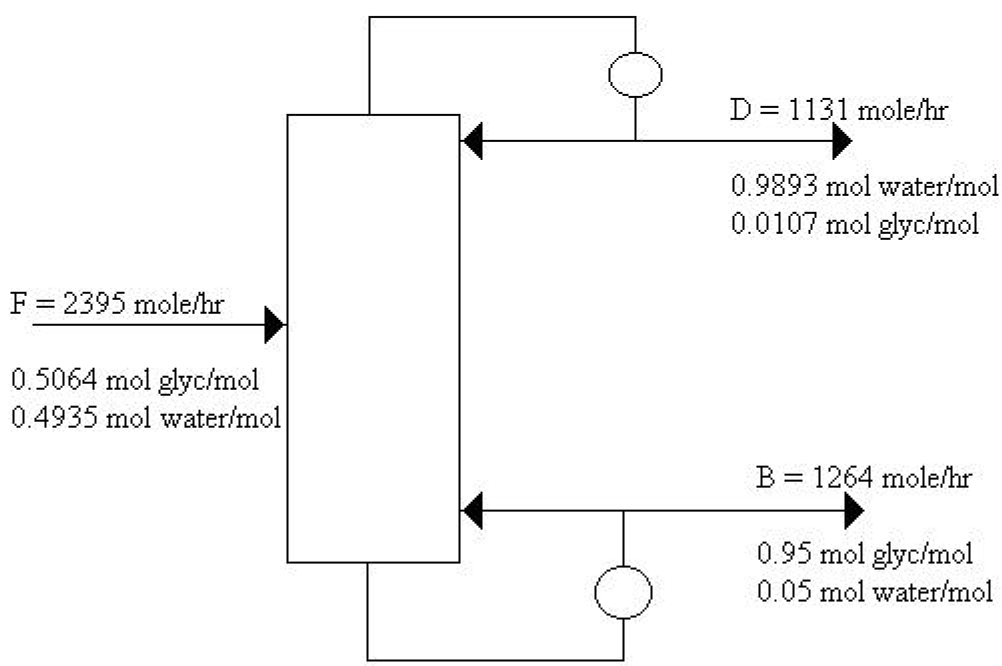
14) For design of commercial towers, value of a/ε^3 should be increased from 15 to 75% for ceramic materials.

Design Data:

System pressure = 5 mmHg absolute pressure

Temp at the Rectifying section (assumed) = 10°C

Temp at the Stripping section (assumed) = 154°C



Basis: 133 Kg/hr of crude Glycerine (84wt%) entering the Distillation column

i.e, 100 Kg Feed contains 84 Kg Glycerine(Mol Wt = 92.09) and 16 Kg Water

(Mol Wt = 18)

i.e, 0.91215 Kmol Glycerine and 0.888Kmol Water = 1.80 Kmol

Thus, mole Glycerine in feed = 0.91215Kmol/ 1.80Kmol feed

= 0.5064 Kmol/Kmol feed = 0.5064mol/mol feed

Similarly, Water in feed = 0.4935 Kmol/Kmol feed = 0.4935mol/mol feed

Now,

Feed(F) = 133 Kg/hr of crude Glycerine (84wt%)

= 133$\times$ (0.84Kg)$\times$(1Kmol Glycerine/92.09 Kg Glycerine) + 133$\times$ (0.16Kg)$\times$(1Kmol Water/18 Kg Water)

= 1213.16 mole glycerine + 1182.2mole Water

= 2395 mole/hr = 0.6652mole/sec

As supplied data, bottom product (B) composition of Glycerine is 0.95moleGlycerine/mole of bottom product.and 99% of the glycerine fed is obtained in the bottom product stream B. So,

$\therefore$2395$\times$ 0.5064mole$\times$(0.99) = B $\times$ (0.95mole Glycerine/mole)

$\therefore$B = 1264 mole/hr = 0.3511mole/sec

{0.95mole Glycerine/mole and 0.05 mole water/mole)

Overall material Balance,

$F = D + B$

$D = 2395 – 1264 = 1131$mole/hr = 0.3141mole/sec

Overall Water Balance,

2395×0.4935mole Water = 1131mole$\times$ X_{water} $+1264$mole$\times 0.05$mole Water

$\therefore$ $X_{water} = 0.9893$mole Water/ mole

$\therefore$ $X_{glycerine} = 0.01073$mole Glycerine/mole

Reflux Ratio: $R_D = L_n/ D = 2$ (Assume)

The internal flow raes are calculated from a series of mass balances

In Rectifying Section:

$V_n = L_n + D$ ………………………………………..$Eq^n(1)$

In Striping Section:

$V_m = L_m – B$ ………………………………………. $Eq^n(2)$

Where,

L_n, L_m = Liquid flows in the rectifying & Stripping sections respectively(mol/hr)

V_n, V_m = Vapour flows in the rectifying & Stripping sections respectively(mol/hr)

F = Feed rate , D = Top Product , B = Bottom Product

$\therefore L_n = R_D \times D = 2\times 1131$mol/hr = 2262mol/hr = 06283mole/sec

$\therefore V_n = L_n + D = 2262 + 1131 = 3393$mol/hr = 0.9425 mol/sec

$\therefore L_m = L_n + F = 2262 + 2395 = 4567$mol/hr = 1.2936 mol/sec

B = 1264 mol/hr = 0.3511mol/sec

$\therefore V_m = L_m - B = 4567 - 1264 = 3393$ mol/hr = 0.9425 mol/sec

The slope of the feed line is given by(assuming Feed at its bubble point)

$\therefore$ Slope = [q/(q-1)] = [1/(1-1)] = ∞

So, m = Tan∞ = 90°

Thus, at Z_f ,the slope will be $\perp$ to feed.

Equation of the operating line in Enriching Section:

$V. Y_{n+1} = L.X_n + D. X_D$

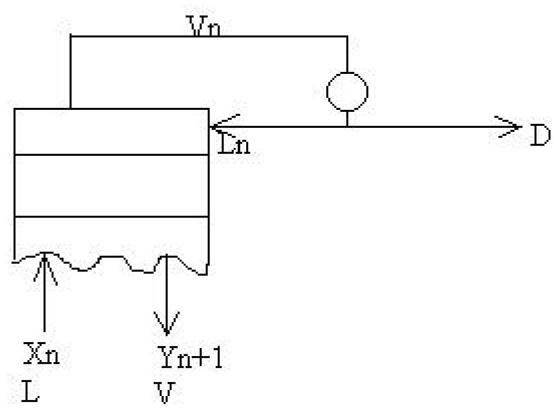
$\therefore Y_{n+1}= (L/V)X_n + (D/V)X_D$

$\therefore Y_{n+1}= [R_D/(R_D +1)]X_n + [X_D/(R_D +1)]$……………………$Eq^n(2)$

This is the equation of straight line X,Y co-ordinate

Slope = $L/V = [R_D/(R_D +1)]$

Y- intercept = $[X_D/(R_D +1)]$



Equation of the operating line in strippin Section:

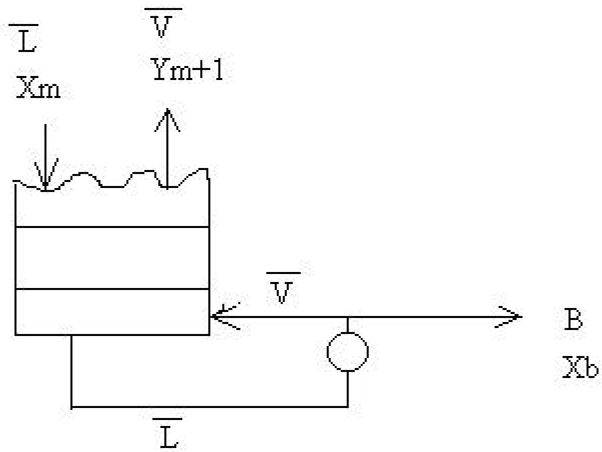
$\bar{L}.X_m = \bar{V}.Y_{m+1} + B.X_B$

$\therefore Y_{m+1} = (\bar{L}/\bar{V}) X_m – (B/\bar{V})X_B$

$\therefore Y_{m+1} = [\bar{L}/(\bar{L}\text{-}B)]X_m –[B/(\bar{L}\text{-}B)]X_B$

This is the equation of straight line, So

Slope = $\bar{L}/\bar{V} = [\bar{L}/(\bar{L}\text{-}B)]$…………………………………**$Eq^n$(3)**



Putting the value in operating line of enriching section3

Slope = L/V = $[R_D/(R_D+1)] = [2/(2+1)] = 0.666$

$\therefore m = \tan\theta = 0.666$

$\therefore \theta = 33.68^\circ$

$\therefore$ Y- intercept = $[X_D/(R_D+1)] = [0.9893/(2+1)] = 0.33$

Now , We plot on the equilibrium curve of Glycerine-Water for Distillation in packed column,

Now, The no. of overall transfer unit N_{OG} must be obtained by graphical integration,

$$N_{OG} = \int_{y1}^{y2} [dy/(y^*-y)]$$

From the graph , we get the following data

x	**y***	**y**	**y*-y**	**[1/(y*-y)]**
0.05	0.99	0.05	0.94	1.0638
0.10	0.999617	0.11	0.889617	1.1240
0.20	0.9988	0.25	0.74988	1.3335
0.30	0.999942	0.39	0.609942	1.6395
0.40	0.99997	0.53	0.46997	2.127795
0.49	0.999980	0.66	0.33998	2.9413
0.60	0.999990	0.73	0.26999	3.7038
0.70	0.999994	0.80	0.199994	5.0
0.80	0.999996	0.865	0.134996	7.4076
0.90	0.999998	0.93	0.069998	14.286
0.99	0.999999	0.9893	0.010699	93.466

The area under the curve in Stripping section from y = 0.05 to y = 0.66 is given by,

Area = N_{OG} (S) = $(0.10-0.05)\times(1.0638+1.124)/2 + (0.25-0.10)\times(1.3335+1.1240)/2 +$

$(0.39-0.25)\times(1.6395+1.3335)/2 + (0.53-0.39)\times(2.127795+1.6395)/2 +$

$(0.66-0.53)\times(2.9413+2.127795)/2$

$= 1.0403$

The area under the curve in Rectifying section from y = 0.66 to y = 0.9893 is given by, Area = N_{OG} (R) = (0.73-0.66)×(3.7038+2.9413)/2 + (0.80-0.73)×(5.0+3.7038)/2 +

(0.865-0.80)×(7.4076+5.0)/2 + (0.93-0.865)×(14.286+7.4076)/2 +

(0.9893-0.93)×(93.4666+14.286)/2

= 4.84

Total No of transfer unit = N_{OG} (R) + N_{OG} (S)

= 1.0403 + 4.84

= 5.8803

For Height Of Transfer Unit :-

The height of a transfer unit for distillation is not available, Thus, By thumb rule These are considered:(Ludwig , vol-2,page-210)

a) Never use HETP less than 12 inch if the Tower dia is 12 inch or larger. For general assumption , Use HETP = 1.5 to 2.0 ft
b) Use HETP = H_{OG} or H_{OL} if other data are not available.
c) Use HETP = Column diameter (over 12 inch dia) if no other information available up to 48” diameter.

Also,In vaccum distillation, The pressure drop should be low between 0.10-0.25 inch Water/ft of packing .

Now,

MW at the top of column = 0.9893×18 + 0.0107×92.09 = 18.79

MW at the bottom of column = 0.05×18 + 0.95×92.09 = 88.3

Rectifying Section:

Liquid Flow rate (L)= 2262mol/hr $\times$ 18.79gm/mole = 42.50Kg/hr = 93.7lb/hr

Vapour Flow rate (V)= 3393mol/hr $\times$ 18.79gm/mole = 63.75Kg/hr = 140.55lb/hr

= 0.03904 lb/sec

Stripping Section:

Liquid Flow rate (L)= 4657mol/hr $\times$ 88.3gm/mole = 411.21Kg/hr = 906.56lb/hr

Vapour Flow rate (V)= 3393mol/hr $\times$ 88.3gm/mole = 299.6Kg/hr = 660.5lb/hr

= 0.1834 lb/sec

1 mole Basis:

0.9893 mol water(MW= 18) = 17.80gm

0.0107 mol Glycerine (MW = 92.09) = 0.985 gm

$\therefore$ Water – 0.9475 gm water/gm

$\therefore$ Glycerine-0.05245 gm glycerine/gm

Here, For the calculation of density of vapour , We assume ideal situation

T_1 = 154°C = 427K ; P = 5mm Hg = 666.61Pa

T_2 = 10°C = 283K

$\therefore \rho_1 = PM_1/RT_1 = [(666.61\times18)/(8.314\times427)]\ mol/M^3 \times 18gm/1mol$

$= 60.838\ gm/M^3$

$\therefore \rho_2 = PM_2/RT_2 = [666.61\times92.09/8.314\times427]\ mol/M^3 \times 92.09gm/1mol$

$= 1589.68\ gm/M^3$

$\therefore 1/\rho = m_1/\rho_1 + m_2/\rho_2 = 0.9475/60.838 + 0.05245/1589.65 = 64.07gm/M^3$

In Stripping Section

$\therefore \rho_{vs} = \rho_G =$ 64.07gm/M^3 = 0.06407Kg/M^3 =0.004lb/ft^3 at Temp - 154°C

In Rectifying Section

$\therefore$ Vapour density (ρ_{VR}) = 0.3076Kg/M^3 = 0.0192lb/ft^3 at Temp-10°C

Similarly for Liquid,

$\therefore 1/\rho_{Lbottom} = m_1/\rho_1 + m_2/\rho_2$ = 0.9475/1000 +0.05245/1260

$\therefore \rho_{Lbottom}$ = 1011Kg/M^3

$\therefore 1/\rho_{Ltop} = m_1/\rho_1 + m_2/\rho_2$ = 0.9893/1260 +0.01018/1000

$\therefore \rho_{Ltop}$ = 1257.32Kg/M^3

$\therefore$ Avg Density of Liquid (ρ_{Lavg}) = (ρ_{Ltop} + $\rho_{Lbottom}$)/2 = (1011 + 1257.32)/2

= 1134 Kg/M^3 = 70.8055lb/ft^3

Avg Liquid Viscosity(μ_L) = 1cp = 0.001N-S/M^2 = 0.001Kg-S/M =6.72$\times 10^{-4}$ lb/ft-S

= 2.42 lb/ft-hr

Now , For Diameter of Stripping Section:

1) calculate

   $(L/G)\times\sqrt{(\rho_G/\rho_L)}$ = (411.21/299.6)$\times\sqrt{(0.06407/1130)}$ = 0.0103

2) Select the pressure drop . Since, The Distillation column is working under high vaccum condition so By Thumb's rule,The pressure drop should be between 0.10-0.25 inch H_2O /ft of packing .

   So, We select pressure drop = 0.10 inch H_2O /ft of packing.

3) We Select packing (Nominal) size = 2" Raschig Ring

   Packing Factor (F) = 65

4) Now From fig 9.13B(page-159,Ludwig,V-2)

At ordinate = 0.0103 and Pressure drop of 0.10 inch H_2O /ft of packing,We get

$(G^2 F\mu^{0.10})/(\rho_G(\rho_L-\rho_G)g_c = 0.018$

$\therefore G^2 = [\{0.018\times4.0\times10^{-3} \times (70.8055-4\times10^{-3})\times32.2\}/\{65\times(1)^{0.1}\}]$

$\therefore G = 0.050257$ lb/ft^2-S

5) Diameter of Packed Tower (D_{Bottom}) = 1.1283√{Gas Rate, lb/Sec(G")}/G

= 1.1283√{0.1834/0.050257}

= 4.1177 ft

Similarly, For Diameter of Rectifying Section:

1) calculate

$(L/G)\times \sqrt{(\rho_G/\rho_L)} = (42.50/63.75)\times\sqrt{(0.3076/1130)} = 0.011$

2) Select the pressure drop . Similarly as above,

We select pressure drop = 0.10 inch H_2O /ft of packing.

3) We Selected packing (Nominal) size = 2" Raschig Ring

Packing Factor (F) = 65

4) Now From fig 9.13B(page-159,Ludwig,V-2)

At ordinate = 0.011 and Pressure drop of 0.10 inch H_2O /ft of packing,We get

$\therefore(G^2 F\mu^{0.10})/(\rho_G(\rho_L-\rho_G)g_c = 0.0178$

$\therefore G^2 = [\{0.0178\times0.0192 \times (70.8055-0.0192)\times32.2\}/\{65\times(1)^{0.1}\}]$

$\therefore G = 0.1094$ lb/ft^2-S

5) Diameter of Packed Tower (D_{Bottom}) = 1.1283√{Gas Rate, lb/Sec(G”)}/G

$= 1.1283\sqrt{0.03904/0.1094}$

$= 0.674$ ft

Here, Since The Diameter of the Stripping Section is the rate Limiting Diameter .So we have to choose the diameter of the Stripping section .

∴Diameter of the Packed Tower (D) = 4.11ft ≈ 4.0 ft

And The Packing material is Raschig ring of Nominal Packing size = 2”

Which is also consistent with rule of Thumb, Which states that the Nominal Packing size should be 2”-3” for the column diameter of 36 inch or larger.

Also,

The max^m pressure drop / ft of Packing = 0.10 inch H_2O /ft of packing

Thus For Height of Transfer Unit:

∴HETP = Column Diameter (over 12 inch Diameter up to 48” diameter)

∴HETP = D = 4 ft

∴ If Height of Transfer unit is not given then

H_{OG} or H_{OL} = HETP

∴H_{OG} = 4 ft

∴ Packing Height (Z) = $H_{OG} \times N_{OG}$ = 4ft × 5.8803 = 23.5 ft

∴Total pressure drop (ΔP) through the Packing height

ΔP = 23.5 ft × 0.10 inch H_2O /ft of packing

= 2.35 inch H_2O

As The Packing Height = 23.5 ft

Trans Disengagement Height should be at least 10% of the Packing Height , But We kept it 15% So,

Total Tower Height = 23.5 + 0.15 $\times$ 23.5 = 27 ft

Also, For Better mass transfer and the channeling should not occur, Packing height per support plate should not exceed 12 ft for Raschig rings or 15-20 ft for most other packing shapes .So

Seperator is used every 12 ft of Packing .

<u>Now,Calculation for Max^{m} allowable velocity(In Rectifying Section):</u>

At Pressure = 1 Atm ; T = 10°C=283K

ρ_G = PM/RT = [{1.01325$\times10^{5}$ $\times$18.79}/{8.314$\times$283}] mol/M^3 $\times$ 18.79gm/mol

= 15204.55gm/M^3 = 15.20 Kg/ M^3 = 0.9489 lb/ft^3

Thus,

$(L/G)\times\sqrt{(\rho_G/\rho_L)}$ = (42.50/63.75)$\times\sqrt{(15.20/1130)}$ = 0.0773

From Fig16.20 (page-698,Peter Timmerhaus, Plant Design)

At Abscissa = 0.0773 , The Flooding ordinate with random packing is given by,

$\therefore [V_m^2\,(a_p/\varepsilon^3)\rho_G\,(\mu'_L)^{0.2}\,]/\,(g\times\rho_L) = 0.13$

Thus, The max^{m} allowable velocity of vapour flow through Packed Tower Distillation column at 1 atm is given by,

$\therefore [V_m^2\ (65)\times0.9489\text{lb/ft}^3\ \times(6.72\times10^{-4}\text{lb/ft-S})^{0.2}\ ]/\ (32.2\ \text{ft/S}^2\times70.8055\text{lb/ft}^3) = 0.13$

$\therefore V_m = 4.55$ ft/S

This was the max allowable vapour velocity at 1 atm and 273K

Now, The Max^m allowable vapour velocity at which flooding will occur at 5mm Hg (0.00657atm) is given by,

$V_{m1} = V_m \times (P_1/P)^{0..5} \times (T/T_1) = 4.55 \times (1/0.00657)^{0.5} \times (273/283) = 54.1466$ ft/S

Thus, The max^m allowable vapour velocity at which flooding will occur is 56.13 ft/S

But for the operation , It is taken as 50-70% of the maximum allowable vapour

Velocity

$\therefore V = 0.5\ V_{m1} = 0.5 \times 54.1466 = 27.073$ ft/S (T=283 K ; P = 5mm Hg)

Here, We have neglected the change in density of liquid due to slight temp

difference of 10°C .

<u>Calculation for Max^m allowable velocity(In Stripping Section):</u>

At Pressure = 1 Atm ; T = 154° C = 427K

$\rho_G = PM/RT = [\ \{1.01325 \times 10^5 \times 88.3\}/\{8.314 \times 427\}]\ mol/M^3 \times 88.3 gm/mol$

$= 222530\ gm/M^3 = 222.53\ Kg/M^3 = 13.98\ lb/ft^3$

Thus,

$(L/G) \times \sqrt{(\rho_G/\rho_L)} = (411.21/299.6) \times \sqrt{(13.98/1130)} = 0.15266$

From Fig16.20 (page-698,Peter Timmerhaus, Plant Design)

At Abscissa = 0.15266 , The Flooding ordinate with random packing is given by,

$\therefore [V_m^2 (a_p/\varepsilon^3)\rho_G (\mu'_L)^{0.2}]/(g\times\rho_L) = 0.10$

Thus, The max^m allowable velocity of vapour flow through Packed Tower Distillation column at 1 atm and 273 K is given by,

$\therefore [V_m^2 \quad (65)\times 13.98 \quad lb/ft^3 \quad \times(6.72\times10^{-4} lb/ft\text{-}S)^{0.2} \quad]/ \quad (32.2\ ft/S^2 \times 70.8055 lb/ft^3) = 0.10$

$\therefore V_m = 1.04$ ft/S

This was the max allowable vapour velocity at 1 atm and 273K

Now, The Max^m allowable vapour velocity at which flooding will occur at 5mm Hg (0.00657atm) is given by,

$V_{m1} = V_m \times (P_1/P)^{0..5} \times (T/T_1) = 1.04 \times (1/0.00657)^{0.5} \times (273/427) = 8.20$ ft/S

Thus, The max^m allowable vapour velocity at which flooding will occur is 8.20 ft/S

But for the operation , It is taken as 50-70% of the maximum allowable vapour

Velocity

$\therefore V = 0.5\ V_{m1} = 0.5\times 8.20 = 4.1$ ft/S (T = 427 K ; P= 5mm Hg)

Here, We have neglected the change in density of liquid due to Temp change.