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DESIGN ON A BUBBLE-CAP COLUMN TO REPLACE NITRIC ACID CONCENTRATOR

By
H. Wallingford

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DESIGN OF A BUBBLE-CAP COLUMN TO REPLACE NITRIC ACID CONCENTRATOR

By H. Wallingford

Design Conditions:

Feed:* 16,500 gal/day; 7.0% by wt. of HNO_3

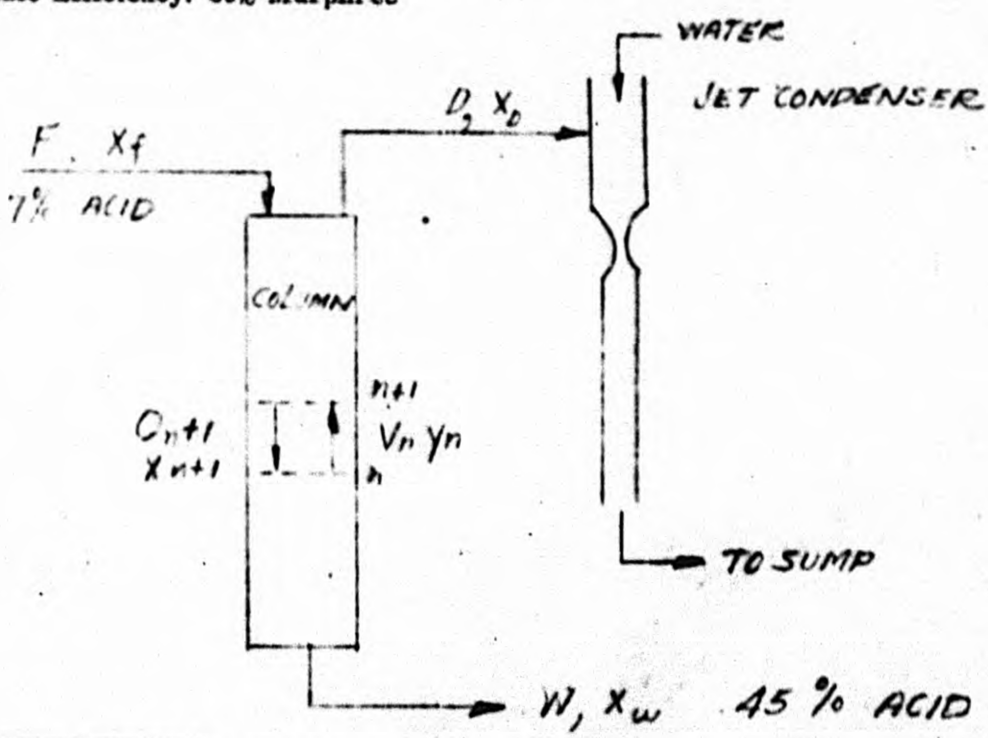
enters at 140°F. / 60°C. .80% solids ($Ca(NO_3)_2$)

Product: 45% by wt. HNO_3

Overall Recovery: 95%

Condenser Pressure: 22" Hg Vacuum (200 mm Hg)

Plate Efficiency: 50% Murphree



* This feed rate is 18% higher than the feed rate during the peak production period, January 1947 – June 1948.

Two types of design are possible:

1. A stripping column, in which the feed enters on the top plate.
2. A rectifying column, with feed entered at the middle and water used for reflux.

Design (1) is selected and presented in these calculations because:

a. Design (1) requires a smaller capacity reboiler.

b. Design (1) requires fewer plates*2 (Note: In Design (2) a higher purity overhead must be obtained in order to boil off the extra water used as reflux and still maintain 95% yield).

$$\text{Feed temperature} = (140 - 32)\frac{5}{9} = 60°C$$

At 60°C, sp.gr. of 7% solution is 1.0188*1

$$F_{lbs.} = 16{,}500\,\frac{gal}{day} \times 8.33\,\frac{lb}{gal} \times 1.0188 = 140{,}029\,\frac{lbs.}{day}$$

lbs. 100% acid in feed = .07 × 140,029 = 9,802#

9,802 × (0.95) = 9312 lbs. acid recovered as 45% solution.

9,312# = .45 $W_{lbs.}$

20,693# =	$W_{lbs.}$ total
9,312#	acid in W
11,381#	water in W

140,029 – 9,802 = 130,227# water in feed

130,227#	water in feed
11,381#	water in waste
118,846#	water in overhead

9,802 × .05 = 490.1# acid in overhead

Compositions:

Feed:

$$\frac{9802}{63.02} = 155.54 \text{ moles acid}$$

$$\frac{130{,}227}{18.02} = \frac{7{,}226.8}{7{,}382.34}\ \begin{matrix}\text{moles water}\\ \text{total}\end{matrix}$$

$$X_F = \frac{7{,}226.8}{7{,}382.34} = .9789$$

Waste:

$$\frac{9312}{63.02} = 147.76 \text{ moles acid}$$

*1Perry's Handbook, p419.

*2See report, July 20, 1949.

$$\frac{11{,}381}{18.02} = \frac{631.58 \text{ moles water}}{779.34 \text{ total}}$$

$$X_W = \frac{631.58}{779.34} = .8104$$

Distillate:

$$\frac{490.1}{63.02} = 7.777 \text{ moles acid}$$

$$\frac{118{,}846}{18.02} = \frac{6{,}595.23 \text{ moles water}}{6603.007 \text{ total}}$$

$$X_D = \frac{6{,}595.23}{6603.01} = .9988$$

Operating Lines:

$$F = W + D$$

$$F x_f = W x_w + D x_d$$

$$F x_f + VnYn = On + 1 \quad Xn + 1 + D x_d$$

$$Yn = \frac{(On + 1)}{Vn} Xn + 1 + \frac{D x_d - FX_f}{Vn}$$

$$On + 1 = F + D$$

where D = amount of vapor that must condense to heat feed to Bpt.

$\Delta H_{vaporization}$ = 18,200 Btu/# mole

Sp.ht. of feed = .927 Btu/°F.lb.*[1]

B pt. of feed = 67.25°C (Perry p401)

Heat required to heat feed = 1.8 (67.25 – 60.0)(.927)(140,029) = 1,692,000 Btu/day.

$$\text{Therefore} = \frac{1{,}692{,}000}{18{,}200} = 93.0 \frac{\#\text{moles}}{\text{day}}$$

$$On + 1 = 7{,}382.34 + 93.0 = 7475.34$$

$$F + Vn = On + 1 + D$$

$$7{,}382.3 + Vn = 7475.3 + 6603.0$$

$$Vn = 6{,}696.0 \text{ moles}$$

$$Yn = \frac{(On + 1)}{Vn} Xn + 1 + \frac{DX_d - FX_f}{Vn}$$

*[1] Landolt Bornstein Physikilisch–Chemische Tabellen 5. Anflaze p1657.

$$Yn = \frac{(On + 1)}{Vn} Xn + 1 - \frac{W x_w}{Vn}$$

$$Yn = \frac{(7475.34)}{6,696.0} Xn + 1 - \frac{631.58}{6,696.0}$$

$$Yn = (1.1164)\ Xn + 1 - .09432$$

When Xn + 1 = Yn
Xn + 1 = Yn = .8104 — Point 1
When Xn + 1 = .95
Yn = .9663

From the McCabe Miele diagram*

8.5 50% eff. steps are required.

Since the kettle or reboiler is likely to have an efficiency approaching 100% (because of the good vapor-liquid content in the boiling liquid several feet in depth) the kettle will probably do the enrichment equivalent to the first two plates on the diagram. Hence ten trays in addition to a kettle will be used. This design provides enough safety factor to allow for such uncertainties as the effect of the dissolved calcium nitrate on the vapor equilibrium relationships. At low water concentrations (80–85 mol %) even a large change in the position of the x-y curve would change the number of plates required very little. At high water concentrations where the pinched region of the McCabe construction occurs, however, the effect of dissolved solids might be serious. It is recommended that the effect of dissolved calcium nitrate on the x-y curve be calculated or determined experimentally.

TRAY DESIGN

Important Design Restrictions

1. Liquid distribution across the plate must be good. This is especially important in vacuum columns with low liquid rates. (Perry's Handbook, p1454).
2. The pressure drop of vapor passing through the tray must not exceed 1.5–2.0" water. This pressure drop is fixed by the unfavorable effect of pressure on the x-y curve. See x-y curves drawn for several pressures slightly above the condenser pressure used in this design (22"Hg).
3. The vapor velocity must be at least 25 ft/sec through the cap slots so that reasonable bubble, spray and foam formation will occur above the tray. This spray or bubbles is essential to high fractionating efficiency (Perry p1441).
4. The vapor velocity must not be high enough to cause entrainment. The familiar equation: $u = Kv\ (\rho 1 - \rho 2)/\rho 2$ gives a limiting vapor velocity, above which

* Note: Upper Operating line is horizontal at $y = X_D$.

entrainment becomes important. (Perry p1449). The superficial vapor velocity used in this design must therefore be less than the μ given by this equation.

A low liquid seal of $\frac{1}{2}$" will be used to minimize pressure drop due to liquid head.

The Perry equation for the maximum vapor velocity at which entrainment is negligible is:

$$u_{max} = K_v \sqrt{\frac{\rho_1 - \rho_2}{\rho_2}} \qquad \text{(Perry, p1450)}$$

where ρ_1 = liquid density
ρ_2 = vapor density
u_{max} = max. permissible vapor velocity, based on column cross section, ft/sec.
K_v = constant (a function of liquid seal and tray spacing).

Bottom of Column

$$\frac{ft^3}{mol} = 359 \times \frac{29.9}{7.9}\ \frac{(460 + 176)}{492} = 1760\ \frac{ft^3}{mol}$$

$$1\ ft^3 \text{ vapor contains } \frac{1}{1760} = \begin{array}{ll} 5.69 \times 10^{-4} & \text{moles total} \\ \underline{.96} & \text{Kettle vapor composition} \\ 5.44 \times 10^{-4} & \text{moles water} \\ .25 \times 10^{-4} & \text{moles } HNO_3 \end{array}$$

$$\rho_2 = (25 \times 10^{-4})(63)_{mol.wt.} + (5.44 \times 10^{-4})(18)_{mol.wt.}$$

$$\rho_2 = .0114\ lbs./ft^3$$

$$\rho_1 = 62.4 \times 1.2119^{*} = 75.5\ lb/ft^3$$

$$\frac{75.5 - .0114}{.0114} = 81.5$$

At Top of Column

$$\frac{ft^3}{mole} = 359 \times \frac{29.9}{7.9}\ \frac{(460 + 152)}{492} = 1690\ \frac{ft^3}{mole}$$

Vapor at top plate is 99.88 mol % water

$$1\ ft^3 \text{ vapor contains } \frac{1}{1690} = 5.92 \times 10^{-4} \text{ moles}$$

$$\rho_2 = (.9988)(5.92 \times 10^{-4})\ 18 + (.0012)(5.92 \times 10^{-4})\ 63$$

$$\rho_2 = .0107$$

$$\rho_1 = 62.4 \times 1.029^{\dagger} = 64.1$$

* Perry, p419.
† Perry, p432.

$$\frac{64.1 - .0107}{.0107} = 77.5$$

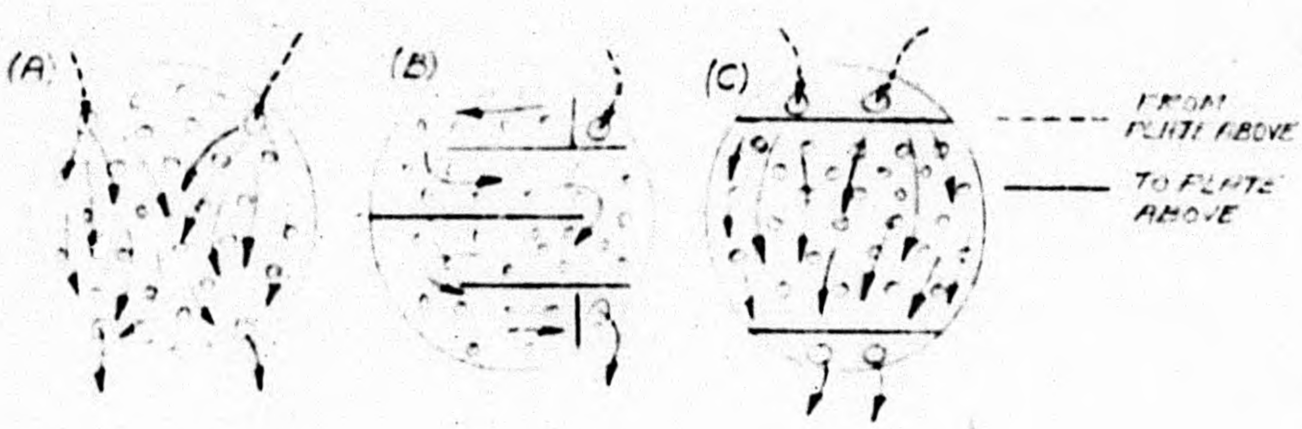
The value of K_v depends on the tray spacing and liquid seal. K_v increases with tray spacing but increases less rapidly above spacings of 18" to 24". The maximum permissible vapor rate is calculated below using the minimum value of $\rho_1 - \rho_2/\rho_2$ in the column and tray spacings of 18" and 24":

Tray spacing in.	K_v	u max. permissible ft/sec. based on tower area
18	0.15	11.6
24	0.185	14.3

RESTRICTIONS IMPOSED BY PRESSURE DROPS

Liquid Gradient Considerations

Attention is directed toward good liquid distribution and low liquid gradient. Several possible designs are:



Design (A) has spaces where liquid inculation is poor and caps are inefficient. Design (B) provides excellent liquid distribution but has the disadvantages of high liquid gradient across the tray (and resultant low cap efficiency) and expensive construction. Design (C) is chosen because it is known to give efficient liquid distribution in large diameter columns (Robinson & Gilliland "Elements of Fractional Distillation," p217) and offers relatively inexpensive construction, and easy access for cleaning.

A number of tray diameters and cap sizes were drawn on scale drawings, the number of caps that could be employed on a tray determined (using 1" to $1\frac{1}{2}$" cap spacing), and the corresponding pressure drops determined from these equations:

$\Delta\rho$ in Riser: $\Delta\rho_1 = .0116 \dfrac{Q^2}{D^4} R$

$\Delta\rho$ in Slots:* $\Delta\rho_2 = K\left(\dfrac{Q}{L}\right)^{0.4} R$ (Equation recommended by Robinson & Gilliland, p214)

$\Delta\rho_3 = Cu^2 \dfrac{\rho_2}{\rho_1}$ (Equation recommended by Perry, p1455)

where Q = vapor flow rate, ft^3/min.per cap
D = riser diameter, in.
R = ratio of density of vapor to that of air
K = constant (function of cap and slot size)
L = total slot width per cap, in.
C = orifice coefficient, 0.51
u = linear vapor velocity in slots, ft/sec.
ρ_2, ρ_1 = densities of vapor and liquid, respent.

$$R = \frac{\text{Mol wt.Vapor}}{\text{Mol wt.Air}} = \frac{20}{27} = .70$$

A conservative value of .73 will be assumed for R. For the calculation of ρ_3, a vapor velocity of 30 ft/sec in the slots is assumed. Therefore, for all plate diameters or cap sizes:

$\Delta\rho_3 = (.51)(30)^2\left(\dfrac{.0107}{64.1}\right) = .93$" water per tray.

The calculated data are:

Tray Diameter 4' 8"

Cap Diameter in.	No. Caps per tray	Riser Diam., in.	ρ_1 in H_2O	ρ_2 in H_2O	Total ρ $\rho_1 + \rho_3$ in. H_2O
7"	29	5"	1.10	.80	2.03
6"	37	4.5"	1.31	.77	2.24
4	86	3"	1.39	.64	2.32

* The pressure drop in the slots was calculated by two different empirical equations.

Tray Diameter 6' 0"

7"	49	5	.384	.645	1.31
6"	64	4.5	.344	.690	1.27
4"	130	3.0	.389	.540	1.32

Sample Calculations

Tray Diag. 4'8", Cap size 7"
From scale drawing, no. caps = 29.

If the annular area between riser and cap is to equal the riser area,

$$(\text{Dia riser})^2 \frac{11}{4} = \frac{1}{2}\frac{11}{4}\, 7^2$$

Dia riser = 5"

whence

$$\Delta\rho_1 = (.0118)\left(\frac{Q}{54}\right)^2 (.73)$$

$$Q = \frac{\dfrac{6696}{24}\,\dfrac{\text{mole}}{\text{day}} \times 1760\,\dfrac{\text{ft}^3}{\text{lb-mole}}}{\dfrac{\text{hr.}}{\text{day}} \times 60\,\dfrac{\text{min}}{\text{hr.}} \times 29 \text{ caps}} = 282\,\frac{\text{ft}^3}{\text{min-cap}}$$

$$\rho_1 = (.0118)\,\frac{(282)^2}{54}\,(.73) = 1.10'' \text{ water}$$

$$\rho_2 = (.35)(.73)\,\frac{(282)^{0.4}}{L}$$

L is taken as 75% of the perimeter of the cap.

L = .75 11(7) = 16.5"

$$\rho_2 = (.35)(.73)\,\frac{(282)^{0.4}}{16.5} = .80'' \text{ water}$$

The total pressure drop is taken as $\Delta\rho_1 + \Delta\rho_3$ since $\Delta\rho_3$ is always greater than $\Delta\rho_2$ and a conservative design is desired.

The total pressure drop in the table (which does not include ρ due to liquid seal) does not change with cap size on a given size tray but does decrease as the tray diameter increases. If the 6 ft. dia. tray is selected the total pressure drop of vapor including liquid seal will be:

$\Delta\rho$ total = 1.3 + .5 = 1.8" H_29

The 7'' caps will be used because the low number of caps required will facilitate cleaning. The superficial vapor velocity is then:

$$u = \frac{6696}{24} \times \frac{1760\ \frac{ft^3}{mol}}{60 \times (6)^2\ \frac{11}{4} \times 60} = 4.84\ ft/sec.$$

This linear velocity is less than u max. predicted by Perry's equation (table p10) and hence entrainment is not likely to be serious.

If the slot velocity is 30 (ft/sec),

$$\text{slot area/tray} = \frac{\frac{6696 \times 1760}{24 \times 3600}}{30} = 4.55\ ft^2/tray$$

$4.55 \times 144 = 655\ in^2$

$$\text{Slot depth} = \frac{655}{49\ \text{caps} \times \frac{16.5}{L}} = .81''$$

portion of cap perimeter cut away for slots

Slots will be rectangular, $^{13}/_{16}$'' deep, and will have a total width of 16.5 inches per cap.

Position of Caps

Caps may be anchored on the tray so that (1) the teeth between slots touch the tray or (2) a space exists between the teeth and the tray. Since this column may be used at a reduced production rate (and hence reduced vapor rate) it is desirable to maintain a reasonable slot velocity by forcing the vapor to pass through the slots. Thus vapor is not permitted to pass under the edge of the cap without being properly dispersed into bubbles. Hence design (1) is chosen to give better performance over a wider range of production rates.

Tray Spacing

The vapor velocity is limited by pressure drop and not by entrainment. Hence the 18'' tray spacing or even 12'' tray spacing would give good performance. However the necessity of occasional cleaning dictates a plate spacing of 24'', so that 18'' manholes can be installed between each tray.

Column Height = 10 trays × 24'' + 2' = 26.
where 2' represent the liquid holdup at bottom of column.

Weir Design

The chord weirs on the feed and discharge sides of the trays will be about ⅔ of the column diameter (72'') or 48''. The notch spacing required to provide good liquid distribution on the tray is estimated to be $1\frac{1}{2}''$. There will be 32 notches in the weir.

Weir Notches

The following equation (Perry p863) is used to select a notch angle and depth for efficient operation.

$$A\,\frac{g\rho}{uH} = 0.575\left(\tan\frac{\alpha}{2}\right)0.996\left(\frac{H^3 g_1 \rho^2}{\mu^2}\right)^{0.49}$$

where ρ = density, lbs./ft^3
u = viscosity, lbs./ft.sec
αd = included angle of notch
H = head above vertex of notch, ft.
g_1 = gravitational constant, 32.2 ft/sec.^{2}
g = flow rate, ft^3/sec. per notch

This equation is to be applied to the notches in the overflow weirs on the discharge side of each tray. Since the liquid flow rate in ft^3/sec. varies from tray to tray, the design calculations are carried out both for the top and the bottom trays. Notches are $1\frac{1}{2}''$ apart, 32 notches per weir.

Top of Column

Solution is 7% HNO_3 by wt. 97.57 mol % water

$$u = .473^{*}\text{ centipoises} = (.000672)(.473) = .000325\ \frac{\text{lbs}}{\text{ft-sec.}}$$

$$= 1.015^{\dagger} \times 62.4 = 63.4\ \text{lbs/ft}^3$$

$$g = \frac{\text{(total weight of water and acid in feed)}}{63.4}$$

$$g = \frac{(7382 \times .9757 \times 18_{\text{mol.wt.}}) + (7382)(1 - .9757)(63\text{ mol wt.})}{63.4\ \frac{\text{lbs.}}{\text{cu.ft.}} \times 24\ \frac{\text{hrs.}}{\text{day}} \times 3600\ \frac{\text{sec.}}{\text{hr.}}}$$

$$g = \frac{1.632\#/\text{sec.}}{63.4} = .0258\ \text{ft}^3/\text{sec.}$$

Bottom of Column

Solution is 45% HNO_3 by wt. 81.04 mol % water.

$$u = .603^{\dagger}\text{ cp.} = (.000672)(.603) = .000405\ \text{lbs/ft-sec.}$$

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* Estimated from I.C.T., Vol V, p10-13.
† I.C.T., Vol V, p10-13.

$u = (1.2119^*)(62.4) = 75.5\ \text{lbs/ft}^3$

$$g = (779.3 \times .8104 \times 18) + (779.3) \times (1 - .8104)63/(75.5)(24)(3600) = .0317\ \frac{\text{ft}^3}{\text{sec.}}$$

By rearrangement, equation (A) becomes

$$g = 0.575\left(\tan\frac{\alpha}{2}\right)^{0.996} H^{2.47}\left(\frac{\mu}{\rho}\right)^{0.02}(32.2)^{0.49}$$

At top of Column

$$H^{2.47} = \frac{.258}{0.575\left(\tan\frac{\alpha}{2}\right)^{0.996}\left(\frac{.000325}{63.4}\right)^{.02}(5.50)\left(\frac{32}{1}\right)} \qquad \text{no.notches/weir}$$

$$H^{2.47} = \frac{.0258}{0.575\left(\tan\frac{\alpha}{2}\right)^{0.996}(.784)(5.5)(32)}$$

$$\text{(B)}\ H^{2.47} = \frac{.000326}{\left(\tan\frac{\alpha}{2}\right)^{0.996}}$$

At bottom of Column

$$H^{2.37} = \frac{.0317}{0.575\left(\tan\frac{\alpha}{2}\right)^{0.996}\left(\frac{.000405}{75.5}\right)^{0.02}(5.50)(32)}$$

$$\text{(C)}\ H^{2.47} = \frac{.000399}{\left(\tan\frac{\alpha}{2}\right)^{0.996}}$$

The notch spacing was estimated so as to provide good liquid distribution over the tray.

Using equations (B) and (C) the following values of weir head are calculated for various values of α.

	H_1 inches		H inches at Per Cent Capacity	
Included Angle	Top Plate	Bottom Plate	75%*	30%**
30°	0.78	0.85	.09	.32
45°	0.66	0.72	.08	.28
60°	0.58	0.62	.06	.23
90°	0.46	0.50	.05	.19

* Perry, p419.

where H_1 = head above vertex of notch, inches

ΔH = decreases in weir head, H_1, caused by decrease in liquid flow rate below design or capacity flow rate (calculated at bottom tray), inches.

The variation of weir head, and hence liquid seal on the bubble caps, with flow rate is smallest when α is large. The larger angle notch is also easier to clean. Notches are made 90°, spaced $1\frac{1}{2}''$ apart and cut $\frac{3}{4}''$ deep.

Sample Calculations

α = 30° Bottom Tray

$$H^{2.47} = \frac{.000399}{\left(\tan\frac{30}{2}\right)^{0.996}} = \frac{.000399}{(.268)^{.996}} = .00149$$

$H = (.00149)^{1/2.47} = .071' = .85''$

At 75% Capacity:

$H^{2.47} = .75(.00149) = .00112$

$H = (.00112^{1/2.47} = .063' = .76''$

$h = .85 - .76 = .09''$

At 30% Capacity

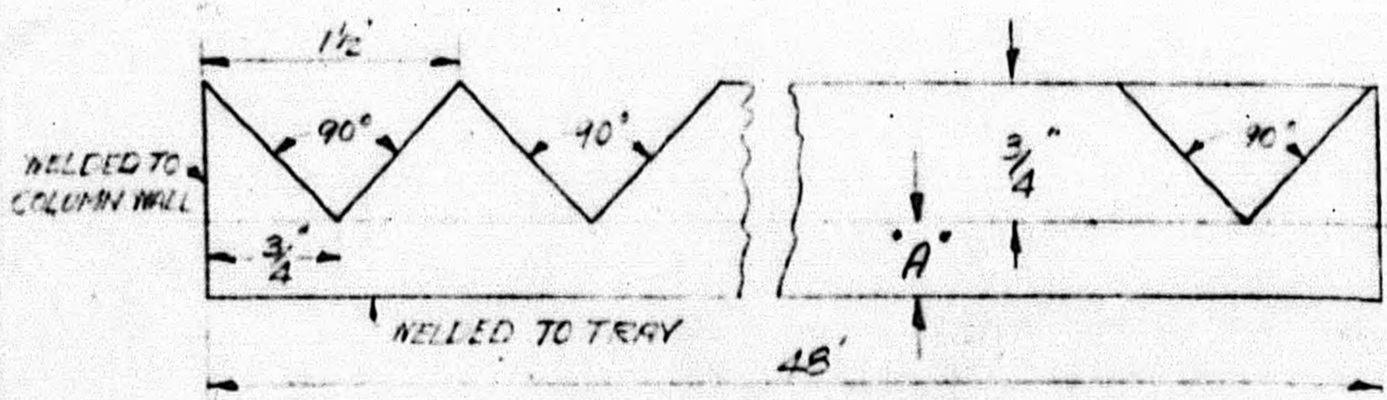
$H^{2.47} = .30(.00149) = .000447$

$H = (.000447)^{1/2.47} = .044' = .53''$

$H = .85 - .53 = .32''$

The liquid seal is $\frac{1}{2}''$ at design capacity and decreases to $(.5 - .19) = .31''$ at 30% capacity. The liquid seal of .31'' will provide efficient fractionation at the reduced production rate.

The weirs on the discharge and feed sides of each tray will be identical in design, except that the discharge weir will be mounted somewhat lower.



Liquid depth on plate = slot depth (in caps) + liquid seal = $^{13}/_{16} + ^{1}/_{2} = ^{21}/_{16} =$ $1^{5}/_{16}''$.

Discharge Weir:

Dimension "A" = $1^{5}/_{16}''$ − (head above weir)

$$\text{"A"} = 1^{5}/_{16}'' - .46 = 1.31 - .46 = .85''$$

The weir on the feed side of the tray will be set higher than the discharge weir to allow the liquid to drain down onto the tray. The dimension "A" will be $1^{1}/_{2}''$. The head of the liquid behind the feed weir is therefore $1^{1}/_{2} + .46 = 2''$ above the tray. The downflow pipe from the tray above will extend down to 1" above the tray.

Down-flow Pipes

Two pipes carry liquid from the discharge weir of one tray to the feed weir on the tray below. Downflow pipes are welded flush with the upper tray and extend to 1" above the lower tray.

The head of liquid in the section of the tray behind the discharge weir must be less than the dimension "A" (p22) which is 0.85".

A head of .60" above the downflow pipes will be assumed in order to calculate the diameter of the pipes.

$V = KLH^{*1.42}$ (Perry p1454)

where V = liquid flow rate, ft³/sec.
H = head over pipe edge, ft.
L = length of weir perimeter (pipe perimeter) ft.

$$V = \frac{.0317\dagger \ \text{ft}^3/\text{sec.}}{2 \text{ pipes}} = .0158 \ \frac{\text{ft}^3}{\text{sec.}} \text{ per pipe}$$

$$.0158 = 3(L)\frac{(.60)}{12}1.42 = 3L\ (.014)$$

$L = .376' = 4^{1}/_{2}''$

If D = pipe dia., in.

$\Pi D = 4.5$

$D = 1.43''$

Standard $1^{1}/_{2}''$ stainless steel pipe will be used (i.d. 1.61"), schedule 40. Downflow pipes will be welded with $1^{1}/_{2}''$ clearance between pipe and column wall.

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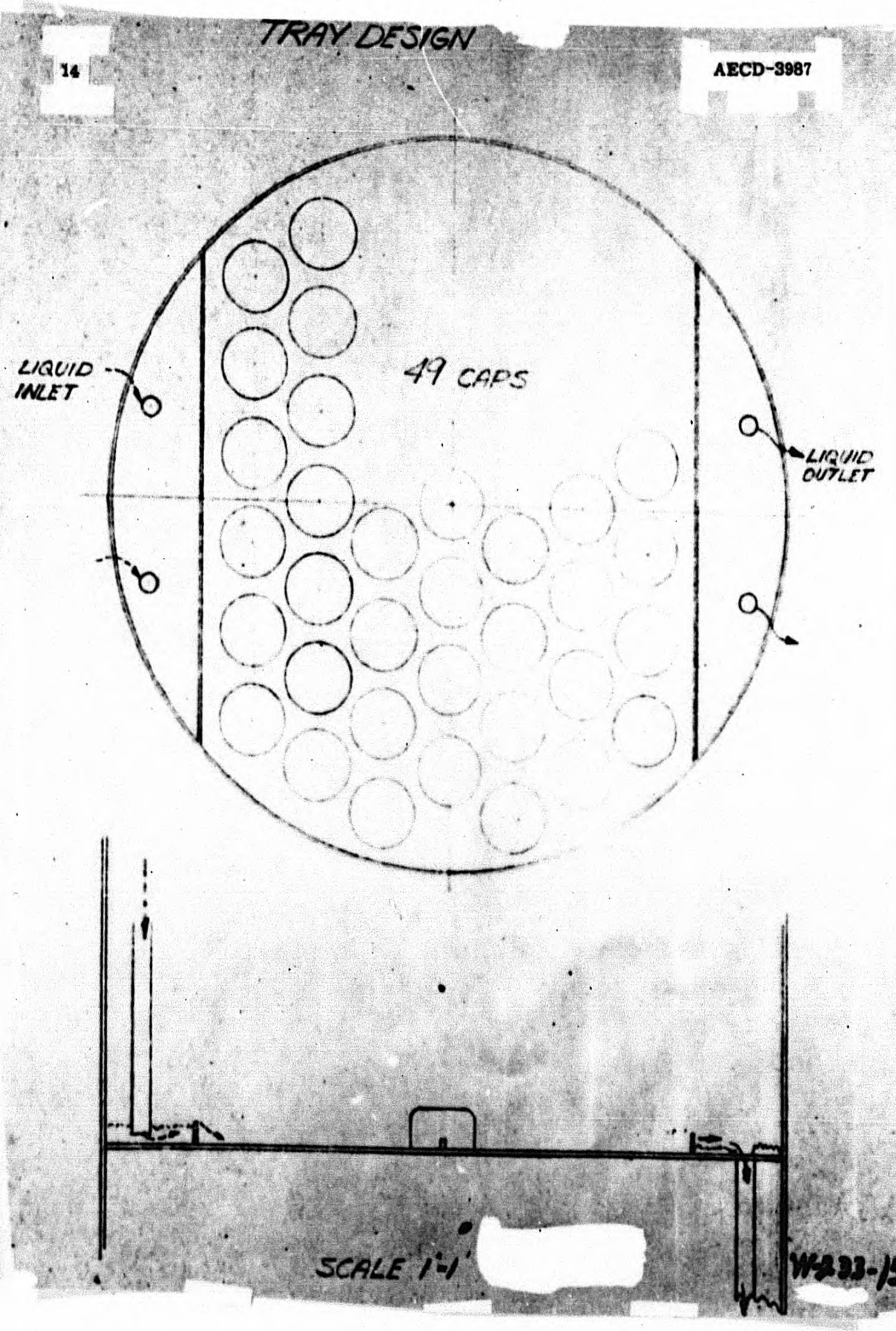
*K is a constant. 3.0. See 1454 Perry.
†P 19.


TRAY DESIGN
LIQUID INLET
49 CAPS
LIQUID OUTLET
SCALE 1"-1'
W-233-15

Liquid Gradient Across Trays

As an approximation, it will be assumed that liquid flows across each tray as liquid flows in a channel or open duct. As average of seven such channels, each about $4\frac{1}{2}$ feet long, $1\frac{1}{2}''$ wide, and filled to a depth of $1\frac{1}{4}''$ exist across each tray (see drawing, page 14).

Checking Formula for Open Ducts*

$$V = C\sqrt{m\frac{F}{L}}$$

where V = linear velocity, ft/sec.

$C = \sqrt{\frac{\alpha g e}{f}}$ where f = Fanning friction factor

m = hydraulic radius

F = liquid head, ft. of liquid

L = length of channel, ft.

For each channel:

$$m = \frac{\text{cross section area}}{\text{wetted perimeter}} = \frac{(1.25)(1.50)}{12\,\frac{\text{in.}}{\text{ft.}}\;2(1.25) + 1.50}$$

$$m = \frac{1.875}{(12)(4)} = .039$$

$$V = \frac{.0317^{\text{ft}^3/\text{sec.}}}{7\text{ channels}}\;\frac{(144)}{1.875}\;\frac{L}{\text{ft}^2} = .348\text{ ft/sec.}$$

$$Re = \frac{DV\rho}{\mu} = \frac{4m\,V\rho}{\mu} = \frac{4(.030)(.348)\,\frac{(75.5\text{ lbs})}{\text{ft}^3}}{.000405}$$

$$Re = 1.01 \times 10^4$$

$$f = (\text{p811, Perry}) = 0.0085$$

$$C = \sqrt{\frac{(2)(32.2)}{.0085}} = 87.0$$

$$V = C\sqrt{m\,F/L} = 87.0\sqrt{\frac{.039 \times F}{4.5}} = .348$$

$$\frac{.039^F}{4.5} = 16 \times 10^{-6}$$

$$F = \frac{(16 \times 10^{-6})}{.039}(4.5) = .0019\text{ ft.}$$

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*Perry's Handbook, p806.

$F = .0019 \times 12 \frac{\text{in.}}{\text{ft}} =$ .02″ water liquid head across plate.

Shell Thickness

The A.P.I.-ASME Code for Unfired Pressure Vessels gives a graph* relating Y and X where:

$$Y = \frac{(27500)}{Sy} \times (p)$$

$$X = \left(\frac{29 \times 10^6}{E}\right) \times \left(\frac{Sy}{27500}\right) \times \left(\frac{L}{D}\right)$$

where E = modulus of electricity, lbs/in^{-2}
Sy = yield strength at operating temperature, psi.
L = distance between stiffeners, measured parallel to axis of vessel, inches.
D = diameter, in.
P = max. external working pressure psi.

The value of E is 28.8×10^6 psi, taken from graph 26, p.85 of VPV codes (1943). The value of Sy is found to be 27,500 psi from Fig. 25, of above reference. The tray spacing, L, is 24″, and the max. external pressure is, atm.(15 psi).

$$Y = \frac{27{,}500}{27{,}500} \times 15 = 15$$

$$X = \frac{29 \times 10^6}{28.8 \times 10^6} \times \frac{27{,}500}{27{,}500} \times \frac{24}{72} = .335$$

From graph, $\frac{(t-c)}{D} = .0023$

where t = shell thickness, inches
c = corrosion allowance, inches.

A corrosion allowance of .25″ was estimated from data in "Materials of Construction for Chemical Engineering Equipment" by Chem. & Met., Stainless Steel type 347, is assumed.

$$\frac{(t - .25)}{72} = .0023$$

$t - .25 = .167$

$t = .417''$

Shell and plates shall be ½″ stainless steel.

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* 1943 Edition, p86.

Heat Balance—Calandria Heat Duty

At 100% capacity:

Heat in = Heat out

Preboiler + $h_fF = DH_D + Wh_w$

where Qreboiler = heat duty of reboiler, or calandria, Btu/day.
h_f = enthalpy of feed, Btu/#mol.
H_D = enthalpy of distillate, Btu/#mol.
h_w = enthalpy of product, Btu/#mol.

$$\begin{array}{l} \quad\quad\quad h_f \quad\quad F \quad\quad\quad\quad\quad D \quad\quad\quad\quad\quad\quad \text{Product} \\ \text{Qreb.} + (2200)(7382) = (6603)(18{,}120) + (779.3)(2700) \end{array}$$

Qreb. + 16,240,000 = 119,646,000 + 2,104,100

$$\begin{array}{rr} & 119{,}646{,}000 \\ & \underline{2{,}104{,}100} \\ & 121{,}750{,}100 \\ \text{Qreb.} = & \underline{16{,}240{,}000} \\ \text{Qreb.} = & 105{,}510{,}100 \text{ Btu/day.} \end{array}$$

At 75% capacity (June 1949 production rate)

.75 (105,510,100) = 79,132,500 Btu/day.*

The calandria now in use might supply this heat.

Cost Estimate for Bubble-Cap Column:

Column diameter = 72 in.
No. Trays = 10
Column Height = 26''

The graph of installed cost vs. column diameter (Data reported in Chemical Engineering, June 1949) gives two methods of estimating cost:

Cost/tray	Both equal $1500
Cost/ft. height	from graph

An average of these two methods is calculated:

$$\begin{array}{rr} 10 \times 1500 = & \$15{,}000 \\ 26 \times 1500 = & \underline{\$39{,}000} \\ 2 & \underline{\$54{,}000} \\ & \$27{,}000 \end{array}$$

* This compares with about 65,000,000 Btu/day, the packed column now in use (yield ≈ 60%).

Probable Installed Cost: (Including instruments and manholes.)	\$27,000
Instruments (10% installed cost)	2,700
Manholes (10 manholes @\$500)	5,000
Contingencies (20% installed cost)	5,400
Total	\$40,100

Summary of Column:

Specifications:

Shell: Diameter: 6' 0" i.d.
Thickness: 1/2"
Height: 26'
Ends: Dished
Material: Stainless Steel 347
Construction: All welded (double welded butt joints)

Trays: Diameter: 6' 0"
Thickness: 1/2"
Risers: No. – 49
Diameter – 5"
Arranged on equilateral triangles, 8 1/2" on each side.
Caps: No. – 49
Diameter – 7"
Perimeter cut away for rectangular slots – 16 1/2"/cap
Slot Depth – 13/16"
Material: Caps, plates and risers all 347 stainless steel
Construction: Risers welded to trays, trays welded to column shell, each cap bolted to riser.
Overflow Weirs: (2 needed)
Length: 48"
Notches: No. – 32
Notch Spacing – 1 1/2"
Included Angle of notch – 90°
Depth – 3/4"
Distance of Vertex of Notch above
Plate – .85" – Discharge weir
1 1/2" – Feed weir
All notches filed to sharp edges.
Material: Stainless 347
Construction: Weirs are welded to tray and column shell.
Thickness: 1/4"

Downflow Pipes:
No. – 2 between trays (Total no. – 18)
Diameter – 1 1/2" std. stainless 347 pipe Schedule 40

Construction—Pipes welded flush with upper plate, extend to 1" of tray below. A clearance of $1\frac{1}{2}$" is to be permitted between downflow pipes and column shell. Each pair of p downflow pipes are separated by $17\frac{1}{2}$".

Material: 347 Stainless

Total Cost: $40,000, excluding calandria and condenser.