How to design Iping for reboiler systems

Interactions between hydraulic requirements and piping configurations require close attention to many fluid and mechanical details, in order to obtain the most efficient and economical distillation units.

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O Familiarity with graphic piping design is an essential requirement for the designer of hydraulic systems. The accuracy of his calculations, predictions of flowrate and pressure differential, reliability oCoperation, and the economy of capital, energy, maintenance and operating costs depend to a great extent on pipe configurations and pipe components:

In these articles, we have recognized. the importance of graphic piping design, and to a limited degree have presented its fundamentals. We will now. evaluate the flow systems and piping design for a distillation column, which is a more integrated unit than the individual systems discussed in our earlier articles.

Layout for distillation columns

A process flow diagram of a typical distillation column with bottom pump, thermosyphon teboiler, overhead condenser, reflux' drum and reflux. pump are shown in Fig. 1 (F/1). The equipment components are located adjacent to each other in the actuaLplant. Also in *FII,* we find elevation and plan drawirigsfor the column. These show how the principal eleIll.ents of a distillation column are usually integrated into an overall plant arrangement. Manholes face access roads (or access aisles at housed installations). Each manhole has a platform for maintenance. Valves and instruments are' located above these platforms for convenient access.

For economy and easy .support; piping should drop immediately upon leaving the tower nozzle, and run parallel, and as close as possible, to. the tower itself. A vertical line lends itself as a. suitable location for the straight run of an.orifice, The horizontal elevations, after the lines leave their vertical run, are governed by the elevations of the main pipe rack. Lines that run directly to equipment at grade (more or less in the direction of the main pipe rack) often have the same eleyation as the pipe bank.

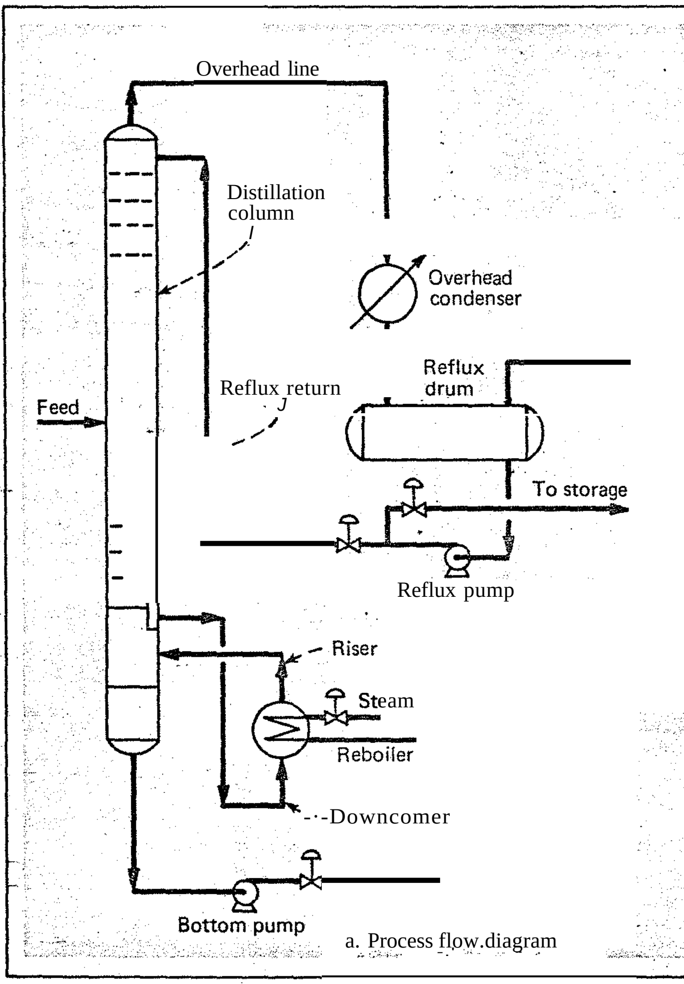
Lines from tower nozzles below the pipe rack should approach the pipe bank roughly 2 ft below the pipe-rack elevation. The same elevation is used for those lines that run to pumps located below the pipe rack.

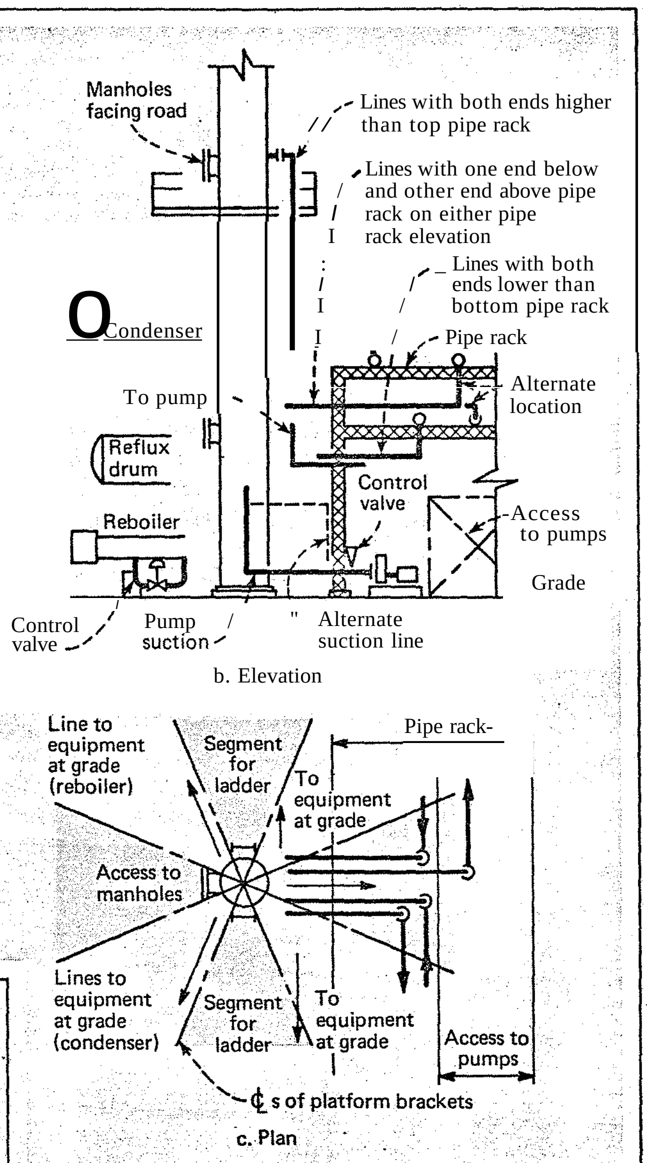
Pump-suction lines can also be arranged on this elevation. They should be as short as possible and run without loops or pockets. Pipelines, droppingJrom above the pipe-rack elevation, will approach the pipe bank roughly 2 ft higher than the elevation of the main pipe bank. This elevation is also used for steam lines to reboilers. These steam lines usually connect to the top of the headers to avoid excessive condensate drainage toward process equipment.

The plan view *(F/*1) of the tower shows the segments of its circumference allotted to piping, nozzles, manholes, platform brackets and ladders. Such a pattern usually leads to a well-organized arrangement for the process equipment and auxiliary components.

From a layout. standpoint, it is preferable to have equal platform-bracket spacing, and the orientati0Ilof brackets lined up along the entire length of the tower. This will minimize interferences between the piping and structural members. According to OSHA, ladders between platforms should not be longer than 30 ft.

Area segments for piping going to equipment at grade are available between the ladders and on both sides of the manholes. Lines approaching the pipe rack





Piping the distillation column requires economy, ease of support and accessibility for good design.

can turn left or right, depending on the'plant's overall arrangement.

On a distillation column, the largest lines are the overhead vapor line, and the reboiler downcomer and return. These lines should have the simplest and most direct configurations, to ininimize pressure loss and cost.

During normal operation, the pump in Fig. 1 transports liquid at equilibrium. This means that the distillation column and reflux drum are elevated to satisfy NPSH (net positive suction head) requirements. The discharge lines often have two destinations. Total-head requirements should be designed and calculated so that operating points fall on the pump's head-capacity curve when pumping to an alternative destination. Alternative discharge lines can have equal capacity and alternative operation, or partial capacity with simultaneous operation. All alternatives should be investigated for the process pumps.

The design for a pumped reboiler circuit is similar to that of a reflux pump system. A bottom pump transports the liquid through an exchanger or fired heater and returns it to the distillation column. Close attention should be paid to possible two-phase flow in pipelines coming after the heater—especially when we want to locate the heater close to the column.

Inserted-type reboilers have no process piping. Larger-diameter towers can have one to four U-tube stub bundles inserted directly into the liquid space through tower nozzles, and extending across the tower diameter. Reboilers with small heat duties are usually designed as helical coils.

Reboiler arrangements

In horizontal thermosyphon reboilers, liquid flows from an elevated drum or tower bottom or tower-trapout boot through a downcomer pipe to the bottom of exchanger shell. The liquid is heated, leaves the reboiler in the return piping as a vapor-liquid mixture, and flows back to the tower or drum.

In vertical reboilers, heating usually occurs on the shellside. In horizontal reboilers, heating is on the tubeside. For a large evaporation rate (for example, 90% of total flow), a kettle-type reboiler is used.

Piping to horizontal reboilers is designed as simply and directly as possible within the limitations of thermal expansion forces.

Symmetrical arrangements between the drawoff and reboiler-inlet nozzles, as well as between the reboiler outlet and return connection on the tower, are preferred in order to have equal flow in the reboiler circuit. A nonsymmetrical piping configuration may also be accepted for a more-economical or more-flexible piping design.

Reboilers often have two outlets and two parallel-pipe segments. When sizing and arranging nonsymmetrical piping; an attempt should be made to equalize the resistance through both legs of the reboiler piping. More resistance in one leg produces a smaller flow than in the other. Hence, uneven heat distribution will occur in the reboiler-one segment of the riser will be hotter than the other.

At startup in reboilers having high, liquid drawoff nozzles, a gravity-flow bypass is usually provided from the tower's liquid space. to a low point of the downcomer.

Valves are rarely included in reboiler piping, except when a standby reboiler is provided, or when two or three reboilers are used and operated at an extremely wide heat-capacity range. Some companies require line blinds to blank off the tower nozzles during shutdown, turnaround and maintenance.

The heating media (steam or a hot process stream) connect to the tubeside of horizontal reboilers. The inlet piping usually has a temperature-regulated control valve (with block valves and bypass globe valve, if required). This is normally arranged at grade near the reboiler's tubeside inlet.

Reboiler elevations

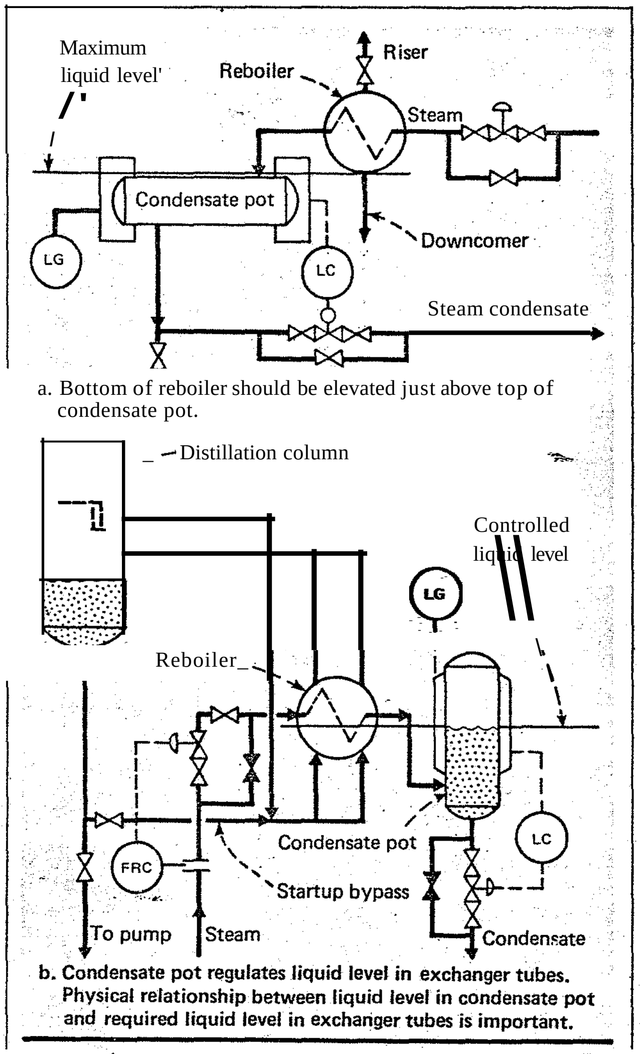
Most reboilers are at grade next to the tower; with centerline elevations of about 3 to 5.5 ft above ground level for exchangers about 1 to 3 ft dia. Exchangers at grade provide economical arrangements-valves and instruments are accessible, tube-bundle handling is convenient, and maintenance is easy. In this arrangement, the static heads are well determined between the exchanger's centerline and the drawoff and return nozzles on the tower. Vertical reboilers are usually supported on the distillation column itself.

Some reboilers have a condensate or liquid-holding pot located after the tubeside outlet, as shown in F2. In such cases, the centerline elevation of the reboiler is somewhat higher than units that do not have these control vessels.

The arrangement in F2a is a high-capacity steam trap. The top of the condenser pot should not be higher than the bottom of the exchanger shell, to avoid flooding the tubes with condensate and adversely affecting the exchanger's heat-transfer duty.

The arrangement in F2b maintains a required condensate level in the reboiler, to provide for a wide range of heat-transfer control. Process conditions determine the precise relationship between the exchanger and the vertical condensate-control pot.

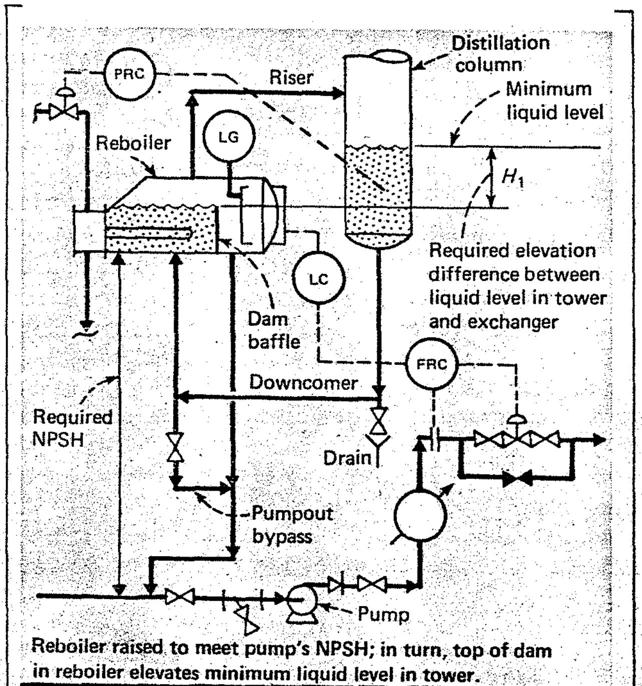
In F3, we show an example where a reboiler has been elevated to meet the NPSH requirement of the centrifugal pump. The elevated reboiler, in turn, raises the tower the minimum liquid level in the bottom of the tower must be higher than the liquid level in the exchanger. The elevation difference (dimen-



a. Bottom of reboiler should be elevated just above top of condensate pot.

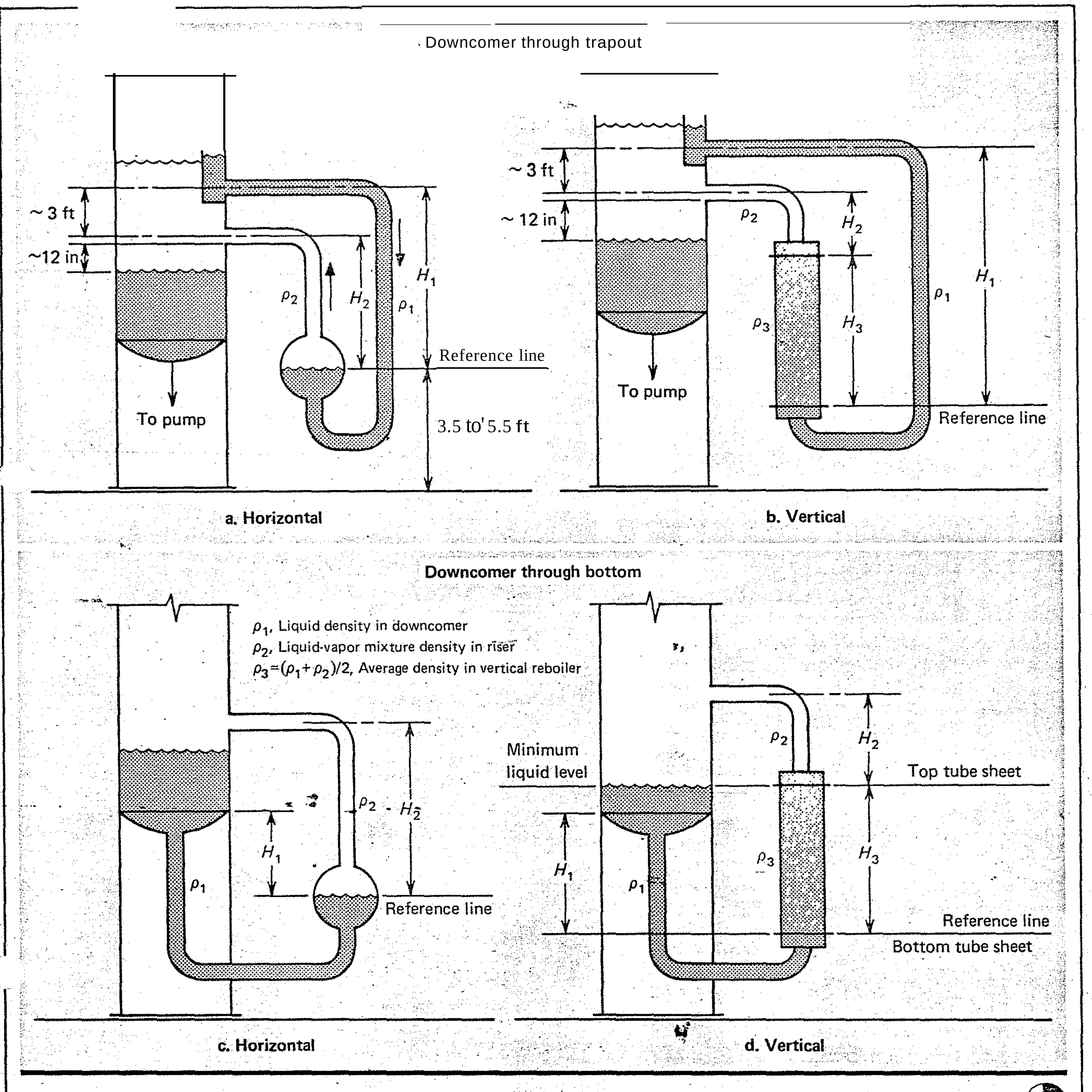
b. Condensate pot regulates liquid level in exchanger tubes. Physical relationship between liquid level in condensate pot and required liquid level in exchanger tubes is important.

Process **conditions determine equipment** elevations, F2



Reboiler raised to meet pump's NPSH; in turn, top of dam in reboiler elevates minimum liquid level in tower.

Reboiler raised to meet pump's NSPH requirement. F3



Reboilers having natural circulation rely on the difference in static head between the downcomer and riser. F4

sion H_1 in *F/3)* provides the positive static head for flow in the reboiler circuit, and overcomes friction losses in the exchanger, and downcomer and return lines.

Designing the reboiler system

Typical reboiler arrangements are shown ically in *F/4*. In all cases, the vessel pressure is the same at the tower's outlet and return nozzles. Circulation is forced by the static-head difference between the liquid column in the downcomer and the vapor-liquid column in the riser. For convenience, reference lines are chosen at the exchanger's centerline for horizontal reboilers, and at the bottom tubesheet for vertical reboilers.

If P_1 is liquid the downcomer at the reference line, and P_2 is backpressure in the riser's vapor-liquid column; the pressure difference ($\Delta P = P_1 - P_2$) must overcome the exchanger and piping friction losses. Therefore, P_1 must be greater than P_2. If ρ_1 is the hot liquid density in the downcomer; then $\rho_1 H_1/144 = P_1$, psi. The backpressure, P_2, can have two alternative expressions:

1. For horizontal exchangers (see F/4a and F/4c):

$$P_2 = \rho_2 H_2/144, \text{ psi} \qquad (1)$$

where:

$$\rho_2 = \frac{W}{\dfrac{W_l}{\rho_l} + \dfrac{W_v}{\rho_v}} = \frac{100}{\dfrac{\%\ Liquid}{\rho_l} + \dfrac{\%\ Vapor}{\rho_v}} \qquad (2)$$

2. For vertical exchangers (see *F/4b* and *F/4d)*:

$$P_Z' = (p_Z H_Z + p_3 H_3)/144, \text{ psi} \qquad (3)$$

where p_Z is again the mixture's density, as expressed by Eq. (2) for horizontal exchangers, and p_3 is the average density of liquid and liquid-vapor mixture in the reboiler:

$$\rho_3 = (\rho_1 + \rho_2)/2 \qquad (4)$$

Eq. (4) provides a conservative estimate of the density gradient in vertical reboilers. Actual density will be less than that expressed by Eq. (4). In all equations, the units for p are lb/ft^3, and for H, ft. We also note that the vertical reboiler should be flooded. The maximum elevation of the top tubesheet should not be higher than the minimum liquid level in the tower.

Hydraulics in horizontal reboilers ..

In the following discussion, the hydraulic conditions only in horizontal exchangers will be developed. (The derivations are the same for vertical exchangers, except that P_Z' will replace P_Z.) For horizontal exchangers:

$$P_I - P_Z = \Delta P = (1/144)(p_I H_I - p_Z H_Z) \qquad (5)$$

If a safety factor of 2 is introduced, then the available pressure difference for friction losses is halved, and:

$$\Delta P = (1/288)(p_I H_I - p_Z H_Z)\,. \qquad (6)$$

The quantity $(H_I - H_Z)$ is usually 3 ft (see F14a). Consequently, a minimum driving force of $\Delta P_{min} = (3/288)p_I \approx 0.01 p_I$ is always available at horizontal exchangers.

The maximum possible driving force depends on the elevation difference between the drawoff nozzle and exchanger centerline (dimension H_I) and on the total evaporation taking place in the reboiler. Neglecting the vapor-column backpressure in the return line, the maximum usable driving force is:

$$\Delta P_{max} = (H_1/288)\rho_1 \qquad (7)$$

In most applications, the actual driving force is not much below this maximum. H_I can range from 6 to 24 ft, depending on the size of the arrangement and on NPSH for a pump taking suction at the bottom of the tower. For these H_I values:

$$\Delta P_{max} = (6/288)p_I \text{ to } (24/288)p_I$$

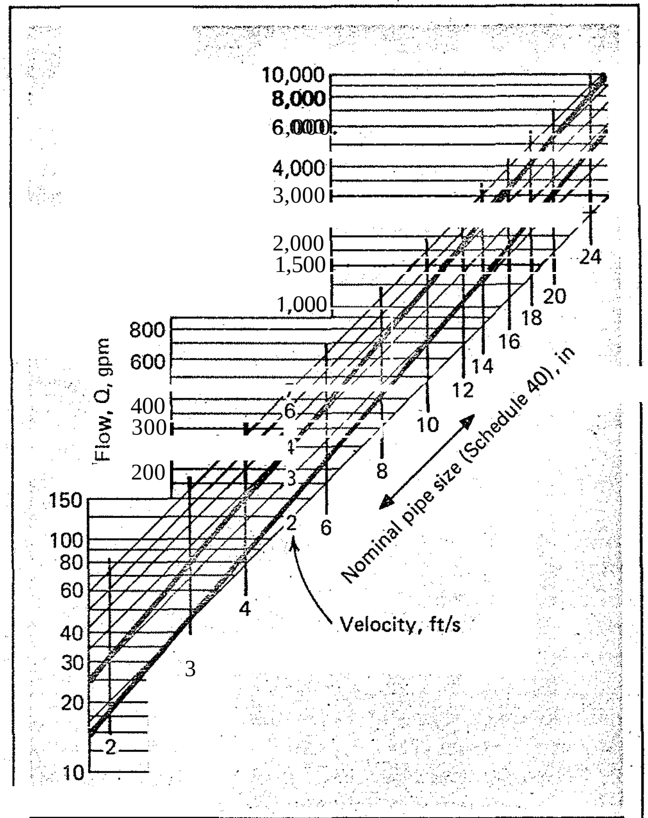
$$\Delta P_{max} = 0.02 p_I \text{ to } 0.08 p_I$$

Thus, the driving force is reduced to a function of the downcomer liquid density at operating temperature. For example, if the piping geometry produces $H_1 = 12$ ft, and $p_I = 50$ lb/ft^3 for kerosene:

$$\Delta P_{max} = (12/288)50 \approx 2.0 \text{ psi}$$

These simple relationships are useful when the evaporation rate is not known and line sizes have to be estimated. The available driving force will be near but less than ΔP_{max}.

ΔP_{max} as evaluated here is, of course, an extreme value taken at total evaporation. In reboilers, partial evaporation usually takes place, which will reduce ΔP_{max}. However, even if the driving force is assumed



Shaded part of chart establishes size of downcomer. F/5

at the maximum value, any inaccuracy is well compensated for by the safety factor of 2, and by the necessity to use commercially available pipe sizes that are normally larger than calculated pipe diameters.

Friction losses in reboilers

The total friction losses in a thermal-circulating reboiler system must be smaller than the available driving force. The pressure loss caused by friction takes place in two main locations: in the exchanger itself, Δp_e, and in the piping, Δp_p. Hence:

$$\Delta p_e + \Delta p_p < \Delta P$$

Friction losses in reboilers, Δp_e, are generally given as 0.25 to 0.5 psi. (A note should indicate whether entrance and exit losses are included.) Unit losses in downcomers and risers are 10 fractions of 1 psi/100 ft.

Calculation procedures for liquid lines have been outlined in Part 2 of this series, *Chem. Eng.*, Jan. 6, 1975, pp. 115–120 (Example 1); and for the two-phase flow risers in Part 8, June 23, 1975, pp. 145-151 (Example 1). We calculate reboiler returns as dispersed flow, regardless of the intersection of the Baker parameters (e.g., Fig. 1, Part 8). To avoid trial-and-error calculations, a selection chart for reboiler pipe size is presented in F/5. This chart is based on limiting velocities for flow in downcomers of 2 to 7 ft/s. We enter the graph with known liquid-flow quantities. We obtain downcomer pipe sizes from the shaded portion of the graph, and also find the corresponding flow velocities for computing the

Reynolds numbers. The riser can be assumed as one or two sizes larger than the downcomer pipe.

. In vertical reboiler circuits, reboiler losses are greater, and pipe losses smaller, than in horizontal circuits. In this case, a safety factor of 1.25 applied to the driving force can be used. In kettle-type reboilers, evaporation rates are high. For these reboilers, a large-diameter return line is usually necessary.

Elevation of the drawoff nozzle

The minimum elevation, H_1, for the downcomer tower nozzle above that for the centerline of the horizontal reboiler may be found from Eq. (6), where $H_2 = H_1 - 3$, ft:

$$H_1 = \frac{288\Delta p - 3\rho_2}{\rho_1 - \rho_2} \qquad (9)$$

The downcomer nozzle cannot be lower than H_1. Δp replaces ΔP in Eq. (6), and is the sum of the downcomer, riser and exchanger friction-losses:

$$\Delta p = \Delta p_d + \Delta p_r + \Delta p_e$$

The value of H_1 is useful when elevation adjustments are made to vessel heights during graphic piping design, or when the vessel can be located at a minimum elevation. The coefficient for ρ_2 in Eq. (9) is the elevation difference between the downcomer and riser nozzles. If this is other than 3 ft, the correct dimension should be inserted.

Many towers have a bottom drawoff pump. NPSH requirements usually, cause the process vessel and the reboiler. drawoff nozzle to be raised higher than that of the reboiler's minimum elevation. This increases the static head in the vertical legs, and also the driving force in the circuit. With increased tower height, it is worthwhile to check the reboiler circuit for a possible reduction in size of the liquid and return lines-'especially where large-diameter lines are necessary. Uneconomical reboiler lines are just carelessly oversized or poorly routed.

Example demonstrates calculations

Let us size the reboiler lines for the kerosene distillation unit, as sketched in F/6. Flow data are:

	Downcomer	Riser,	
	Liquid	Liquid	Vapor
Flowrate, W, lb/h	85,000	59,500	25,500
Density (hot), ρ, lb/ft³	36.7	36	1.31
Viscosity, μ, cp	0.6	0.5	0.01
Molecular weight, M	—	—	53

The flow data for the riser reflect that 70% of total flow is liquid and 30% vapor. We obtain the vapor density,* ρ_v, in the riser from:

$$\rho_v = MP'/10.72Tz$$

$$\rho_v = 53(181.7)/(10.72)(682) = 1.31 \text{ lb/ft}^3$$

We find the mixture density,* ρ_2, in the riser by substituting in Eq. (2):

$$\rho_2 = \frac{100}{(70/36) + (30/1.31)} = 4.05 \text{ lb/ft}^3$$

And, we calculate volume flowrate,* Q, from:

$$Q = W/500S$$

$$Q = 85{,}000/500(36.7/62.37) = 289 \text{ gpm}$$

We will begin by sizing the downcomer and then calculate the overall pressure loss in it. We follow up with similar computations for the riser-remembering that the line has vapor/liquid flow. The simplest part of this analysis is finding the preliminary size of the reboiler lines. The computations for checking whether these line sizes are adequate require considerable detail. We should also note that the reboiler has two inlets and two outlets (F/6). Consequently, we find full flow (100%) in each line from the column to the "tee" connection on the reboiler's inlet and outlet piping; but only 50% of total flow in each line segment after the tee. In the following procedures, we will see how this flow arrangement affects our design calculations.

Downcomer-For a liquid flow of 289 gpm, we find the pipe size for the downcomer as 6 in from the selection chart of F/5. (For a 6-in, Schedule 40 pipe, I.D. = 5.761 in, $d^5 = 6{,}346$ in⁵.)

In order to calculate the unit loss for the 100% flow section, we must calculate the Reynolds number from:

$$N_{Re} = 50.6(Q/d)(\rho/\mu)$$

$$N_{Re} = 50.6\left(\frac{289}{5.761}\right)\left(\frac{36.7}{0.6}\right) = 155{,}300$$

For this Reynolds number, we obtain the friction factor, f, as 0.0182 from the chart in Part 8 of this series (*Chem. Eng.*, June 23, 1975, p. 147), and calculate the unit loss from:

$$\Delta p_{100} = 0.0216 f \rho_1 (Q^2/d^5)$$

$$\Delta p_{100} = 0.0216(0.0182)(36.7)[(289)^2/6{,}346]$$

$$\Delta p_{100} = 0.19 \text{ psi/100 ft}$$

We can then find the unit loss at 50% flow:

$$N_{Re} = 0.5(155{,}300) = 77{,}650;\ f = 0.022$$

$$\Delta p_{100}(50\%) = 0.19\left(\frac{144.5}{289}\right)^2\left(\frac{0.022}{0.0182}\right)$$

$$\Delta p_{100}(50\%) = 0.0574 \text{ psi/100 ft}$$

We now determine the equivalent length of pipe and fittings for each segment from tables in Part 2 of this series (*Chem. Eng.*, Jan. 6, 1975, pp. 115–120), as follows:

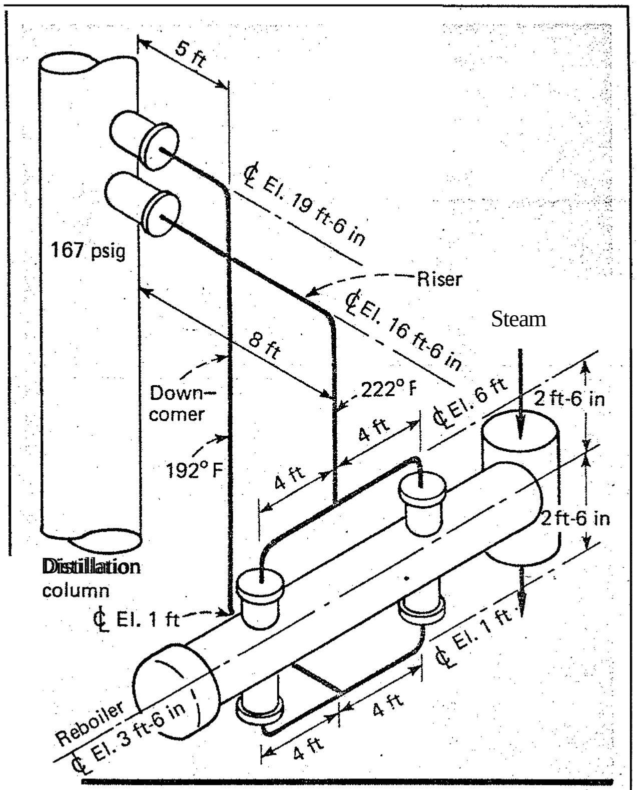
	Segment for 50% Flow, Ft	Segment for 100% Flow, Ft
Actual length	6	26
Entrance loss		18
Elbows*	10	20
Sharp tee	30	—
Exit loss	36	—
Total	82	64

*One elbow for 50% flow segment, 2 for 100% flow.

Overall pressure loss of the downcomer:

$$\Delta P = 0.19(64/100) + 0.057(82/100) = 0.167 \text{ psi}$$

*See Part 1 of this series, *Chem. Eng.*, Dec. 23, 1974, pp. 58–66, for complete details for these calculations.



Reboiler flow relations and elevations for problem.

Riser—Since the downcomer is a 6-in pipe, let us use a riser having a nominal size of 8 in. (For an 8-in Schedule 40 pipe, I.D. = 7.981 in, d^5 = 32,380 in.^{5}, A = 0.3474 ft^2.)

In order to calculate the unit loss for the 100% flow section, we must calculate the vapor-phase Reynolds number from:

$$N_{Re} = 6.31\, W_v/d\mu_v$$
$$N_{Re} = 6.31(25,500)/(7.981)(0.01)$$
$$N_{Re} = 2 \times 10^6$$

For this Reynolds number, we obtain a friction factor, f, of 0.014 from the chart in Part 8 of series; and then calculate the vapor-phase unit loss from:

$$\Delta p_{100} = 0.000336(f/\rho_v)(W_v^2/d^5)$$
$$\Delta p_{100} = 0.000336(0.014/1.31)[(25,500)^2/32,380]$$
$$\Delta p_{100} = 0.072 \text{ psi/100 ft}$$

Since the riser handles vapor/liquid flow, we must now determine the two-phase flow modulus* from:

$$X^2 = (W_l/W_v)^{1.8}(\rho_v/\rho_l)(\mu_l/\mu_v)^{0.2}$$
$$X^2 = \left(\frac{59,500}{25,500}\right)^{1.8}\left(\frac{1.31}{36}\right)\left(\frac{0.6}{0.01}\right)^{0.2} = 0.38$$

With this value of X^2, we find the two-phase flow modulus, ϕ^2, as 14 from a chart in Part 8 of this series.* We calculate the two-phase-flow unit loss* from:

$$\Delta p_{100} = 0.072 \times 14 = 1.01 \text{ psi/100 ft}$$

*See Part 8 of this series, *Chem. Eng.*, June 23, 1975, pp. 145–151 for complete details of these calculations.

Let us estimate that the unit loss in the 50% flow section is one-third that in the 100% flow segment. Hence:

$$\Delta p_{100}(50\%) = 0.34 \text{ psi/100 ft}$$

Again, we determine the equivalent length of pipe and fittings for each segment of the riser as follows:

	Segment for 50% Flow, Ft	Segment for 100% Flow, Ft
Actual length	6	18
1 Elbow	14	14
1 Sharp tee	24	
1 Entrance loss	24	
1 Exit loss		48
Total	68	80

Overall pressure loss of the riser:

$$\Delta P = 1.01(80/100) + 0.34(68/100) = 1.04 \text{ psi}$$

In summary, the total pressure loss is obtained by adding the loss in the downcomer, riser and reboiler:

6-in Downcomer. .	0.167 psi
8-in Riser	1.02 psi
Reboiler	0.35 psi
Total ΔP	1.557 psi

By substituting the appropriate values into Eq. (6) for this example, we determine the available pressure difference, as:

$$\Delta P = (1/288)[(36.7)(13.5) - (4.05)(10.5)] = 1.57 \text{ psi}$$

The available pressure difference of 1.57 psi is greater than the calculated pressure losses of 1.557 psi. Therefore, the design and sizes are acceptable.

Finally, we check the minimum elevation of the drawoff nozzle above the centerline of the reboiler by substituting into Eq. (9):

$$\frac{288(1.535) - 3(4.05)}{31.7 - 4.05} = 15.55 \text{ ft}$$

Since the drawoff nozzle is actually 16.0 ft above the reboiler's centerline (F/6), the minimum value of 15.55 ft is acceptable.

The next article in this series will appear in the Sept. 15, 1975, issue. This article will review the design of pipelines for the hydraulic and thermal conditions occurring in overhead condensing systems.

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