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Revision Memo

- 08/04
- Highlights of revisions:
 - Equation changes on figure 1

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SCOPE

This subsection covers the use of flow maps for determining two-phase (vapor-liquid) flow regimes and calculation methods for determining pressure drop and liquid holdup for isothermal two-phase flow in horizontal and vertical pipes. Methods for calculating pressure drop across orifices, valves, pipe fittings, as well as sudden enlargements and contractions, are also included. Guidelines for designing manifolds and perforated pipes for distributing vapor-liquid mixtures are also given. These methods are considered to be among the best currently available for general use by simple hand calculations.

REFERENCES

DESIGN PRACTICES (SUBSECTION A, B, AND C OF THIS SECTION)

OTHER LITERATURE

1. O. Baker, "Multiphase Flow in Pipelines," Oil and Gas Journal, 56, 156-67 (November 10, 1958).
2. A. E. Dukler, et. al., "Frictional Pressure Drop in Two-Phase Flow," AIChE Journal, 10, 38-51 (1964).
3. AGA/API Monograph on Project NX-28, "Gas-Liquid Flow in Pipelines."
 - a. Vol. I, Research Results, A. E. Dukler (May, 1969).
 - b. Vol. II, Design Manual, O. Baker, et. al (October, 1970).
4. O. Flanigan, "Effect of Uphill Flow on Pressure Drop in Design of Two-Phase Gathering Systems," Oil and Gas Journal, 56, 132-141 (March 10, 1958).
5. G. W. Govier and K. Aziz, "The Flow of Complex Mixtures in Pipes," Krieger, New York (1977).
6. J. L. Greene, "Symmetrical Piping Arrangement Solves Two-Phase Flow Distribution Problems," Hydrocarbon Processing, 46, 141-43 (1967).
7. K. C. Hong, "Two-Phase Flow Splitting at a Pipe Tee," J. Petroleum Technology, 290-296 (February, 1978).
8. G. A. Hughmark and B. S. Pressburg, "Holdup and Pressure Drop With Gas-Liquid Flow in a Vertical Pipe," AIChE Journal, 7, 677-82 (1961).
9. J. C. Leung, "A Generalized Correlation for One-Component Homogeneous Equilibrium Flashing Choked Flow," AIChE Journal, 32, 1743-46 (1986).
10. F. N. Nazario, "Sizing Pressure Relief Valves in Flashing and Two-Phase Service - an Alternative Procedure," EE.28E.90 (April, 1990).
11. L. L. Simpson, "Sizing Piping for Process Plants," Chemical Engineering, 75, 192-214 (June 17, 1968).
12. Y. W. Wong and R. D. Patel, "Improved Vapor-Liquid Manifolds," EE.23E.88 (February, 1988).
13. Y. W. Wong, "Single Component Critical Flashing Flow," 89CET355 (April 11, 1989).

DEFINITIONS

Many general fluid dynamics terms defined in [Section XIV-A](#) are applicable to two-phase flow. Additional definitions specific to two-phase flows are provided below.

TWO-PHASE FLOW DEFINITIONS

CRITICAL POINT

The location along the flow path that experiences the steepest pressure gradient in critical flow. For a valve or a nozzle, the critical point is located at the narrowest point of the device and for a crack or a short pipe, it is usually at the exit.

FLOW MALDISTRIBUTION

In branching flows, such as tee junctions or perforated pipe distributors, the flow maldistribution measures the preferential distribution of flow among alternative paths. As a percentage, it is calculated as follows.

$$\left(\frac{\text{Largest mass flow in any branch}}{\text{Smallest mass flow in any branch}} - 1 \right) \times 100\%$$

In the preceding relation, the branch flows refer to branches or flow paths downstream of the split and not to the main branch.

QUALITY

Mass fraction of vapor in vapor-liquid flow.

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STAGNATION

Location upstream of the critical point where flow is stagnant or moving very slowly.

VELOCITY SLIP

In actual two-phase flow there is slip between vapor and liquid, with vapor flowing at a higher average velocity.

BASIC DESIGN CONSIDERATIONS

The considerations discussed below affect the basis for the calculation procedures given later in this subsection.

FLOW REGIMES IN HORIZONTAL OR SLIGHTLY INCLINED PIPES

In two-phase flow, interactions between liquid and vapor phases, as influenced by their physical properties and flow rates and by the size, roughness and orientation of the pipe, cause the fluids to flow in various types of patterns. These patterns are called flow regimes. Only one type of flow exists at a given point in a line at any given time. However, as flow conditions change, the flow regime may change from one type to another.

Seven principal flow regimes have been defined to describe flow found in horizontal or slightly inclined pipes. These flow regimes are described below, in order of increasing vapor velocity. In the accompanying sketches, the direction of flow is from left to right.

BUBBLE FLOW - liquid occupies the bulk of the cross-section and vapor flows in the form of bubbles along the top of the pipe. Vapor and liquid velocities are approximately equal. If the bubbles become dispersed throughout the liquid, then this is sometimes called froth flow. In uphill flow bubbles retain their identity over a wider range of conditions. In downhill flow the behavior is displaced in the direction of plug flow.



PLUG FLOW - as the vapor rate increases, the bubbles coalesce, and alternating plugs of vapor and liquid flow along the top of the pipe with liquid remaining the continuous phase along the bottom. In an uphill orientation, the behavior is displaced in the direction of bubble flow; downhill, stratified flow is favored.



STRATIFIED FLOW - as the vapor rate continues to increase, the plugs become a continuous phase. Vapor flows along the top of the pipe and liquid flows along the bottom. The interface between phases is relatively smooth and the fraction occupied by each phase remains constant. In uphill flow, stratified flow rarely occurs with wavy flow being favored. Downhill, stratified flow is somewhat enhanced, as long as the inclination is not too steep.



WAVY FLOW - as the vapor rate increases still further, the vapor moves appreciably faster than the liquid, and the resulting friction at the interface forms liquid waves. The wave amplitude increases with increasing vapor rate. Wavy flow can occur uphill, but over a narrower range of conditions than in a horizontal pipe. Downhill, the waves are milder for a given vapor rate and the transition to slug flow, if it occurs at all, takes place at higher vapor rates than in horizontal pipes.

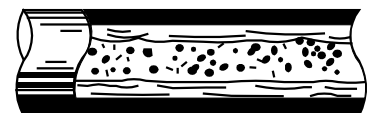


SLUG FLOW - when the vapor rate reaches a certain critical value, the crests of the liquid waves touch the top of the pipe and form frothy slugs. The velocity of these slugs, and that of the alternating vapor slugs, is greater than the average liquid velocity. In the body of a vapor slug the liquid level is depressed so that vapor occupies a large part of the flow area at that point. Uphill, slug flow is initiated at lower vapor rates than in horizontal pipe. Downhill, it takes higher vapor rates to establish slug flow than in horizontal pipe, and the behavior is displaced in the direction of annular flow. **SINCE SLUG FLOW MAY LEAD TO PULSATION AND VIBRATION IN BENDS, VALVES and other flow restrictions, it should be avoided where possible.**



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ANNULAR FLOW - the liquid flows as an annular film of varying thickness along the wall, while the vapor flows as a high-speed core down the middle. There is a great deal of slip between phases. Part of the liquid is sheared off from the film by the vapor and is carried along in the core as entrained droplets. At the same time, turbulent eddies in the vapor deposit droplets on the liquid film. The annular film on the wall is thicker at the bottom of the pipe than at the top, the difference decreasing with distance from slug flow conditions. Downstream of bends, most of the liquid will be at the outer wall.



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In annular flow, the effects of friction pressure drop and momentum outweigh the effect of

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gravity, so that pipe orientation and direction of flow have less influence than in the previous flow regimes. Annular flow is a very stable flow regime. For this reason and because vapor-liquid mass transfer is favored, this flow regime is advantageous for some chemical reactions.

SPRAY FLOW (ALSO KNOWN AS MIST FLOW OR DISPERSED FLOW) - when the vapor velocity in annular flow becomes high enough, all of the liquid film is torn away from the wall and is carried by the vapor as entrained droplets. This flow regime is almost completely independent of pipe orientation or direction of flow.



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FLOW REGIMES IN VERTICAL PIPES

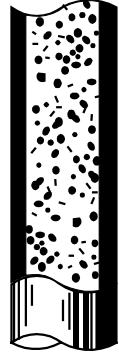
Flow behavior in vertical pipes, where gravity plays an important role, has been less extensively investigated than has flow in horizontal pipes. Most of the available information on vertical flow pertains to upflow.

Conditions under which certain flow regimes exist depend largely on the orientation of the pipe and the direction of flow. In a situation where stratified or wavy flow would exist in a horizontal pipe, tilting the pipe downward increases the relative velocity of the liquid, making a larger part of the flow area available for the vapor. On the other hand, tilting the pipe upward causes the liquid to drain back downhill until enough has accumulated to block off the entire cross-section. The vapor can then no longer get past the liquid, and therefore pushes a slug of liquid through the inclined section of the line.



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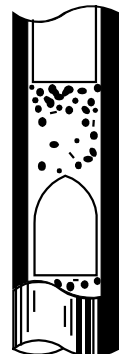
Five principal flow regimes have been defined to describe vertical flow. These flow regimes are described below, in order of increasing vapor velocity. In the accompanying sketches, the direction of flow is upward. **BUBBLE FLOW** - upward flowing liquid is the continuous phase, with dispersed bubbles of vapor rising through it. The velocity of the bubbles exceeds that of the liquid, because of buoyancy. As vapor flow rate is increased, the sizes, number and velocity of the bubbles increase. The bubbles retain their identity, without coalescing into slugs, at higher vapor rates than in a horizontal pipe.



SLUG FLOW - as the vapor rate increases, bubbles coalesce into slugs which occupy the bulk of the cross-sectional area. Alternating slugs of vapor and liquid move up the pipe with some bubbles of vapor entrained in the liquid slugs. Surrounding each vapor slug is a laminar film of liquid which flows toward the bottom of the slug. As the vapor rate is increased, the lengths and velocity of the vapor slugs increase.

Slug flow can occur in the downward direction, but is usually not initiated in that orientation. However, if slug flow is well established in an upward leg of a coil, it will persist in a following downward leg, provided that other conditions remain the same.

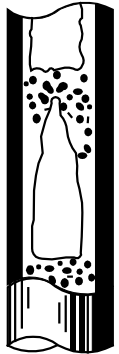
IN DESIGNING FOR TWO-PHASE FLOW IT IS NORMAL PRACTICE TO TRY TO AVOID SLUG FLOW, since this regime can lead to serious pressure fluctuations and vibration, especially at vessel inlets and in bends, valves and other flow restrictions. This could lead to serious equipment deterioration or operating problems. When slug flow cannot be avoided (for instance, in thermosyphon reboilers), one should avoid flow restrictions and use long-radius bends to make turns as smooth as possible.



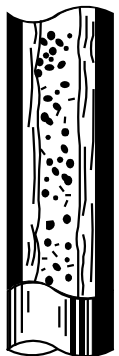
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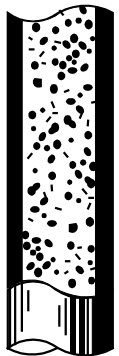
FROTH FLOW - as the vapor rate increases further, the laminar liquid film is destroyed by vapor turbulence and the vapor slugs become more irregular. Mixing of vapor bubbles with the liquid increases and a turbulent, disordered pattern is formed with ever shortening liquid slugs separating successive vapor slugs. The transition to annular flow is the point at which liquid separation between vapor slugs disappears and the vapor slugs coalesce into a continuous, central core of vapor. Since froth flow has much in common with slug flow, the two regimes are often lumped together and called slug flow. In the downward direction, froth flow behaves much the same as slug flow does, except that the former is more easily initiated in this orientation, particularly if conditions are bordering on those for annular flow.



ANNULAR FLOW - this flow regime is similar to annular flow in horizontal pipe, except that the slip between phases is affected by gravity. In upflow, the annular liquid film is slowed down by gravity, which increases the difference in velocities between vapor and liquid. In downflow, the reverse is true, with gravity speeding up the liquid and reducing the difference in velocities between vapor and liquid. On the other hand, the liquid film thickness is more uniform around the circumference of the pipe than in horizontal flow. Annular flow tends to be the dominant regime in vertical downflow.



MIST FLOW - this flow regime is essentially the same as spray flow in horizontal pipe. The very high vapor rates required to completely disperse the liquid essentially eliminate the effects of orientation and direction of flow. In identification of vertical two-phase flow regimes, annular and mist flow are often considered together (and called annular-mist).



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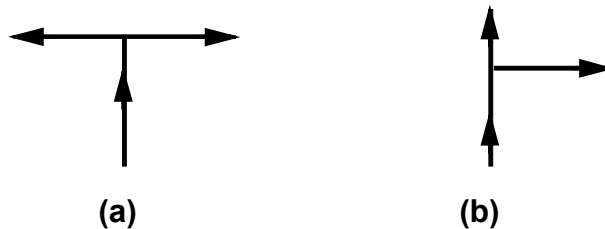
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EFFECT OF FITTINGS ON TWO-PHASE FLOW

Fittings may strongly affect the flow of vapor-liquid mixtures. Bends tend to cause phase segregation. Because of differences in momenta, liquid tends to move to the outer wall while vapor tends to stay close to the inner wall. The segregation can be minimized by the use of blanked-off tees instead of elbows. The fluid should enter through one straight-through end and exit through the branch. Orifices, valves, and other flow restrictions tend to disperse the two phases into each other. For example, if the flow upstream of an orifice is in a separated flow regime such as stratified or slug the flow regime immediately downstream of the orifice will be dispersed flow in most cases. It may take 100 pipe diameters downstream of the orifice for the flow to re-establish the separated flow pattern.

BASIC DESIGN CONSIDERATIONS

It has been shown that the behavior of a two-phase mixture in a tee is highly unpredictable, i.e., Liquid or vapor may prefer one branch over the other resulting in uneven splitting, so that the volume fraction of vapor (or liquid) differs in the two split streams. However, it is generally true that if the flow entering the tee is a dispersed flow the split will be more even than when the entering stream is in a separated flow regime. The splitting of a two-phase stream in a pipe tee is also dependent on the configuration of the flow. In general the splitting is more even if it is done in a more symmetrical fashion. For example, splitting is more even in an impacting tee (a) than in a straight-through tee (b) as shown below.



In order to maintain symmetry, elbows immediately upstream of an impacting tee should be mounted perpendicular to the plane of the tee. If this is not possible a blanked-off tee should be used instead of the elbow.

PRESSURE DROP IN STRAIGHT PIPE

Two methods are described in this subsection for calculating two-phase pressure drop in straight pipe. Method a assumes a homogeneous mixture of vapor and liquid, with no slip between the phases. Method b allows for slip between the two phases, but assumes that the local ratio of liquid velocity to vapor velocity is constant independent of position. Neither method requires the prediction of the two-phase flow regime. Method b is more accurate than method a, especially for vertical upflow but requires a trial-and-error procedure.

OTHER PRESSURE DROP

For two-phase pressure drop through other than straight pipe, the fluid is treated as a single ("liquid") phase with the average properties of the mixture, and the method given in section xiv-b for liquid flow is used. This applies to valves and other fittings; orifices, nozzles and venturis; sudden contractions and expansions; and combining and dividing of streams. An exception is that for orifices, nozzles and venturis: the pressure recovery factor is not used.

PERFORATED-PIPE DISTRIBUTORS

THE DESIGN OF PERFORATED-PIPE DISTRIBUTORS FOR SINGLE-PHASE FLUID AS DOCUMENTED IN SECTION XIV-B may be applied to pseudo-liquid (vapor volume flow rate $\leq 5\%$ of the total flow) or pseudo-vapor (liquid volume flow rate $\leq 5\%$ of the total flow) two-phase mixtures. With two-phase flow there is the potential complication that liquid may flow preferentially through some of the holes, and vapor through the others. For pseudo-liquids and pseudo-vapors, this effect will be small. However, for mixed-phase fluids (vapor volume flow rate $> 5\%$ or liquid volume flow rate $> 5\%$ of the total flow) it may be quite severe. As a result the section xiv-b procedure is not recommended for mixed-phase flow. Currently no general procedures are available for mixed-phase fluids, and the design should be considered on an individual basis. ExxonMobil engineering should be consulted for such a design.

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FLASHING CRITICAL FLOW

When a liquid or a vapor-liquid mixture under pressure is released to the surroundings through a pressure relief valve or a crack in a pipe or vessel, flashing can occur along the flow path due to the reduction in pressure. If the pressure drop is severe enough, a situation can arise in which the release rate becomes independent of the surrounding pressure. In other words, the release rate cannot be increased by further reduction in the surrounding pressure. This condition is called critical flow.

If the two-phase mixture can be considered as a homogeneous mixture, i.e., There is no velocity slip between the vapor and the liquid, and equilibrium between the phases is maintained during the expansion, the critical mass flux and the critical pressure ratio can be calculated when the stagnation conditions, i.e., Pressure, temperature, and quality, are given. The critical pressure ratio is the ratio of the pressure at the critical point to the stagnation pressure.

In reality, phase equilibrium is seldom achieved during the expansion. This is especially true for liquid systems and for two-phase mixtures with low quality (< 0.1). This normally results in a higher critical mass flux than that predicted by assuming phase equilibrium. Methods that include the non-equilibrium effects are available and usually require computer solution. ExxonMobil engineering should be consulted for these more sophisticated calculation methods.

CALCULATION PROCEDURE

The following design methods, equations, and guidelines should be used with the "basic design considerations" given above. The first part gives procedures for determining the flow regime. The succeeding parts deal with methods for calculating pressure drop and liquid holdup in straight pipes, pressure drop in piping components such as valves, orifices and pipe fittings, integrated pressure drop calculation for piping systems, guidelines for designing perforated-pipe distributors and manifolds for distributing two-phase flow to parallel equipment, and finally a procedure for calculating critical flashing flow.

DETERMINING FLOW REGIME

The flow regime is a very important design consideration for two-phase flow since it has significant effects on mechanical design, heat transfer, flow splitting, etc. For example, slug flow can lead to pipe vibration and should be avoided whenever possible. Depending on the flow regime, a two-phase fluid can have quite different heat transfer characteristics. As mentioned in the basic design considerations section, the splitting of a two-phase mixture in a tee is very sensitive to its flow pattern. Knowledge of the flow regime is thus essential to design of equipment involving two-phase flow.

Flow regime maps for the most common and significant pipe orientations: horizontal and vertical upflow are given in [Figures 1](#) and [2](#), respectively, as functions of the flow rates and system properties. It should be noted that these maps predict flow regimes for fully-developed flow, which may require up to 100 pipe diameters downstream of an obstruction or fitting to establish. Further, these maps are mainly based on air-water data obtained from small diameter pipes [typically 1/2 in. (13 mm) to 3 in. (76 mm) dia.]. Therefore, the maps are more reliable for systems with properties similar to the air-water system flowing in pipes not much larger than 3 in. (76 mm) dia. For systems with vastly different properties than air-water or for flow in large pipes (> 3 in. (76 mm) dia.), The maps should be considered only as qualitative guides to the flow pattern. Also, when using a flow map it must be recognized that the boundaries between flow regimes do not mean that there is a sharp transition from one regime to the other.



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HORIZONTAL FLOW - FIGURE 1 is a flow map for horizontal flow. Caution should be used in applying it to slightly inclined upflow or downflow since the effect of gravity on the two-phase mixture can be quite strong. For example, stratified flow is favored in inclined pipes with downflow. For flow regime estimation in inclined pipes contact ExxonMobil engineering.

THE FOLLOWING PROCEDURE IS RECOMMENDED FOR DETERMINING THE FLOW REGIME WHEN THE PIPE SIZE AND FLOW CONDITIONS ARE GIVEN:

1. Calculate the parameter X in [Figure 1](#) from the following equation:

$$X = F_1 \left(\frac{1-x}{x} \right) \left[\frac{\rho_G^{1/2} \mu_L^{1/3}}{\rho_L^{1/6} \sigma} \right] \quad \text{Eq. (1)}$$

where:

- X = Abscissa of [Figure 1](#)
- F_1 = 530.7 (Customary)
= 2100 (Metric)
- ρ_G = Vapor density, lb_m/ft³ (kg/m³)
- ρ_L = Liquid density, lb_m/ft³ (kg/m³)
- σ = Liquid surface tension, dynes/cm (mN/m)
- x = Mass fraction vapor (quality), dimensionless
- μ_L = Liquid viscosity, cp (Pa•s)

2. Calculate the parameter Y in [Figure 1](#) from the following equation:

$$Y = F_2 \frac{xW}{d^2 \rho_G^{1/2} \rho_L^{1/2}} \quad \text{Eq. (2)}$$

where:

- Y = Ordinate of [Figure 1](#)
- F_2 = 396 (Customary)
= 3.24 x 10⁷ (Metric)
- W = Total mass flow rate, klb_m/h (kg/s)
- d = Inside diameter of pipe, in. (mm)

3. Using X and Y, determine the flow regime from [Figure 1](#).

WHEN A PARTICULAR FLOW REGIME IS DESIRED, THE FOLLOWING PROCEDURE IS RECOMMENDED FOR CALCULATING THE PIPE SIZE TO ACHIEVE THE FLOW PATTERN:

1. Calculate the parameter X from [Eq. \(1\)](#).
2. Using this value of X and [Figure 1](#) pick a value of Y in the desired flow regime. Keep in mind that the boundaries between flow regimes are not sharp as indicated in the map, so that it is prudent to pick a value of Y away from the boundaries.

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3. Use Y to calculate the pipe diameter d from the following equation:

$$d = F_3 \left(\frac{1}{\rho_G^{1/4} \rho_L^{1/4}} \right) \left(\frac{xW}{Y} \right)^{1/2} \quad \text{Eq. (3)}$$

where:

$$\begin{aligned} F_3 &= 19.9 && \text{(Customary)} \\ &= 5695 && \text{(Metric)} \end{aligned}$$

VERTICAL FLOW - Figure 2 is a flow map for vertical upflow. Maps for vertical downflow are not usually needed since in most cases such flows are in the annular/mist flow regime. As a result a vertical downflow map is not given. ExxonMobil engineering should be contacted if such a map is needed.

When the pipe size and flow conditions are given the flow regime can be found by the following procedure:

1. Calculate the parameter Ψ from the following equation:

$$\Psi = F_4 \left(\frac{\rho_L}{\sigma} \right)^{1/4} \quad \text{Eq. (4)}$$

where:

$$\begin{aligned} F_4 &= 1.04 && \text{(Customary)} \\ &= 0.52 && \text{(Metric)} \end{aligned}$$

2. Calculate V_{sl} from the following equation:

$$V_{sl} = F_5 \frac{(1-x)W}{\rho_L d^2} \quad \text{Eq. (5)}$$

where:

$$\begin{aligned} V_{sl} &= \text{Superficial velocity of liquid, ft/s (m/s)} \\ F_5 &= 50.93 && \text{(Customary)} \\ &= 1.273 \times 10^6 && \text{(Metric)} \end{aligned}$$

3. Calculate the parameter ϕ from the following equation:

$$\phi = F_6 \left(\frac{\rho_L}{\sigma} \right)^{1/4} (\rho_G)^{1/3} \quad \text{Eq. (6)}$$

where:

$$\begin{aligned} F_6 &= 2.45 && \text{(Customary)} \\ &= 0.486 && \text{(Metric)} \end{aligned}$$



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4. Calculate V_{sg} from the following equation:

$$V_{sg} = F_5 \frac{xW}{\rho_G d^2} \quad \text{Eq. (7)}$$

where:

$$\begin{aligned} V_{sg} &= \text{Superficial velocity of vapor, ft/s (m/s)} \\ F_5 &= 50.93 \quad (\text{Customary}) \\ &= 1.273 \times 10^6 \quad (\text{Metric}) \end{aligned}$$

5. Calculate ΨV_{sl} and ϕV_{sg} and determine the flow regime from [Figure 2](#) (A for customary units and B for metric units). For calculating the pipe size which will give a desired flow regime in vertical upflow, the following procedure is recommended:

1. Calculate the parameter ϕ from [Eq. \(6\)](#).
2. Pick a value of ϕV_{sg} from [Figure 2](#) that will give the desired flow regime. For example pick $\phi V_{sg} > 70$ ft/s (21.3 m/s) for an annular flow regime and $\phi V_{sg} < 0.4$ ft/s (0.122 m/s) for a bubble flow regime.
3. Calculate V_{sg} based on the value of ϕV_{sg} picked in Step 2 and ϕ calculated in Step 1.
4. Calculate the pipe diameter from the following equation:

$$d = 12 \left(\frac{0.354 \times W}{\rho_G V_{sg}} \right)^{1/2} \quad (\text{Customary}) \quad \text{Eq. (8a)}$$

$$d = 1000 \left(\frac{1.273 \times W}{\rho_G V_{sg}} \right)^{1/2} \quad (\text{Metric}) \quad \text{Eq. (8b)}$$

5. Calculate V_{sl} from [Eq. \(5\)](#) using the d calculated in Step 4.
6. Calculate ΨV_{sl} where Ψ is calculated from [Eq. \(4\)](#).
7. Determine the flow regime based on ΨV_{sl} from Step 6 and ϕV_{sg} from Step 2 using [Figure 2](#) and check whether it is the desired one. If not, repeat Steps 2 to 7 until the desired flow regime is obtained.

PRESSURE DROP IN STRAIGHT PIPE

The methods given below for calculating pressure drop in circular pipes are independent of the flow regime. For non-circular conduits, these methods may be used with the equivalent hydraulic diameter d_{eq} defined by the following relation. It should be pointed out that the equivalent hydraulic diameter is a measure of the size of the cross sectional area available for flow, but cannot be used in place of the pipe diameter for flow rate calculations (see the note following the definition in subsection a).

$$d_{eq} = 4 \left(\frac{\text{Cross-section Area}}{\text{Conduit Perimeter}} \right) \quad \text{Eq. (9)}$$

The modified homogeneous method (method a) is the simplest since it assumes no velocity slip between the vapor and the liquid and is independent of liquid holdup. Both methods a and b are included in the two-phase pressure drop section of pegasys (pc engineering application system). If liquid holdup is needed method b, which is based on dukler's similarity method, should be used. Method b is slightly more accurate than method a when applied to horizontal flow **and may be expected to be accurate to $\pm 30\%$** . For vertical flow in which the pressure drop due to gravity may dominate, method b is appreciably more reliable. This is because the two-phase density calculated by method b is based on liquid holdup and as a result is more accurate. For more SOPHISTICATED METHODS SUCH AS THOSE DEPENDENT ON FLOW REGIME AND METHODS FOR INCLINED PIPES, EXXONMOBIL ENGINEERING SHOULD BE CONSULTED.

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METHOD A - MODIFIED HOMOGENEOUS METHOD

1. Assume the downstream pressure if upstream pressure is given or vice versa and calculate an average line pressure.
2. Calculate the two-phase mixture density, ρ_{ns} , based on homogeneous flow assumption:

$$\rho_{ns} = \rho_L \lambda + \rho_G (1 - \lambda) \quad \text{Eq. (10)}$$

and

$$\lambda = \frac{Q_L}{Q_L + Q_G} \quad \text{Eq. (11)}$$

where:

ρ_{ns} = Two-phase, no slip density, lb/ft³ (kg/m³)

λ = Volume fraction liquid, dimensionless

Q_L = Liquid volumetric flow rate, ft³/s (m³/s)

Q_G = Gas volumetric flow rate, ft³/s (m³/s)

3. Assume that the average mixture viscosity, $\mu_{2\phi}$, is equal to the liquid viscosity:

$$\mu_{2\phi} = \mu_L \quad \text{Eq. (12)}$$

where:

$\mu_{2\phi}$ = Two-phase viscosity, cP (Pa•s)

4. Calculate the average mixture velocity, $V_{2\phi}$:

$$V_{2\phi} = F_7 \frac{(Q_L + Q_G)}{d^2} \quad \text{Eq. (13)}$$

where:

$V_{2\phi}$ = Average mixture velocity, ft/s

F_7 = 184 (Customary)

= 1.273 x 10⁶ (Metric)

5. Calculate the two-phase Reynolds Number, assuming homogeneous flow (no slip):

$$\text{Re}_{ns} = F_8 \frac{d V_{2\phi} \rho_{ns}}{\mu_{2\phi}} \quad \text{Eq. (14)}$$

where:

Re_{ns} = No-slip Reynolds Number, dimensionless

F_8 = 124 (Customary)

= 10⁻³ (Metric)

6. Find the Fanning friction factor, f , from [Figure 1](#) or [2](#) in Subsection B, using Re_{ns} calculated from [Eq. \(14\)](#):



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7. Calculate the pressure drop due to friction $(\Delta P)_f$:

$$(\Delta P)_f = \frac{f V_{2\phi}^2 \rho_{ns} L}{6 g_c d} \quad (\text{Customary}) \quad \text{Eq. (15a)}$$

$$(\Delta P)_f = \frac{2 f V_{2\phi}^2 \rho_{ns} L}{d} \quad (\text{Metric}) \quad \text{Eq. (15b)}$$

where:

L = Pipe length, ft (m)

$(\Delta P)_f$ = Frictional pressure drop, psi (kPa)

g_c = Gravitational conversion factor, 32.174 ft lb_m/lb_f s²

8. For calculating the pressure drop due to elevation changes, one assumes no pressure recovery in the downhill legs. When the pressure gain on a negative leg must be known, consult ExxonMobil Engineering.

- a. Calculate the superficial vapor velocity, V_{sg} . This is the velocity the vapor would have if there were no liquid in the line.

$$V_{sg} = F_7 \frac{Q_G}{d^2} \quad \text{Eq. (16)}$$

where:

V_{sg} = Superficial vapor velocity, ft/s (m/s)

F_7 = 184 (Customary)

= 1.273 x 10⁶ (Metric)

- b. Calculate the pressure drop due to elevation changes $(\Delta P)_e$:

$$(\Delta P)_e = \frac{E_H \rho_L \Sigma H}{144} \quad (\text{Customary}) \quad \text{Eq. (17a)}$$

$$(\Delta P)_e = 9.8 \times 10^{-3} E_H \rho_L \Sigma H \quad (\text{Metric}) \quad \text{Eq. (17b)}$$

where:

$(\Delta P)_e$ = Pressure drop due to elevation change, psi (kPa)

E_H = Elevation head factor from [Figure 3](#), using V_{sg} calculated from [Eq. \(16\)](#)

ΣH = Summation of all elevation uplegs measured in the vertical direction, ft. Note: This is not the same as the net elevation change between pipe inlet and outlet.



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9. Pressure drop due to acceleration is usually small and can normally be neglected. However, it should be checked, as indicated below:

a. Calculate the acceleration group, J :

$$J = \frac{18 (W_L + W_G) W_G \bar{P}}{g_c d^4 P_1 P_2 \bar{\rho}_G} \quad \text{(Customary)} \quad \text{Eq. (18a)}$$

$$J = \frac{162 \times 10^9 (W_L + W_G) W_G \bar{P}}{d^4 P_1 P_2 \bar{\rho}_G} \quad \text{(Metric)} \quad \text{Eq. (18b)}$$

where:

- J = Acceleration group, dimensionless
 W_L = Liquid mass flow rate, klb_m/h (kg/s)
 W_G = Vapor mass flow rate, klb_m/h (kg/s)
 g_c = Dimensional constant, 32.174 ft lb_m/lb_f s²
 d = Inside diameter of pipe, in. (mm)
 P_1 = Upstream pressure, psia (kPa)
 P_2 = Downstream pressure, psia (kPa)
 $\bar{P}$ = Average pressure, $\frac{P_1 + P_2}{2}$, psia (kPa)
 $\bar{\rho}_G$ = Average vapor density, $\frac{\rho_{G1} + \rho_{G2}}{2}$, lb_m/ft³ (kg/m³)

- b. If $J \leq 0.1$, acceleration may be neglected, and the total pressure drop can be calculated from [Eq. \(19\)](#). See Step 10a. If $J > 0.1$, a trial-and-error procedure is required (Step 10b).

10. Calculate the total pressure drop:

a. If the acceleration as determined in Step 9 is negligible, calculate the total pressure drop $(\Delta P)_t$ from [Eq. \(19\)](#):

$$(\Delta P)_t = (\Delta P)_f + (\Delta P)_e \quad \text{Eq. (19)}$$

b. If the acceleration is not negligible, use the values $(\Delta P)_f$ and $(\Delta P)_e$ from [Eqs. \(15\) and \(17\)](#) in [Eq. \(20\)](#):

$$(\Delta P)_t = \frac{(\Delta P)_f + (\Delta P)_e}{1 - J} \quad \text{Eq. (20)}$$

where:

- $(\Delta P)_t$ = Total pressure drop due to friction, elevation, and acceleration, psi (kPa)

11. Check the assumed average pressure (Step 1) and repeat the procedure if more accuracy is required, particularly when acceleration (Step 10b) must be accounted for:



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METHOD B - SIMILARITY METHOD (LIQUID HOLDUP INCLUDED)

Use the following procedure for calculating pressure drop and liquid holdup in **HORIZONTAL FLOW** when flow rate and either upstream or downstream pressure are given.

1. Assume the downstream pressure if the upstream pressure is given or vice versa and calculate the average line pressure.
2. Calculate λ , the volume fraction liquid, from [Eq. \(11\)](#).
3. Calculate the mixture viscosity, $\mu_{2\phi} = \mu_L \lambda + \mu_G (1 - \lambda)$. Eq. (21)
4. Calculate the average mixture velocity, $V_{2\phi}$ from [Eq. \(13\)](#).
5. Calculate the two-phase Reynolds Number, $Re_{2\phi}$. This is a trial-and-error procedure consisting of the following steps:
 - a. Estimate a value of R_L , the liquid holdup. (Use λ and [Figure 4](#) for estimate.)
 - b. Calculate $Re_{2\phi}$ using λ , $\mu_{2\phi}$, and $V_{2\phi}$ from:

$$Re_{2\phi} = F_8 \frac{d V_{2\phi} \rho_{2\phi}}{\mu_{2\phi}} \quad \text{Eq. (22)}$$

$$\rho_{2\phi} = \frac{\rho_L \lambda^2}{R_L} + \frac{\rho_G (1 - \lambda)^2}{1 - R_L} \quad \text{Eq. (23)}$$

where:

$Re_{2\phi}$ = Two-phase Reynolds Number, dimensionless

$$F_8 = \begin{matrix} 124 & \text{(Customary)} \\ 10^{-3} & \text{(Metric)} \end{matrix}$$

$\rho_{2\phi}$ = Two-phase density, lb_m/ft³ (kg/m³)

- c. Use [Figure 4](#) with λ and $Re_{2\phi}$ to obtain a new value of R_L . If the assumed and calculated values of R_L agree within 5%, the accuracy is sufficient. If not, Step 5b must be repeated with the new value of R_L .
 - d. When the assumed and calculated values of R_L agree within 5%, use the last calculated value of R_L to calculate $Re_{2\phi}$ from [Eq. \(22\)](#).
6. Calculate $f_{1\phi}$, the single-phase Fanning friction factor from [Eq. \(24\)](#):

$$f_{1\phi} = 0.0014 + \frac{0.125}{(Re_{2\phi})^{0.32}} \quad \text{Eq. (24)}$$

7. Look up $f_{2\phi}/f_{1\phi}$ from [Figure 5](#) and calculate $f_{2\phi}$ from this ratio and the value of $f_{1\phi}$ calculated in Step 6.
8. Calculate pressure drop due to friction, $(\Delta P)_f$, from [Eq. \(25\)](#):

$$(\Delta P)_f = \frac{f_{2\phi} V_{2\phi}^2 \rho_{2\phi} L}{6 g_c d} \quad \text{(Customary)} \quad \text{Eq. (25a)}$$

$$(\Delta P)_f = \frac{2 f_{2\phi} V_{2\phi}^2 \rho_{2\phi} L}{d} \quad \text{(Metric)} \quad \text{Eq. (25b)}$$

