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## Basics of Fluid Flow

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Single and Two phase fluid flow calculations

*Prepared by*

Yasser Kassem
Preface

I would like to present a simple calculation guide for friction losses, and other fluid calculation for both single liquid phase, single gas phase, and two-phase flow.

Best Regards

Yasser Kassem
Jan. 2nd, 2020

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Chapter One

Fluid flow and pressure drop

1. Introduction to facility piping and pipeline systems

The produced Oil and gas fluid produced must be transported to a facility where it is separated into oil, water, and gas; treated to remove impurities such as H₂S, CO₂, H₂O, and solids; processed into specific end products and refined or stored for eventual sales. Figure 1 is a simplified block diagram that illustrates the basic "wellhead to sales" concept. The diagram begins with wellhead choke, which is used to control the rate of flow from each well. The fluid from the well travels through a flow line to the production facility where the fluid is separated, conditioned, treated, processed, measured, and refined or stored.

The facility piping and pipeline systems associated with producing wells include, but are not limited to, the well flow line, trunk line, facility (on-plot) interconnecting equipment piping within the production facility, gathering or sales pipelines, and transmission pipelines. A brief description of the aforementioned facility piping and pipeline systems follows.

1.1.1 Flow line
A well flow line identifies a two-phase line from a wellhead to a production manifold. Flow lines range in size from 2 in. to 20 in.

1.1.2 Trunk line
A trunk line is a larger line that connects two or more well flow lines that carries the combined well streams to the production manifold. Trunk lines range from 10 in. to 42 in.

1.1.3 Manifold
A manifold is a combination of pipes, fittings, and valves used to combine production from several sources and direct the combined flow into appropriate production equipment. A manifold may also originate from a single inlet stream and divide the stream into multiple outlet streams. Manifolds are generally located where many flow lines come together, such as gathering stations, tank batteries, metering sites, separation stations, and offshore platforms. Manifolds also are used in gas lift injection systems, gas/water injection systems, pump/compressor stations, gas plants, and installations where fluids are distributed to multiple units. A production manifold accepts the flow streams from well flow lines and directs the combined flow to either test, or production separators and tanks.

1.1.4 Facility (on-plot) interconnecting piping
Facility piping consists of piping within a well-defined boundary of processing plants, piping compressor stations, or pumping stations. The piping is used for conducting a variety of fluids within those boundaries as required.

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1.1.5 Gathering line
A gathering line consists of the line downstream of field manifolds or separators containing fluid flow from multiple wells and leading to the production facility. The gathering line may handle condensed hydrocarbon liquids, water, and corrosive gas and may require special design considerations.

1.1.6 Transmission line
A transmission line consists of a cross-country piping system for transporting gas or liquids. The inlet is normally the custody transfer point or the production facility boundary with the outlet at its final destination, for example, processing plants and refineries. Transmission lines are usually long and have large diameters.

1.2 Introduction to fluid flow design
When designing facility piping and/or pipeline systems, it is essential to optimize the line size and determine pump and/or compressor requirements. Several factors that should be considered when determining the size of a line to meet the design requirements are the following:

1. Volume of fluid
2. Distance
3. Pressure loss

1.2.1 Volume of fluid
The main consideration in line sizing is the volume of fluid that must be transported through the piping system. The exact volume is rarely known during the initial design stage. An estimate is normally made for initial design purposes. Excess capacity reduces line profitability, while too small a line might need to be expanded in the future.

1.2.2 Distance
For pipelines, the distance between the entry point and the delivery point must be known. The designer needs to know the type of terrain the pipeline must traverse and the elevation profile along the right-of-way as it affects pressure loss and power requirements. The designer must also
be knowledgeable of environmental conditions, ecological, historical, and archaeological sites as they might impact the pipeline routing, thereby increasing the length of the pipeline.

1.2.3 Pressure loss
The pressure loss as the fluid flows through the piping system is a key factor in both facility and pipeline design. Available piping inlet pressure must be known, as well as if there is any particular outlet requirement at the delivery point.
The design process begins with sizing lines for a given fluid flow rate. One must conform to the following:
1. Applicable codes, standards, and recommended practices
2. Company design criteria as contained in applicable facility specifications
3. Local regulatory requirements
4. The design process requires determination of the line size and wall thickness.

1.2.4 Line size determination
1.2.4.1 Pressure drop considerations
Pressure drop is used to avoid the installation of excessive brake horsepower required to boost the pressure for transporting the fluid. It is used to conform to the available piping inlet and discharge pressures and the allowable pressure gradient standards.

1.2.4.2 Fluid velocity considerations
Fluid velocities are used to prevent excessive water hammer, erosive velocities, and liquids and/or solids from dropping out of the flow stream. (Refer to 1.5.3) “Velocity of fluids (Liquid and gas).

1.2.5 Wall thickness determination
1.2.5.1 Maximum internal/external pressure considerations
Internal burst pressure is based on initial well conditions and other flow line considerations. The external collapse resistance is a consideration in offshore locations and in onshore locations where the overburden loads are large.
The applicable design standards are ASME B31.3, ASME B31.4, and ASME B31.8.

1.3 Fluid flow principles
1.3.1 Pressure changes
As a fluid flows through a pipe, its pressure changes. Calculation of these pressure changes is necessary to size pipe. Fluid pressure changes can occur due to the following:
1. Acceleration effects
2. Elevation effects
3. Frictional effects

1.3.1.1 Acceleration effects
Acceleration effects in production facility piping systems are generally negligible and are ignored.
1.3.1.2 Elevation effects
Elevation effects are a result of hydrostatic gravity effects and can occur even in still fluid in inclined pipes. Elevation effects are important in wellbore pressure gradients, cross-country pipelines, and subsea pipelines. They are less important in production facilities and process piping, except pump suction lines and flow lines.

1.3.1.3 Frictional effects
Frictional effects, or pressure drop, are of primary importance in production facilities, flow lines, and pipeline design. As a fluid travels down a pipe, flow is retarded by frictional shear stresses with the pipe walls. The pressure levels decrease downstream as energy is used to overcome the frictional effects. The only exception occurs in downwardly inclined sections of pipe where elevation effects may overcome the pressure-decreasing effects of friction. The faster the fluid travels in the pipe, the greater the frictional stresses and the greater the pressure gradient.

1.3.2 Steady-state conditions
Most piping design is performed assuming non-fluctuating flow conditions. Two situations in which transients must be taken into account involve “water hammer” in liquid lines and “line pack and draft” in gas lines. Considerations of the above conditions are necessary primarily for pipeline operations.

1.4 Fluid types
Facility piping and pipelines transport various types of fluids. These include the following:

1.4.1 Gases
- Unprocessed natural gas (rich gas) consists primarily of methane with some heavier hydrocarbons.
- Processed natural gas (lean gas) consists primarily of methane, although small amounts of heavier fractions may still be present.
- Nonhydrocarbon components consist of nitrogen and hydrogen sulfide and carbon dioxide may also be present.
- Natural gas liquids (NGLs) consist primarily of the intermediate-molecular-weight hydrocarbon components such as propane, butane, and pentanes plus.

1.4.2 Crude oil
Crude oil consists of the heavier hydrocarbon fractions that are generally liquid at atmospheric conditions in storage tanks. Volatile oils are stabilized to prevent excessive vapor formation or “weathering” in storage or transport tankers. Piped fluid will remain liquid due to adequate operating pressure.

1.4.3 Water
Produced well streams frequently contain dissolved salts and minerals that are usually corrosive.

1.4.4 Two-phase fluids
Two-phase fluids usually consist of natural gas and condensate or crude oil and associated gas. Flow lines from the well to the production facility are designed for two-phase flow.
1.4.5 Combinations
Produced well fluids contain the following:
1- Hydrocarbons (gases and liquids)
2- Water
3- Varying amounts of CO₂ and H₂S

Liquid hydrocarbons and some water combine both physically and mechanically to form an emulsion that has a higher viscosity value than that of oil or water.

1.5 Fluid characteristics
1.5.1 Physical properties
The physical properties of the transported fluid play an important role in determining the pipe diameter and selecting the pipe material and the associated equipment. They are also important in determining the power required to transport the fluid. The most important fluid properties that affect piping and pipeline design are the following:
1- Composition
2- Density
3- Viscosity
4- Vapor pressure
5- Water content
6- CO₂ and H₂S content
7- Compressibility

1.5.1.1 Composition
Well stream compositions are usually stated as mole fractions. Knowledge of composition is necessary to predict fluid properties such as density, viscosity, and phase behavior. If a compositional analysis is not available, one must rely on a "black oil" characterization in which API gravity, gas gravity, gas-oil ratio, and water-liquid ratio are given. The use of empirical black oil property correlations provides reasonable values for density, viscosity, and phase behavior.

1.5.1.2 Liquid Density
There are several definitions of fluid density that are used in upstream oil and gas operations, such as density, specific gravity or relative density, and API gravity. The density of a fluid is defined as mass per unit volume with unit lbm/ft.³ (kg/m³). Density is a thermodynamic property and is a function of pressure, temperature, and composition. Liquid densities are higher than gas densities and are affected less by pressure and temperature. Gas densities are increased by increasing pressure and decreased by increasing temperature. The density of a fluid is an important property in calculating the elevation pressure drop since elevation pressure drop is the product of density and elevation change. A liquid’s density is often specified by giving its specific gravity relative to water at standard conditions of 60 °F and 14.7 psia (15.6 °C and 101.4 kPa). Thus,

\[ \rho = 62.4 \text{ (SG)} \]  Eq. 1.1

where
\( \rho = \) density of liquid (lbm/ft.³),
SG = specific gravity of liquid relative to water.

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API gravity is a special function of relative density. It is a reverse graduation scale of relative density, where lighter fluids have higher API gravities. For example, a light oil would typically have an API gravity between 30 and 40, while water would have an API gravity of 10. API gravity is defined as:

\[
0^{\text{API}} = \frac{141.5}{\text{Sp.Gr} @ 60 \, \text{Degrees} \, F} - 131.5 \tag{Eq. 1.2}
\]

The density of a mixture of oil and water can be determined by the volume weighted average of the two densities and is given by:

\[
\rho = \left( \frac{\rho_w Q_w + \rho_o Q_o}{Q_T} \right) \tag{Eq. 1.3}
\]

where
- \( \rho = \) density of liquid (lb/ft.\(^3\)),
- \( \rho_o = \) density of oil (lb/ft.\(^3\)),
- \( \rho_w = \) density of water (lb/ft.\(^3\)),
- \( Q_w = \) water flow rate, barrel per day, (BPD),
- \( Q_o = \) oil for rate (BPD),
- \( Q_T = \) total liquid flow rate (BPD).

The specific gravity or relative density of a liquid is indicated relative to water, and that of a gas is indicated relative to air. The specific gravity is measured at certain pressure and temperature conditions. Usually “Standard” conditions which are taken as 60 °F (15.6 °C) and 14.7 psi (1.01325 bar)

The average specific gravity of some oil field liquids are the following:

<table>
<thead>
<tr>
<th>Fluid</th>
<th>Specific Gravity</th>
</tr>
</thead>
<tbody>
<tr>
<td>Crude oil</td>
<td>0.825</td>
</tr>
<tr>
<td>Condensate</td>
<td>0.75</td>
</tr>
<tr>
<td>Butane</td>
<td>0.58</td>
</tr>
</tbody>
</table>

Table 1-1: Specific gravity of some oil fluids.

The specific gravity of an oil and water mixture can be calculated by

\[
(SG)_m = \left( \frac{(SG)_w Q_w + (SG)_o Q_o}{Q_T} \right) / Q_T \tag{Eq. 1.4}
\]

where
- \( (SG)_m = \) specific gravity of liquid,
- \( (SG)_o = \) specific gravity of oil,
- \( (SG)_w = \) specific gravity of water,
- \( Q_w = \) water flow rate, barrel per day, (BPD),
- \( Q_o = \) oil for rate (BPD),
- \( Q_T = \) total liquid flow rate (BPD).

1.5.2: Hydrocarbon gas physical properties
Most of compounds in crude oil and natural gas consist of molecules made up of hydrogen and carbon, therefore these types of compounds are called hydrocarbon.
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The smallest hydrocarbon molecule is Methane (CH$_4$) which consists of one atom of Carbon and four atoms of hydrogen. It may be abbreviated as C$_1$ since it consisted from only one carbon atom. Next compound is Ethane (C$_2$H$_6$) abbreviated as C$_2$, and so on Propane (C$_3$H$_8$), Butane (C$_4$H$_{10}$)...etc.

Hydrocarbon gases are C$_1$–C$_4$), with the increase of carbon number, liquid volatile hydrocarbon is found (e.g. Pentane C$_5$ is the first liquid hydrocarbon at standard conditions).

1.5.2.1: Molecular weight and apparent molecular weight

The molecular weight of a compound is the sum of the atomic weight of the various atoms making up that compound. The Mole is the unit of measurements for the amount of substance, the number of moles is defined as follows:

\[
\text{Mole} = \frac{\text{Weight}}{\text{Molecular weight}} \quad \text{Eq. 1-5}
\]

Expressed as

\[
\text{n} = \frac{m}{M} \quad \text{Eq. 1-6}
\]

or, in units as

\[
\text{lb-mole} = \frac{\text{lb}}{\text{lb/lb-mole}} \quad \text{Eq. 1-7}
\]

<table>
<thead>
<tr>
<th>Compound</th>
<th>Formula</th>
<th>Molecular weight</th>
<th>Boiling Point °F @ 14.7 psia</th>
<th>Relative Density of gas (air=1)</th>
<th>Critical Temp. °R</th>
<th>Critical pressure. Psia</th>
</tr>
</thead>
<tbody>
<tr>
<td>Methane</td>
<td>CH$_4$</td>
<td>16.043</td>
<td>-259</td>
<td>0.5539</td>
<td>343.0</td>
<td>666</td>
</tr>
<tr>
<td>Ethane</td>
<td>C$_2$H$_6$</td>
<td>30.070</td>
<td>-128</td>
<td>1.0382</td>
<td>549.6</td>
<td>706.6</td>
</tr>
<tr>
<td>Propane</td>
<td>C$_3$H$_8$</td>
<td>44.097</td>
<td>-44</td>
<td>1.5225</td>
<td>665.6</td>
<td>615.5</td>
</tr>
<tr>
<td>n-Butane</td>
<td>C$<em>4$H$</em>{10}$</td>
<td>58.124</td>
<td>10.8</td>
<td>2.0068</td>
<td>734.5</td>
<td>527.9</td>
</tr>
<tr>
<td>i-Butane</td>
<td>C$<em>4$H$</em>{10}$</td>
<td>58.124</td>
<td>31.1</td>
<td>2.0068</td>
<td>765.2</td>
<td>550.9</td>
</tr>
<tr>
<td>n-Pentane</td>
<td>C$<em>5$H$</em>{12}$</td>
<td>72.151</td>
<td>82.1</td>
<td>2.4911</td>
<td>829.1</td>
<td>490.4</td>
</tr>
<tr>
<td>i-Pentane</td>
<td>C$<em>5$H$</em>{12}$</td>
<td>72.151</td>
<td>97</td>
<td>2.4911</td>
<td>845.5</td>
<td>488.8</td>
</tr>
<tr>
<td>n-Hexane</td>
<td>C$<em>6$H$</em>{14}$</td>
<td>86.178</td>
<td>156</td>
<td>2.9755</td>
<td>895.5</td>
<td>436.6</td>
</tr>
<tr>
<td>n-Heptane</td>
<td>C$<em>7$H$</em>{16}$</td>
<td>100.205</td>
<td>209</td>
<td>3.4598</td>
<td>972.6</td>
<td>396.8</td>
</tr>
<tr>
<td>n-Octane</td>
<td>C$<em>8$H$</em>{18}$</td>
<td>114.232</td>
<td>258</td>
<td>3.9441</td>
<td>1023.9</td>
<td>360.7</td>
</tr>
<tr>
<td>n-Nonane</td>
<td>C$<em>9$H$</em>{20}$</td>
<td>128.259</td>
<td>303</td>
<td>4.4284</td>
<td>1070.5</td>
<td>330.7</td>
</tr>
<tr>
<td>Carbon dioxide</td>
<td>CO$_2$</td>
<td>44.01</td>
<td>-109.1</td>
<td>1.5197</td>
<td>547.4</td>
<td>1070.0</td>
</tr>
<tr>
<td>Hydrogen sulfide</td>
<td>H$_2$S</td>
<td>34.082</td>
<td>-67.5</td>
<td>1.1769</td>
<td>772.5</td>
<td>1306.5</td>
</tr>
<tr>
<td>Oxygen</td>
<td>O$_2$</td>
<td>32</td>
<td>-297</td>
<td>1.1050</td>
<td>278.2</td>
<td>731.4</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>N$_2$</td>
<td>28.01</td>
<td>-320.4</td>
<td>0.9674</td>
<td>227.1</td>
<td>492.5</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>H$_2$</td>
<td>2.0159</td>
<td>-423</td>
<td>0.0696</td>
<td>59.8</td>
<td>190.7</td>
</tr>
<tr>
<td>Air</td>
<td></td>
<td>28.96</td>
<td>-317.6</td>
<td>1.0000</td>
<td>238.7</td>
<td>551.9</td>
</tr>
<tr>
<td>Water</td>
<td>H$_2$O</td>
<td>18.015</td>
<td>211.95</td>
<td>0.6221</td>
<td>1164.8</td>
<td>3200.1</td>
</tr>
</tbody>
</table>

Table 1-2 Physical constants of light hydrocarbons and some inorganic gases. Adapted from GPSA, Engineering Hand Book.

Example 1.1:

Methane molecule consists of one carbon atom with atomic weight = 12 and 4 hydrogen atoms with atomic weight = 1 each. Molecular weight for Methane (CH$_4$) = (1 × 12) + (4 × 1) = 16 lb/lb-
mole. Similarly, Ethane (C\textsubscript{2}H\textsubscript{6}) molecular weight = (2 \times 12) + (6 \times 1) = 30 lb/lb-mole. Hydrocarbon up to four carbon atoms are gases at room temperature and atmospheric pressure. Reducing the gas temperature and/or increasing the pressure will condense the hydrocarbon gas to a liquid phase. By the increase of carbon atoms in hydrocarbon molecules, consequently the increase in molecular weight, the boiling point increases and a solid hydrocarbon is found at high molecular weight.

Physical constants of light hydrocarbon and some inorganic gases are listed in Table 1-2.

1.5.2.2: Apparent molecular weight of gas mixture
For compounds, the term molecular weight is used, while, for hydrocarbon mixture the term apparent molecular weight is commonly used. Apparent molecular weight is defined as the sum of the products of the mole fractions of each component times the molecular weight of that component. As shown in Eq. 1-8

\[ MW = \sum Y_i (MW)_i \]  
Eq. 1-8

where
\[ y_i = \text{molecular fraction of } i^\text{th} \text{ component}, \]
\[ MW_i = \text{molecular weight of } i^\text{th} \text{ component}, \Sigma y_i = 1. \]

Example 1.2:
Determine the apparent molecular weight for the gas mixture in Table 1-3:

<table>
<thead>
<tr>
<th>No</th>
<th>Component</th>
<th>Mole Fraction</th>
<th>MW</th>
<th>( Y_i \times MW )</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Nitrogen N\textsubscript{2}</td>
<td>0.01</td>
<td>28</td>
<td>0.28</td>
</tr>
<tr>
<td>2</td>
<td>Carbon dioxide CO\textsubscript{2}</td>
<td>0.015</td>
<td>44</td>
<td>0.66</td>
</tr>
<tr>
<td>3</td>
<td>Methane (C\textsubscript{1})</td>
<td>0.77</td>
<td>16.043</td>
<td>12.35</td>
</tr>
<tr>
<td>4</td>
<td>Ethane (C\textsubscript{2})</td>
<td>0.11</td>
<td>30.070</td>
<td>3.308</td>
</tr>
<tr>
<td>5</td>
<td>Propane (C\textsubscript{3})</td>
<td>0.06</td>
<td>44.097</td>
<td>2.665</td>
</tr>
<tr>
<td>6</td>
<td>i-Butane (i-C\textsubscript{4})</td>
<td>0.02</td>
<td>58.124</td>
<td>1.16</td>
</tr>
<tr>
<td>7</td>
<td>n-Butane(C\textsubscript{4})</td>
<td>0.01</td>
<td>58.124</td>
<td>0.58</td>
</tr>
</tbody>
</table>

Table 1-3 Gas mixture for Example 1-2

Solution: Using Table 1-2 & Equation 1-8

\[ MW = \sum Y_i (MW)_i \]
\[ MW = (\text{Mole Fraction of component } 1 \times \text{MW of component } 1) + (\text{Mole Fraction of component } 2 \times \text{MW of component } 2) + (\text{Mole Fraction of component } 3 \times \text{MW of component } 3) + \ldots etc. \]

The following table can be made:

<table>
<thead>
<tr>
<th>No</th>
<th>Component</th>
<th>Mole Fraction</th>
<th>MW</th>
<th>( Y_i \times MW )</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Nitrogen N\textsubscript{2}</td>
<td>0.01</td>
<td>28</td>
<td>0.28</td>
</tr>
<tr>
<td>2</td>
<td>Carbon dioxide CO\textsubscript{2}</td>
<td>0.015</td>
<td>44</td>
<td>0.66</td>
</tr>
<tr>
<td>3</td>
<td>Methane (C\textsubscript{1})</td>
<td>0.77</td>
<td>16.043</td>
<td>12.35</td>
</tr>
<tr>
<td>4</td>
<td>Ethane (C\textsubscript{2})</td>
<td>0.11</td>
<td>30.070</td>
<td>3.308</td>
</tr>
<tr>
<td>5</td>
<td>Propane (C\textsubscript{3})</td>
<td>0.06</td>
<td>44.097</td>
<td>2.665</td>
</tr>
<tr>
<td>6</td>
<td>i-Butane (i-C\textsubscript{4})</td>
<td>0.02</td>
<td>58.124</td>
<td>1.16</td>
</tr>
<tr>
<td>7</td>
<td>n-Butane(C\textsubscript{4})</td>
<td>0.01</td>
<td>58.124</td>
<td>0.58</td>
</tr>
</tbody>
</table>

http://www.oilprocessing.net/oil/
The apparent molecular weight is 21.36

1.5.2.3: Gas Specific Gravity and Density
The density of a gas is defined as the mass per unit volume as follows

\[
\text{Density} = \frac{\text{mass}}{\text{volume}} \quad \text{Eq. 1-9}
\]

The specific gravity of a gas (G) is the ratio of the density of the gas to the density of air at standard conditions of temperature and pressure.

\[
G = \frac{\rho(\text{gas})}{\rho(\text{air})} \quad \text{Eq. 1-10}
\]

Where

- \(\rho(\text{gas})\) = density of gas
- \(\rho(\text{air})\) = density of air

Both densities must be computed at the same pressure and temperature, usually at standard conditions. It may be related to the molecular weight as follows

\[
G = \frac{\text{MW}(\text{gas})}{\text{MW}(\text{air})} \quad \text{Eq. 1-11}
\]

Since molecular weight of air is 28.96 (table 1-2)

Specific gravity of gas \(G = \frac{\text{MW}(\text{gas})}{28.96}\) \quad \text{Eq. 1-12}

Example 1-3:
Determine the specific gravity of the gas mixture in example 1-2.
Solution: Apparent molecular weight of gas mixture is 21.36
Gas specific gravity = 21.36/28.96 = 0.7376

Since the gas is a compressible fluid, its density varies with temperature and pressure, calculating the gas density at a certain pressure and temperature will be explained after discussing the general gas law and gas compressibility factor.

1.5.2.4: General Gas Law
The general (Ideal) Gas equation, or the Perfect Gas Equation, is stated as follows:

\[
PV = nRT \quad \text{Eq. 1-13}
\]

Where

- \(P\) = gas pressure, psia (=psig + base pressure, usually 14.7 psi)
- \(V\) = gas volume, ft\(^3\)
- \(n\) = number of lb moles of gas (mass/molecular weight)
- \(R\) = universal gas constant, psia ft\(^3\)/lb mole \(\circ\)R
- \(T\) = gas temperature, \(\circ\)R \((\circ\)R = 460 + \(\circ\)F)

The universal gas constant \(R\) is equal to 10.73 psia ft\(^3\)/lb mole \(\circ\)R in field units.
Equation 1-13 is valid up to pressures of about 60 psia. As pressure increases above 60 psia, its accuracy becomes less and the system should be considered a non-ideal gas equation of state.

\[ PV = znRT \quad \text{Eq. 1-14} \]

Where

\( z = \) gas compressibility factor.

1.5.2.5 Compressibility and compressibility factor (\( z \))

Compressibility is the measure of the change in volume a substance undergoes when a pressure is exerted on the substance. Liquids are generally considered to be incompressible. For instance, a pressure of 16,400 psig will cause a given volume of water to decrease by only 5% from its volume at atmospheric pressure. Gases on the other hand, are very compressible. The volume of a gas can be readily changed by exerting an external pressure on the gas.

The Compressibility factor, \( Z \) is a dimensionless parameter less than 1.00 that represents the deviation of a real gas from an ideal gas. Hence it is also referred to as the gas deviation factor. At low pressures and temperatures \( Z \) is nearly equal to 1.00 whereas at higher pressures and temperatures it may range between 0.75 and 0.90. The actual value of \( Z \) at any temperature and pressure must be calculated taking into account the composition of the gas and its critical temperature and pressure. Several graphical and analytical methods are available to calculate \( Z \). Among these, the Standing-Katz, and CNGA methods are quite popular. The critical temperature and the critical pressure of a gas are important parameters that affect the compressibility factor and are defined as follows.

The critical temperature of a pure gas is that temperature above which the gas cannot be compressed into a liquid, however much the pressure. The critical pressure is the minimum pressure required at the critical temperature of the gas to compress it into a liquid.

As an example, consider pure methane gas with a critical temperature of 343 \( ^{0} \text{R} \) and critical pressure of 666 psia (Table 1-2).

The reduced temperature of a gas is defined as the ratio of the gas temperature to its critical temperature, both being expressed in absolute units (\( ^{0} \text{R} \)). It is therefore a dimensionless number. Similarly, the reduced pressure is a dimensionless number defined as the ratio of the absolute pressure of gas to its critical pressure.

Therefore, we can state the following:

\[ T_r = \frac{T}{T_c} \quad \text{Eq. 1-15} \]
\[ P_r = \frac{P}{P_c} \quad \text{Eq. 1-16} \]

Where

\( P = \) pressure of gas, psia
\( T = \) temperature of gas, \( ^{0} \text{R} \)
\( T_r = \) reduced temperature, dimensionless
\( P_r = \) reduced pressure, dimensionless
\( T_c = \) critical temperature, \( ^{0} \text{R} \)
\( P_c = \) critical pressure, psia

Example 1-4:

Using the preceding equations, the reduced temperature and reduced pressure of a sample of methane gas at 70 \( ^{0} \text{F} \) and 1200 psia pressure can be calculated as follows:

\[ T_r = \frac{(70 + 460)}{343} = 1.5 \]
Books are available [here](http://www.oilprocessing.net/oil/)

\[ P_r = \frac{1200}{666} = 1.8 \]

For natural gas mixtures, the terms pseudo-critical temperature and pseudo-critical pressure are used. The calculation methodology will be explained shortly. Similarly, we can calculate the pseudo-reduced temperature and pseudo-reduced pressure of a natural gas mixture, knowing its pseudo-critical temperature and pseudo-critical pressure.

The Standing-Katz chart, Fig. 1.2 can be used to determine the compressibility factor of a gas at any temperature and pressure, once the reduced pressure and temperature are calculated knowing the critical properties.

Pseudo-critical properties allow one to evaluate gas mixtures. Equations (1-17) and (1-18) can be used to calculate the pseudo-critical properties for gas mixtures:

\[
P'_c = \sum y_i P_{ci} \quad \text{Eq. 1-17}
\]

\[
T'_c = \sum y_i T_{ci} \quad \text{Eq. 1-18}
\]

where

- \( P'_c \) = pseudo-critical pressure,
- \( T'_c \) = pseudo-critical temperature,
- \( P_{ci} \) = critical pressure at component i, psia
- \( T_{ci} \) = critical temperature at component i, \( ^\circ \text{R} \)
- \( y_i \) = mole fraction of each component in the mixture, (\( \sum y_i = 1 \)).

**Example 1-5:**

Calculate the compressibility factor for the following gas mixture at 100\(^\circ\text{F}\) and 800 psig:

<table>
<thead>
<tr>
<th>Compound</th>
<th>Molecular weight</th>
<th>Mole fraction ( y_i )</th>
<th>Relative Density of gas (air=1)</th>
<th>Critical Temp. ( ^\circ\text{R} ), ( T_{ci} )</th>
<th>Critical pressure, Psia, ( P_{ci} )</th>
<th>( T'<em>c ) = ( \frac{T</em>{ci} y_i}{T_r} )</th>
<th>( P'<em>c ) = ( \frac{P</em>{ci} y_i}{P_r} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>CH(_4)</td>
<td>16.043</td>
<td>0.5</td>
<td>0.5539</td>
<td>343.0</td>
<td>666</td>
<td>171.5</td>
<td>333</td>
</tr>
<tr>
<td>C(_2)H(_6)</td>
<td>30.070</td>
<td>0.3</td>
<td>1.0382</td>
<td>549.6</td>
<td>706.6</td>
<td>164.9</td>
<td>212</td>
</tr>
<tr>
<td>C(_3)H(_8)</td>
<td>44.097</td>
<td>0.1</td>
<td>1.5225</td>
<td>665.6</td>
<td>615.5</td>
<td>66.6</td>
<td>61.6</td>
</tr>
<tr>
<td>C(_4)H(_10)</td>
<td>58.124</td>
<td>0.1</td>
<td>2.0068</td>
<td>734.5</td>
<td>527.9</td>
<td>73.5</td>
<td>52.8</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td><strong>1.0</strong></td>
<td></td>
<td></td>
<td><strong>464.5</strong></td>
<td><strong>659.4</strong></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Table 1-5 for Example 1-5.

Using Equation 1-15 and 1-16

\[
T'_r = \frac{(100+460)}{464.5} = 1.2
\]

\[
P'_r = \frac{(800+14.7)}{659.4} = 1.23
\]

From fig.1-2. Compressibility factor is approximately, \( z = 0.72 \)

Calculating the compressibility factor for example 1-4, of the gas at 70 \(^\circ\text{F}\) and 1200 psig, using Standing-Katz chart, fig. 1-2. \( z = 0.83 \) approximately. For \( (T_r = 1.5, P_r = 1.8) \).

Another analytical method of calculating the compressibility factor of a gas is using the “California Natural Gas Association” CNGA equation as follows:
Where
\[ P_{\text{avg}} = \text{Gas pressure, psig.} \quad [\text{psig} = (\text{psia} - 14.7)] \]
\[ T_f = \text{Gas temperature, } ^\circ\text{R} \]
\[ G = \text{Gas gravity (air = 1.00)} \]

The CNGA equation for compressibility factor is valid when the average gas pressure \( P_{\text{avg}} \) is greater than 100 psig. For pressures less than 100 psig, compressibility factor is taken as 1.00. It must be noted that the pressure used in the CNGA equation is the gauge pressure, not the absolute pressure.

Example 1-6:
Calculate the compressibility factor of a sample of natural gas (gravity = 0.6) at 80 \(^\circ\)F and 1000 psig using the CNGA equation.

Solution:
From the Eq. (1.19), the compressibility factor is
\[ Z = \frac{1}{1 + \left( \frac{1000 \times 344400(10)^{1.785 \times 0.6}}{(80 + 460)^{3.825}} \right)} = 0.8746 \]

The CNGA method of calculating the compressibility, though approximate, is accurate enough for most gas pipeline hydraulics work and process calculations.

1.5.2.6: Gas density at any condition of Pressure and temperature
Once the molecular weight of the gas is known, the density of a gas at any condition of temperature and pressure is given as:
\[ \rho_g = \frac{(MW)P}{RTZ} \quad \text{Eq. 1-20} \]

Since, Specific gravity of gas \( G = \frac{MW(\text{gas})}{28.96} \) and \( R=10.73 \), then
\[ \rho_g = 2.7 \frac{GP}{TZ} \quad \text{Eq. 1.21} \]
\[ \rho_g = 0.093 \frac{(MW)P}{TZ} \quad \text{Eq. 1.22} \]

where
\( \rho_g = \text{density of gas, lb/ft}^3 \),
\( P = \text{pressure, psia} \),
\( T = \text{temperature, } ^\circ\text{R} \),
Z = gas compressibility factor,
MW = apparent molecular weight of the gas.

Figure 1-2 Compressibility Factor for lean sweet natural gas (Surface Production Operations).
Example 1-7:
Calculate the pseudo-critical temperature and pressure for the natural gas stream composition given in example 1-2, calculate the compressibility factor, and gas density at 600 psia and 100°F.

Solution:

<table>
<thead>
<tr>
<th>No</th>
<th>Component</th>
<th>Mole Fraction $Y_i$</th>
<th>MW</th>
<th>$Y_i \times MW$</th>
<th>$T_{ec}$ °R</th>
<th>$Y_i \times T_{ec}$ °R</th>
<th>$P_{ec}$ psia</th>
<th>$Y_i \times P_{ec}$ psia</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>N$_2$</td>
<td>0.01</td>
<td>28</td>
<td>0.28</td>
<td>227.1</td>
<td>2.271</td>
<td>492.5</td>
<td>4.925</td>
</tr>
<tr>
<td>2</td>
<td>CO$_2$</td>
<td>0.015</td>
<td>44</td>
<td>0.66</td>
<td>547.4</td>
<td>8.211</td>
<td>1070</td>
<td>16.05</td>
</tr>
<tr>
<td>3</td>
<td>Methane (C$_1$)</td>
<td>0.77</td>
<td>16.043</td>
<td>12.35</td>
<td>343</td>
<td>264.11</td>
<td>666</td>
<td>512.82</td>
</tr>
<tr>
<td>4</td>
<td>Ethane (C$_2$)</td>
<td>0.11</td>
<td>30.070</td>
<td>3.308</td>
<td>549.6</td>
<td>60.456</td>
<td>706.6</td>
<td>77.726</td>
</tr>
<tr>
<td>5</td>
<td>Propane (C$_3$)</td>
<td>0.06</td>
<td>44.097</td>
<td>2.665</td>
<td>665.6</td>
<td>39.936</td>
<td>615.5</td>
<td>36.93</td>
</tr>
<tr>
<td>6</td>
<td>i-Butane (i-C$_4$)</td>
<td>0.02</td>
<td>58.124</td>
<td>1.16</td>
<td>734.5</td>
<td>14.69</td>
<td>527.9</td>
<td>10.558</td>
</tr>
<tr>
<td>7</td>
<td>n-Butane(C$_4$)</td>
<td>0.01</td>
<td>58.124</td>
<td>0.58</td>
<td>765.2</td>
<td>7.652</td>
<td>550.9</td>
<td>5.509</td>
</tr>
<tr>
<td>8</td>
<td>n-Pentane (C$_5$)</td>
<td>0.005</td>
<td>72.151</td>
<td>0.361</td>
<td>845.5</td>
<td>4.2275</td>
<td>488.8</td>
<td>2.444</td>
</tr>
<tr>
<td></td>
<td><strong>Total</strong></td>
<td><strong>1.0</strong></td>
<td><strong>21.36</strong></td>
<td><strong>451.5</strong></td>
<td><strong>667</strong></td>
<td><strong>667</strong></td>
<td><strong>667</strong></td>
<td></td>
</tr>
</tbody>
</table>

Table 1-6 solution of Example 1-7.

From the table MW= 21.36

$T_{ec} = 451.5$ °R

$P_{ec} = 667$ psia

From Eq. (1-15) and Eq. (1-16)

$T_r = T/T_{ec} = (100+460)/451.5 = 1.24$

$P_r = P/P_{ec} = 600/667 = 0.9$

Compressibility factor $z$ could be calculated from figure 1-2, or from Eq. 1-19.

Value from figure, $z = 0.83$

From Equation 1-19  $z = 0.87$

*For further calculations, we will calculate $z$ value using [Eq. 1-19]*

Using Eq. 1-22. density of gas

$\rho_g = 0.093 \left(\frac{21.36}{560}\right)600 \times 0.83 = 2.56$ lb/ft$^3$

Comparing $\rho_g$ at standard condition (z=1)

$\rho_g$ at standard condition = $0.093 \left(\frac{21.36}{560}\right)14.7 \times 1 = 0.056$ lb/ft$^3$

We can conclude that density increases with pressure while the volume decreases.

1.5.2.7: Gas volume at any condition of Pressure and temperature

Volume of a gas is the space occupied by the gas. Gases fill the container that houses the gas. The volume of a gas generally varies with temperature and pressure.

Volume of a gas is measured in cubic feet (ft$^3$).

Gas volume are commonly referred to in "standard" or "normal" units.

Standard conditions commonly refer to gas volumes measured at: 60°F and 14.696 psia.

The Gas Processors Association (GPA) Set standard molar volume conditions is:

$379.49$ std ft$^3$/lb- mol at 60°F, 14.696 psia. (= 379.5 approximately).

Therefore, each mole (n) contains about 379.5 cubic feet of gas (ft$^3$) at standard conditions.

http://www.oilprocessing.net/oil/
Therefore, by knowing the values of mass and density at certain pressure and temperature, the volume occupied by gas can be calculated.

Example 1-8:
Calculate the volume of a 10 lb mass of gas (Gravity = 0.6) at 500 psig and 80 °F, assuming the compressibility factor as 0.895. The molecular weight of air may be taken as 29 and the base pressure is 14.7 psia.
Solution:
The molecular weight of the gas (MW) = 0.6 x 29 = 17.4
Pressure =500+14.7 = 514.7 psia
Temperature = 80+460 = 540 °R
Compressibility factor z= 0.895
The number of lb moles n is calculated using Eq. 1-6. n=m/(MW)

\[ n = \frac{10}{17.4} \]
Therefore, \( n = 0.5747 \) lb mole
Using the real gas Eq. (1-14), \( PV=nzRT \)

\[ V = 0.895 \times 0.5747 \times 10.73 \times 540. \]
Therefore, \( V = 5.79 \) ft³

Example 1-9:
Calculate the volume of 1 lb mole of the natural gas stream given in the previous example at 120°F and 1500 psia (compressibility factor Z = 0.811).
Solution:
Using Eqn. (1-14), \( PV = nzRT \)

\[ V = 0.811 \times 1 \times 10.73 \times (120+460)/1500. \]
Therefore, \( V = 3.37 \) ft³

Example 1-10:
One thousand cubic feet of methane is to be compressed from 60°F and atmospheric pressure to 500 psig and a temperature of 50°F. What volume will it occupy at these conditions?
Solution:
Moles CH₄ (n) = 1000 / 379.5 = 2.64
At final conditions, (Compressibility factor z must be calculated), from equations 1-15 and1-16
\[ T_r = \frac{(460 + 50)}{344} = 1.88 \]
\[ P_r = \frac{(500 + 14.7)}{673} = 0.765 \]
From Figure 1-2. (Z = 0.94)
From Eqn. 1-14, \( PV = nzRT \)

\[ V = \frac{2.64 \times 10.7 \times 510}{514.7} = 26.3 \] ft³

Example 1-11:
One pound-mole of C₃H₈ (44 lb) is held in a container having a capacity of 31.2 cu ft. The temperature is 280°F. "What is the pressure?
Solution:
Volume = \( V = 31.2 \) ft³
A Trial-and-error solution is necessary because the compressibility factor Z is a function of the unknown pressure. Assume Z = 0.9.
Using Eqn. 1-14, \( PV = nzRT \)

\[ P \times 31.2 = 0.9 \times 1.0 \times 10.7 \times (460 + 280) \]
\[ P = 229 \text{ psia} \]
From table 1-2, Eqns. 1-15 and 1-16

\[ P_r = \frac{229}{616} = 0.37, \]
\[ T_r = \frac{665}{665} = 1.113 \]

According to Figure 1.2, the value of “Z” should be about 0.915 rather than 0.9. Thus, recalculate using Eqn. 1-14, the pressure is 232 rather than 229 psia.

Example 1-12:
Calculate the volume of gas (MW=20) will occupy a vessel with diameter 24 in, and 6 ft. length. At pressure 200 psia and temperature 100°F. (Assume compressibility factor z=0.9), and what will be the volume of gas at 14.7 psia and 60°F. Then calculate gas density and mass inside the container at pressure 200 psia and temperature 100°F.

Volume of vessel = \( \pi L r^2 \)

\[ V = 3.14 \times 6 \times (24/2^2) \times (2 \times 12)^2 = 18.8 \text{ ft}^3. \]

(We divided by 2 to get \( r \) from the diameter, and divided by 12 to convert from in. to ft.)

\[ T = 460 + 100 = 560 \text{ °R} \]

Using Eqn. 1-14, \( PV=nRT \)

\[ n = 18.8 \times 200 / (0.9 \times 10.73 \times 560) \]

\[ n = 0.7 \text{ lb. moles. (Remember gas volume ft}^3 = 379.5 \times n) \]

Volume of gas at 200 psia and 100°F = 0.7 \times 379.5 = 266 \text{ ft}^3

n of Gas at 14.7 psia and 60°F (z=1) = 18.8 \times 14.7 / (1 \times 10.73 \times 520)

n = 0.0495 lb. moles

Volume of gas at 14.7 psia and 60°F = 0.0495 \times 379.5 = 18.8 \text{ ft}^3

From this example (1-12), the gas volume will equal to the container volume at standard conditions (14.7 psia and 60°F).

Gas density is calculated using Eqn. 1-22

\[ \rho_g = 0.093 \times \frac{(MW)P}{TZ} \text{ lb/ft}^3 \]

Density of gas \( \rho_g = 0.093 \times 20 \times 200 / (0.9 \times 560) = 0.738 \text{ lb/ft}^3 \)

Mass of gas inside the vessel = Volume \times density = 0.738 \times 265 = 196 \text{ lb mass}

1.5.3: Velocity of fluids (Liquid and gas), (ft/s)

Single-phase liquid lines should be sized primarily on the basis of flow velocity. For lines transporting liquids in single-phase from one pressure vessel to another by pressure differential, the flow velocity should not exceed 15 feet/second at maximum flow rates, to minimize flashing ahead of the control valve. If practical, flow velocity should not be less than 3 feet/second to minimize deposition of sand and other solids. At these flow velocities, the overall pressure drop in the piping will usually be small. Most of the pressure drop in liquid lines between two pressure vessels will occur in the liquid dump valve and/or choke.

Flow velocities in liquid lines may be calculated using the following derived equation:

\[ V_l = 0.012 \frac{Q_l}{d_l^2} \quad \text{Eq. 1.23} \]

where

\[ V_l = \text{average flow velocity, feet/second.} \]
\[ Q_l = \text{liquid flow rate, barrels/day.} \]
\[ d_l = \text{pipe inside diameter, inches.} \]
The velocity of gas equal the volume flow rate \((\text{ft}^3)\) “Flow rate at operating conditions, (not at standard conditions) per second divided by flow area \((\text{ft}^2)\).

The velocity of gas flow in a pipeline represents the speed at which the gas molecules move from one point to another. Unlike a liquid pipeline, due to compressibility the gas velocity depends upon the pressure and, hence, will vary along the pipeline even if the pipe diameter is constant. The highest velocity will be at the downstream end, where the pressure is the least. Correspondingly, the least velocity will be at the upstream end, where the pressure is higher.

Consider a pipe transporting gas from point A to point B. Under steady state flow, at (A), the mass flow rate of gas is designated as \((M)\) and will be the same as the mass flow rate at point \((B)\), if between \((A)\) and \((B)\) there is no injection or delivery of gas. The mass being the product of volume and density, we can write the following relationship for point \((A)\):

\[
M = Q \rho \quad \text{Eq. 1.24}
\]

The volume rate \((Q)\) can be expressed in terms of the flow velocity \((V)\) and pipe cross sectional area \((A)\) as follows:

\[
Q = V A \quad \text{Eq. 1.25}
\]

Therefore, combining the above Equations, and applying the conservation of mass to points \((A)\) and \((B)\), we get

\[
M = V_1 A_1 \rho_1 = V_2 A_2 \rho_2 \quad \text{Eq. 1.26}
\]

where subscripts 1 and 2 refer to points \((A)\) and \((B)\), respectively. If the pipe is of uniform cross section between \((A)\) and \((B)\), then \(A_1 = A_2 = A\).

Therefore, the area term in Equation 1.26 can be dropped, and the velocities at \((A)\) and \((B)\) are related by the following equation:

\[
V_1 \rho_1 = V_2 \rho_2 \quad \text{Eq. 1.27}
\]

Example 1-13:

Calculate the gas velocity for gas flow rate 100 MMscfd through 24 in. internal diameter gas pipe, the gas specific gravity is 0.7, pressure 500 psia, Temperature 100 °F, and assume compressibility factor 0.85.

Solution: Using Eqn. 1-14, \(PV = n z RT\), and remember that \(n = V (\text{ft}^3)/379.5\).

\[
n = 100 \times \frac{10^6}{379.5}
\]

Gas volume at operating conditions \(V = 100 \times 10^6 \times 0.85 \times 10.73 \times 560 / (379.5 \times 500) = 2,695,000 \text{ ft}^3/\text{day}
\]

Gas flow rate cubic foot per second = \(2,695,000 / (24 \times 60 \times 60) = 31.2 \text{ ft}^3/\text{sec}
\]

Area of flow = \(\pi r^2 = 3.14 \times 12 \times 12 / (144) = 3.14 \text{ ft}^2\)

(We divided by 144 to convert \(r^2\) from \(\text{in}^2\) to \(\text{ft}^2\).)

Velocity of gas = \(31.2/3.14 = 9.9 \text{ ft/s.}\)

The gas velocity may be calculated directly from the following equation:

\[
\text{Velocity} = 6 ZTQ/(100,000 \times Pd^2) \quad \text{ft/s.} \quad \text{Eq. 1.28}
\]

Where \(Q = \text{Flow rate, scfd.}\)

d = diameter in inches.

Books are available [here](http://www.oilprocessing.net/oil/)
1.5.3.1 Maximum Recommended Velocity
The avoidance of pipe damage sets an upper limit on the capacity of the pipe. One criterion used to estimate the critical fluid velocity above which pipe damage may occur is found in API RP 14E, which suggests that a critical erosional velocity is expressed as

\[ V_{\text{max}} = \frac{C}{\sqrt{\rho}} = C \sqrt{\frac{ZRT}{29GP}} \]  
Eq. 1.29

where
- \(\rho\) = mixture density (lbm/ft.\(^3\)),
- \(V_e\) = erosional velocity threshold (ft./s),
- \(G\) = gas sp. (air=1),
- \(P\) = pressure psia,
- \(T\) = temperature, °R
- \(Z\) = compressibility factor
- \(R\) = gas constant,
- \(C\) = 125 for intermittent service,
- 100 for continuous service,
- 60 for corrosive service.

The erosional velocity represents the upper limit of gas velocity in a pipeline. The maximum recommended velocity of dry gas in pipes is 100 ft/s, (60 ft/s for wet gas), and to be less than the erosional velocity which is defined as:

For gas in Example 1-13, the erosional velocity \(V_{\text{max}}\) is:

\[ V_{\text{max}} = 100 \sqrt{0.85 \times 10.73 \times 560/(29 \times 0.7 \times 500)} \]
\[ V_{\text{max}} = 70.9 \text{ ft/s}. \]

1.5.4 Viscosity of Fluids
Viscosity is a measure of a fluid's internal resistance to flow. It is determined either by measuring the shear force required to produce a given shear gradient or by observing the time required for a given volume of liquid to flow through a capillary or restriction. When measured in terms of force, it is called absolute or dynamic viscosity. When measured with respect to time, it is called kinematic viscosity. A fluid’s kinematic viscosity is equal to its absolute viscosity divided by its density. The unit of absolute viscosity is poise or centipoise (cP). The unit of kinematic viscosity is “Stoke” or “centistokes” (cSt). The relationship between absolute and kinematic viscosity is given by

\[ \mu = \rho \ u_k = (SG) \ u_k \]  
Eq. 1.30

where
- \(\mu\) = absolute viscosity (cP),
- \(u_k\) = kinematic viscosity (cSt),
- \(\rho\) = density (gm/cm.\(^3\)),
- \(SG\) = specific gravity relative to water.

Fluid viscosity varies with temperature. For liquids, viscosity decreases with increasing temperature. As shown in Figure 1.3, liquid water at 70 °F has an absolute viscosity of approximately one centipoise (cP). The common English system unit of viscosity is lbm/ft.-s. The conversion between metric and English units are listed in the next table

http://www.oilprocessing.net/oil/
Multiply | By | To obtain
--- | --- | ---
ft²/sec | 92903.04 | Centistokes
lbf·sec/ft² (lb/ft²·sec) | 47880.26 | Centipoises
Centipoises | 1/density (g/cm³) | Centistokes
lbf·sec/ft² (lb/ft²·sec) | 32.174/density (lb/ft³) | ft²/sec
Centipoise | 0.000672 | lbf·sec/ft² (lbf·sec/ft²)

Table 1- 7 Viscosity conversion factors

Figure 1.3 Physical properties of water.

1.5.2.1 Crude oil Viscosity.
The viscosity of oil is highly dependent on temperature and is best determined by measuring the viscosity at two or more temperatures and interpolating to determine the viscosity at any other temperature. When data are not available, the viscosity of a crude oil can be approximated from Figures 1.4, and 1.5, provided the oil is above its cloud point temperature, that is, the temperature at which wax crystals begin to form when the crude oil is cooled (is the temperature at which paraffins first become visible in a crude sample). Figures 1.4, and 1.5, present kinematic viscosity for "gas-free" or stock tank crude oils. Although viscosity is generally a function of API gravity, it is not always true that a heavier crude (lower API gravity) has a higher viscosity than a lighter crude (higher API gravity). Therefore, Figures 1.4, and 1.5, should be used with caution. As shown in Table 1.8, the viscosity of crude varies from very low to very high.

As shown in figure 1.4 for crude "B." Solid phase high-molecular-weight hydrocarbons, "paraffins", can dramatically affect the viscosity of the crude sample. The effect of the cloud point on the temperature viscosity curve is shown for crude "B" in Figure 1-4. This change in the temperature-viscosity relationship can lead to significant errors in estimation. Therefore, care should be taken when one estimates viscosities near the cloud point.

The pour point is defined as the lowest temperature (5 °F) at which the oil will flow. The lower the pour point, the lower the paraffin content of the oil.

http://www.oilprocessing.net/oil/
Figure 1-4, typical viscosity-temperature curves for crude oils. (Courtesy of ASTM D-341.)

(Light crude oil (300–400 API), Intermediate crude oil (200–300), & Heavy crude oil (less than 200 API))

Figure 1-5, Oil viscosity vs. gravity and temp. (Courtesy of Paragon Eng. Services, Inc.)
In the absence of any laboratory data, correlations exist that relate viscosity and temperature, given the oil gravity. The following equation relating viscosity, gravity, and temperature was developed by Beggs and Robinson after observing 460 oil systems:

\[ \mu = 10^x - 1 \]  
Eq. 1-31

where
- \( \mu \) = oil viscosity, cp,
- \( T \) = oil temperature, °F,
- \( x = y (T)^{-1.63} \),
- \( y = 10^2 \)
- \( z = 3.0324 - 0.02023G \),
- \( G \) = oil gravity, API@ 60 °F.

*Figure 1-5 is a graphical representation of another correlation.*

<table>
<thead>
<tr>
<th>Crude</th>
<th>Country</th>
<th>Density (15 °C) (kg/m³)</th>
<th>Viscosity (40 °C) cSt</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ekofisk</td>
<td>Norway</td>
<td>804</td>
<td>2</td>
</tr>
<tr>
<td>Arabian</td>
<td>Saudi</td>
<td>859</td>
<td>6</td>
</tr>
<tr>
<td>Light Kuwait</td>
<td>Arabia</td>
<td>870</td>
<td>10</td>
</tr>
<tr>
<td>Bintulu</td>
<td>Kuwait</td>
<td>886</td>
<td>6</td>
</tr>
<tr>
<td>Schoonebee</td>
<td>Sarawak</td>
<td>904</td>
<td>200</td>
</tr>
<tr>
<td>Langunillas</td>
<td>Netherlands</td>
<td>967</td>
<td>800</td>
</tr>
<tr>
<td>Boscanc</td>
<td>Venezuela</td>
<td>1005</td>
<td>20,000</td>
</tr>
</tbody>
</table>

Table 1.8. Represents the independence of viscosity on crude oil density.

1.5.2.2 Oil-Water Mixture Viscosity

The viscosity of produced water depends on the amount of dissolved solids in the water as well as the temperature, but for most practical situations, it varies from 1.5 to 2 centipoise at 50°F, 0.7 to 1 centipoise at 100°F, and 0.4 to 0.6 centipoise at 150°F.

When an emulsion of oil and water is formed, the viscosity of the mixture may be substantially higher than either the viscosity of the oil or that of the water taken by themselves. The modified Vand’s equation allows one to determine the effective viscosity of an oil-water mixture and is written in the form

\[ \mu_{\text{eff}} = (1+2.5 \phi +10 \phi^2) \mu_c \]  
Eq. 1-32

where
- \( \mu_{\text{eff}} \) = effective viscosity, cp
- \( \mu_c \) = viscosity of the continuous phase (Oil), cp
- \( \phi \) = volume fraction of the discontinuous phase (Water).

1.5.2.3 Viscosity of gases

Viscosity of a fluid relates to the resistance to flow of the fluid. Higher the viscosity, more difficult it is to flow. Viscosity is a number that represents the drag forces caused by the attractive forces in adjacent fluid layers. It might be considered as the internal friction between molecules, separate from that between the fluid and the pipe wall.

The viscosity of a gas is very small compared to that of a liquid. For example, a typical crude oil may have a viscosity of 10 centipoise (cp), whereas a sample of natural gas has a viscosity of 0.019 cp.

Viscosity may be referred to as absolute or dynamic viscosity measured in cp or kinematic viscosity measured in centistokes (cSt). Other units of viscosity are lb/ft·sec for dynamic viscosity and ft²/s for kinematic viscosity.
Fluid viscosity changes with temperature. Liquid viscosity decreases with increasing temperature, whereas gas viscosity decreases initially with increasing temperature and then increases with further increasing temperature. Figure 1-6 can be used to estimate the viscosity of a hydrocarbon gas at various conditions of temperature and pressure if the specific gravity of the gas at standard conditions is known. It is useful when the gas composition is not known. It does not make corrections for H₂S, CO₂, and N₂. It is useful for determining viscosities at high pressure.

1.5.5 Phase behavior
Phase behavior refers to the mole fraction ratio of vapor to liquid of the fluid and a prediction of its value as a function of pressure, temperature, and composition. For NGL and volatile oils, calculation of the fluid phase behavior is necessary for the determination of pressure drop-flow rate relations. If the fluid composition is known, flash calculations can be performed to determine the phase envelope and the relative amounts of liquid and vapor in the two-phase region.
1.5.6 Waxy crude
The design and operation of facility piping and pipelines carrying crudes with very high pour points or with high wax content require special consideration. This section discusses waxy crude, its behavior and special design, and operational challenges involved in both the design and operation of waxy piping and pipelines. Refer to Figure 1.7 for vapor-pressure-temperature relationship chart for light petroleum products.

1.5.6.1 Paraffin wax
Waxy crude contains heavy paraffins. Heavy paraffins are saturated hydrocarbons that typically contain between 18 and 34 carbon atoms in a chain. The wax formed is of crystalline structure and can be soft with a high percentage of trapped oil similar to Vaseline or can form hard deposits like candle wax.

1.5.6.2 Waxy crude behavior
When the temperature drops too low for wax to remain dissolved in a crude, it precipitates out of solution and deposits itself on the inside wall of facility piping and pipelines or inside facility process components. The “cloud point” or the “wax appearance temperature” (WAT) is defined as the temperature at which wax crystals can first be detected. When the temperature in facility piping or pipeline system drops below the cloud point, wax crystals begin to form and deposit. Any additional decrease in temperature causes additional wax to come out of solution until the crude in the piping system gels up. The temperature at which this occurs is called the “pour point.” When the crude in the pipe gels up, a certain force (yield stress) is required to shear the waxy crude and restart the flow.

In practice, a crude is considered to have a high wax content when there is more than 10% wax, while a crude is considered to have low wax content when there is less than 4%. Some examples of low-wax-content crudes are shown in Table 1.9. Even though the wax content of each of the crudes is 4%, there are significant differences in pour point, that is, the temperature at which the crude gels up. As is shown, the Malaysian Labuan crude gels up at 48 °F (9 °C), while the Saudi Arabian light crude would gel only if the temperature drops below -32.8 °F (-36 °C).

Table 1.10 shows some examples of crude with high wax content and relatively high pour points. Temperatures in a pipeline between 68° and 86 °F (20-30 °C) are not uncommon, such as the shutdown of a subsea pipeline. Under this condition, these crudes will gel up. To understand the behavior of waxy crude, one needs to know the following:

- Cloud point
- Pour point
- Wax content
- Yield stress

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تحميل مجاني لأول كتاب بالعربية عن معالجة البترول والغاز وهو ترجمة للكتاب الإنجليزي
http://oilprocessing.net/data/documents/ogparabic.pdf

ممكن تفتح الملف وتحفظه بعد ذلك، ومن الممكن أن تختار من بين الماس وهو على الرابط (حفظ الملف أو الهدف ك... ) وتقوم بتثبيته مباشرة على اللاب توب أو الكمبيوتر

http://www.oilprocessing.net/oil/
Vapour Pressure Psig

Figure 1-7 Vapor pressure for light hydrocarbons.

http://www.oilprocessing.net/oil/
1.5.6.3 Wax prediction

The potential for wax formation in piping and pipelines can be predicted by sampling and analyzing the crude for the following properties:

- Cloud point—cold finger technique or cross polar microscopy
- Pour point—differential scanning calorimetry
- Wax content—solid-phase analysis
- Carbon distribution—gas chromatography

In order to achieve accurate results, it is imperative representative samples must be taken and that the correct experimental procedures be followed. Even if this is done, reproducibility is difficult, and one is very likely to get a range of results, rather than one definitive value, for example, cloud point.

1.5.6.4 Design and operational challenges

Waxy crude with high cloud points and pour points poses special design and operational challenges. These include the following:

- Wax deposits on the internal pipe walls of flow lines and pipeline reduce internal diameter, which causes pressure and production losses.
- Removal of wax deposits requires production shutdowns or extra pigging, which results in loss of revenue or extra costs.
- Flow lines and pipelines may need to be designed with costly insulation to ensure the crude remains warm and above the wax deposition temperature.
- Chemicals are added to prevent wax formation and deposition, which is expensive.
- Chemicals may need to be injected before a shutdown to prevent crude from gelling up in the pipeline, if it cools below the pour point.

Designers and pipeline operations personnel need to thoroughly evaluate the aforementioned factors when confronted with waxy crude. When deciding how to manage the production and
transportation of waxy crude, it is imperative that one carefully consider both capital (CAPEX) and operating expenses (OPEX) over the life cycle of the field for different options.

1.5.6.5 Wax management
A development plan for a waxy crude field will be different and more likely more expensive than a non-waxy crude. Thus, early prediction is important. There are commercially available wax prediction tools available that will predict the wax deposition rate. However, one needs to be cautious, as the results may not be reliable. There are several options to choose from when one needs to manage waxy crude production. These include the following:

- Wax inhibitors, growth modifier, or dispersants
- Pour point reducer
- Dilution (reduces wax content)
- Pipeline insulation
- Heating or steam tracing the pipeline
- Regular pipeline pigging

1.6 Flow conditions
1.6.1 Flow potential
The flow potential is the total pressure drop available to transport fluid in a section of pipe. This potential, the difference in inlet and discharge pressures, includes the elevation pressure effect due to hydraulic pressure in an inclined pipe. If the flow potential does not equal or exceed that required to flow a given quantity of fluid, a larger diameter pipe, pump, or compressor must be specified.

1.6.2 Flow regimes
When a fluid moves through a pipe, two distinct types of flow are possible, laminar and turbulent. Laminar flow occurs in fluids moving with small average velocities. As the velocity increases, a “critical” point is reached at which the flow regime changes to turbulent flow. This “critical” point varies depending upon the pipe diameter, the fluid density and viscosity, and the velocity of flow. Turbulent flow becomes apparent as the velocity is increased above a critical velocity. In laminar flow, the fluid particles move along the length of the pipe in a very orderly fashion, with little or no sideways motion across the width of the pipe. Turbulent flow is characterized by random, disorganized motion of the particles, from side to side across the pipe as well as along its length. The two types of fluid flow are described by different sets of equations. In general, for most practical situations, the flow will be turbulent.

![Laminar and turbulent flow in pipes.](http://www.oilprocessing.net/oil/)
1.6.3 Reynolds Number

A useful factor in determining which type of flow is involved is the Reynolds number. This is the ratio of the dynamic forces of mass flow to the shear resistance due to fluid viscosity and is given by the following formulas.

\[ \text{Re} = \frac{DV}{\rho/\mu} = 1488 \quad \text{or} \quad \text{Re} = \frac{124DV}{d/\mu} \quad \text{Eq. 1.33} \]

where
- \( \text{Re} \) = Reynolds number, dimensionless
- \( V \) = average gas velocity, ft/s
- \( D \) = pipe inside diameter, ft
- \( d \) = pipe inside diameter, in.
- \( \rho \) = gas density, lb/ft\(^3\)
- \( \mu \) = gas viscosity, lb/ft-s (1 cp = 0.000672 lb/ft-s (lb/ft./s)) (or = Centipoise divided by 1488, or (Centistoke times specific gravity) divided by 1488.)
- \( \mu \) = absolute viscosity, cP

The flow is considered to be laminar flow when the Reynolds number is below 2000. Turbulent flow is said to exist when the Reynolds number is greater than 4000. When the Reynolds numbers is between 2000 and 4000, the flow is called critical flow, or undefined flow. Therefore

- \( \text{Re} \leq 2000 \) Flow is laminar
- \( \text{Re} > 4000 \) Flow is turbulent
- \( 2000 < \text{Re} \leq 4000 \) Flow is critical flow

In terms of the more commonly used units in the gas pipeline industry, the following formula for Reynolds number is more appropriate:

\[ \text{Re} = 1.35 \times 10^{-5} \left( \frac{S\text{Q}}{\mu d} \right) \quad \text{Eq. 1.34} \]

where
- \( S \) = gas specific gravity (air = 1)
- \( \text{Q} \) = gas flow rate, standard ft\(^3\)/day (scfd)
- \( d \) = pipe inside diameter, in.
- \( \mu \) = gas viscosity, lb/ft-s (1 cP = 0.000672 lb/ft-s)

Or

\[ \text{Re} = 20,100 \left( \frac{S\text{Q}_{g}}{\mu d} \right) \quad \text{Eq. 1.35} \]

\( \mu \) = gas absolute viscosity, cP
\( Q_{g} \) = gas flow rate, MMscf/D

Reynolds number for liquids

The Reynolds number can be expressed in more convenient “oil field” terms. For liquids, as follows:

\[ \text{Re}_{l} = 92.1 \left( \frac{\text{Q}_{l}}{\mu d} \right) \quad \text{Eq. 1.36} \]

where
- \( \mu \) = absolute viscosity (cP),
- \( d \) = pipe ID (in.),
- (SG) = specific gravity of liquid relative to water,
- \( Q_{l} \) = liquid flow rate (BPD).
1.6.4 Pipe roughness
In laminar flow, the pipe wall roughness does not materially affect the pressure gradient. In turbulent flow, the pipe wall roughness has a marked effect on the pressure gradient. As the Reynolds number increases, the fluid boundary layer on the pipe wall becomes thinner, exposing the irregularities in the pipe wall to the high velocity outside the boundary layer. Techniques for quantifying relative roughness, or ratio of wall irregularity height to pipe diameter, are discussed later in this chapter.

1.6.5 Rate
Fluid flow rate \( V = Q/A \) through a pipe is primarily controlled by pressure loss due to friction. The higher the flow rate, the higher the frictional pressure drop. Reducing frictional pressure drop requires lower velocities. The capacity of a pipe is determined by the velocity developed at the design or allowable pressure drop (flow potential).

1.6.6 Velocity limitations
When fluid flow rates increase, fluid velocities can increase until the pipe wall is actually damaged. Refer to 1.5.3 Velocity of fluids.

1.6.7 Temperature
Fluid temperature can affect the pressure drop-flow rate relations and thus affects the choice of pipe size. Fluid temperature can also affect the fluid density and thus influence the erosional velocity limitations.

1.6.7.1 Gas considerations
When a gas with even small amounts of heavier hydrocarbons (propane and heavier) is compressed and/or cooled sufficiently, the heavier components may condense into a separate liquid phase. The presence of two phases flowing concurrently in a pipe causes a considerably higher frictional pressure drop than an equivalent mass flow rate of gas in the same diameter pipeline.
The presence of even 1% by volume liquid phase can cause a 20-30% increase in frictional pressure drop. The presence of liquid can also increase elevation pressure drop. Since liquids tend to be much heavier than gases, the presence of liquid in the flow increases the effective fluid density, thus increasing hydrostatic pressure.

1.6.7.2 Liquid considerations
Temperature strongly influences liquid viscosity. Viscosity increases exponentially with decreasing temperature. The increasing viscosity causes a decreasing Reynolds number and an increasing frictional pressure gradient. If the Reynolds number becomes sufficiently low and the flow becomes laminar, the frictional pressure gradient becomes inversely proportional to viscosity.
1.7 Special considerations

1.7.1 Emulsions
The characteristics of oil-water emulsions vary greatly and are difficult to characterize or predict. Caution should be exercised and experienced help sought in dealing with emulsion in fluid flow calculations.

1.7.2 Pigging
A pig is a spherical or cylindrical device that is pumped through a pipe.

1.7.2.1 Liquid removal
Liquid accumulation at the bottom of a gas pipeline can cause erratic “slug” flow of accumulated liquids and a reduction of the effective pipe cross-sectional flow area, which reduces line capacity. The passage of a pig through the line functions like a piston and pushes the accumulated liquid toward a slug catcher, thus restoring the line capacity. Pigs used for liquid removal are generally inflatable spheres, rather than like rubber balls. By adjusting the inflation pressure of the sphere, the tightness of fit in the pipe can be altered.

1.7.2.2 Corrosion protection
Water in a low section of pipe can become a corrosion hazard and must be removed. Often, a slug of corrosion inhibitor is injected and a second slightly underinflated pig is run to allow the inhibitor to coat the pipe wall.

1.7.2.3 Cleaning
Special types of scrapper pigs are run to remove paraffin, scale, rust, or construction debris from the inside of the pipe wall.

1.7.2.4 Monitoring
Special monitoring pigs are run to inspect the condition of the pipe. Feeler gauges on the pipe wall can check for wall thickness loss due to corrosion. Inclinometers and other position sensors have been placed in pigs to monitor pipe movement due to soil shifting. Increasing sophisticated pigs are being used for a widening range of pipeline monitoring tasks.

1.7.2.5 Auxiliary equipment
Much care and planning must be exercised before pigging a pipeline. Special equipment to launch and receive pigs must be provided. Line, fittings, and valves must be examined to assure passage and controlled removal of the pig from the line. Pig velocity and travel must also be controlled through fluid pumping rates, upstream and downstream pressures, and the use of “pig sig” equipment that provides positive external signal information about internal arrival or passage.

1.7.3 Water hammer
The preceding discussion assumed that the fluid flow is steady-state. Transients have been assumed to be unimportant in the sizing of piping. One aspect of liquid flow line sizing in which transients can be extremely important involves “water hammer.” Whenever a valve in a liquid line is closed, all the liquid flowing in the pipe upstream of the closed valve must be slowed and brought to rest. Since this liquid column can be massive and is relatively incompressible, a large
transient pressure at the downstream end of the pipe may be required. If the valve closure is abrupt, and the fluid initial velocity is sufficiently high, then a destructive pressure surge may result.

This pressure surge travels upstream in the flow line and may exceed the burst pressure of the pipe. The familiar knocking of household plumbing resulting from rapid faucet closure is an example of this pressure pulse deforming the piping and causing it to knock or hammer against its supports. "Water hammer" is mainly found only in liquid lines due to the near incompressibility of liquid.

1.7.4 Line packing
Fluid transient effects can also be important in long gas pipelines as a method of smoothing demand peaks. Since gas is much more compressible than liquid, potentially destructive liquid pressure surges do not occur as they may in liquid lines.

The compressibility of gas permits the pipeline to serve as a long, slender, pressure vessel for the storage of gas. By allowing the pressure levels in a gas pipeline to fluctuate, the instantaneous gas input to a pipeline does not have to equal the instantaneous gas offtake from the pipeline.

The practice of pumping the line pressure up to a high pressure during periods of low demand permits the pipeline to deliver gas quantities in excess of its steady-state capacity under conditions of high demand. The process of raising pipeline operating pressures in anticipation of greater future demand is called "line packing."

1.7.5 Line drafting
The process of drawing pipeline inventory down during periods of peak demand is called "line drafting." While pipeline pack and draft are of small importance in production operations, gas transmission companies are quite sophisticated in using pack and draft to smooth their demand curves. The capacity to "pack" a gas pipeline is related to the difference in the normal operating pressure and the MAWP.

1.7.6 Phase flow regimes
1.7.6.1 Single-phase flow
Single-phase liquid flows are characterized by a virtually constant density but may be strongly influenced by viscosity effects. Single-phase gas flows are characterized by low viscosity, which result in high Reynolds numbers. Pipe wall roughness tends to be a more important effect in determining frictional pressure drop than does viscosity.

The density of gases varies greatly with both pressure and temperature and is the cause for the development of separate pressure drop-flow rate equations for two-phase flow.

In some single-phase flow conditions, a small volume of gas may be entrained in the liquid flow, such as a liquid dump line from a separator, or a small amount of liquid may be carried in the pipe in gas flow, such as the gas outlet line off a separator. These small amounts usually have a negligible effect on pressure loss and are not considered in single-phase flow calculations.

However, there are certain flow conditions where sufficient volumes of a second gas or liquid phase exist to produce an appreciable effect on pressure loss. The pressure drop in such lines must be considered using techniques for two-phase flow.

1.7.6.2 Two-phase flow
When vapor and liquid flow in a pipe simultaneously, the flow situation becomes much more complicated. Liquid and vapor phases will distribute according to flow conditions, pipeline

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geometry, and the effect of gravity. Pressure drop-flow rate relations change as well. Using the best correlations available for pressure drop and liquid holdup, predictions may be in error by as much as +/- 20% for perfectly horizontal pipe and +/- 50% for inclined and looped systems. A phase diagram should be prepared for the flowing fluid. Plotting the anticipated flowing pressure and temperature will determine the number of phases present. Generally, in production operations, two-phase flow exists from the wellbore to the first separator in the production facility. Gas from the separator is considered single-phase flow even though entrained liquids are present. Liquid from the separator is considered single-phase flow even though gas is present due to pressure drop through a liquid dump valve. Different liquid and gas distributions in the pipe occur for different flow conditions.

1.8 Volumetric and Mass Flow Rates

1.8.1 Volumetric Flow Rate

The *volumetric flow rate* \( Q_V \) of a system is a measure of the volume of fluid passing a point in the system per unit time. The volumetric flow rate can be calculated as the product of the cross-sectional area \( A \) for flow and the average flow velocity \( V \).

\[
Q_V = A \times V \quad \text{Eq. 1.37}
\]

Where

- \( Q_V \) = Volumetric flow rate, \( \text{ft}^3/\text{sec} \)
- \( A \) = Cross-sectional area, \( \text{ft}^2 \)
- \( V \) = Velocity, \( \text{ft/sec} \)

Example 1-14:
A pipe with an inner diameter of 4 inches (4/12 ft.) contains water that flows at an average velocity of 14 feet per second. Calculate the volumetric flow rate of water in the pipe.
Solution:
\[
Q_V = A \times V = \left( \frac{\pi \times (4/12)^2}{4} \right) \times 14
\]
\[
= 3.14 \times \frac{1}{4} \times 14
\]
\[
= 1.22 \text{ ft}^3/\text{sec}
\]

1.8.2 Mass Flow Rate

The *mass flow rate* \( m \) of a system is a measure of the mass of fluid passing a point in the system per unit time. The mass flow rate is related to the volumetric flow rate as shown in Equation 1-38, where \( \rho \) is the density of the fluid.

\[
m = \rho \times Q_V \quad \text{Eq. 1.38}
\]

where

- \( Q_V \) = volumetric flow rate is in cubic feet per second, \( \text{ft}^3/\text{sec} \)
- \( \rho \) = the density is in pounds-mass per cubic foot, \( \text{lbm}/\text{ft}^3 \)
- \( m \) = mass flow rate measured in pounds-mass per second, \( \text{lbm/sec} \)

From Equations 1-38 & 1-37:

\[
m = \rho \times A \times V \quad \text{Eq. 1.39}
\]

Example 1-15:
The water in the pipe of example 1-14, had \( \rho = 62.44 \text{ lbm/ft}^3 \). Calculate the mass flow rate.

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Solution:
\[ m = \rho Qv \]
\[ m = 62.44 \times 1.22 = 76.2 \text{ lbm/sec} \]

1.9 Networks
Piping systems are composed of more than a single line of constant diameter pipe. Several pipes may be manifolled together, and even runs of a single line may change diameter. For very complicated systems, hand calculations may be prohibitively complicated, and computer programs must be used. For simple series and parallel lines, composite pressure drop-flow rate relations can be calculated using a couple of simple procedures.

1.9.1 Pipes in series
Pressure drops are additive in pipes in series. For two sections of pipe, the downstream section having a different diameter from the upstream section, the overall pressure drop can be calculated by determining the pressure drop of the upstream and downstream sections separately. The composite pressure drop is then calculated by adding the pressure drops of the upstream and downstream sections. Note that the volume flow rate \( (Q) \) is the same in the upstream and downstream segments of pipes in a series.

1.9.2 Pipes in parallel
For two pipe segments in parallel, the pressure drops are identical for each segment and the flow rates are additive. Figure 1.9 shows the comparison between these two conditions.

Figure 1.9. Series and parallel piping systems.

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1.10 Continuity Equation

The continuity equation is simply a mathematical expression of the principle of conservation of mass. For a control volume that has a single inlet and a single outlet, the principle of conservation of mass states that, for steady-state flow, the mass flow rate into the volume must equal the mass flow rate out. The continuity equation for this situation is expressed by Equation

\[ m_{\text{inlet}} = m_{\text{outlet}} \quad \text{Eq. 1.40} \]

\[ (\rho Av)_{\text{inlet}} = (\rho A v)_{\text{outlet}} \quad \text{Eq. 1.41} \]

One of the simplest applications of the continuity equation is determining the change in fluid velocity due to an expansion or contraction in the diameter of a pipe.

Example 1-16:
Steady-state flow exists in a pipe that undergoes a gradual expansion from a diameter of 6 in. to a diameter of 8 in. The density of the fluid in the pipe is constant at 60.8 lbm/ft³. If the flow velocity is 22.4 ft/sec in the 6 in. section, what is the flow velocity in the 8 in. section?

Solution:
From the continuity equation we know that the mass flow rate in the 6 in. section must equal the mass flow rate in the 8 in. section. Letting the subscript (1) represent the 6 in. section and (2) represent the 8 in. section we have the following.

\[ m_1 = m_2 \]

\[ \rho_1 A_1 v_1 = \rho_2 A_2 v_2 \]

\[ v_2 = v_1 \left( \frac{\rho_1}{\rho_2} \right) \left( \frac{A_1}{A_2} \right). \quad \text{since, } \rho_1 = \rho_2 \]

\[ = v_1 \left( \frac{\pi r_1^2}{\pi r_2^2} \right) \]

\[ = 22.4 \times \frac{3^2}{4^2} = 12.6 \text{ ft/sec} \]

Example 1-17:
The inlet diameter of a pump shown in Figure 1-10 is 28 in. while the outlet flow through the pump is 9200 lbm/sec. The density of the water is 49 lbm/ft³. What is the velocity at the pump inlet?

Solution:
\[ A_{\text{inlet}} = (3.14) (14/12)^2 = 4.28 \text{ ft}^2 \]

\[ m_{\text{inlet}} = m_{\text{outlet}} = 9200 \text{ lbm/sec} \]

\[ (\rho Av)_{\text{inlet}} = 9200 \text{ lbm/sec} \]

\[ V_{\text{inlet}} = 9200 / (4.28 \times 49) = 43.9 \text{ ft/sec} \]

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Example 1-18:
A piping system has a "Y" configuration for separating the flow as shown in Figure 1-11. The diameter of the inlet leg is 12 in., and the diameters of the outlet legs are 8 and 10 in. The velocity in the 10 in. leg is 10 ft/sec. The flow through the main portion is 500 lbm/sec. The density of water is 62.4 lbm/ft$^3$. What is the velocity out of the 8 in. pipe section?

Solution:
\[
A_8 = 3.14 \times (4/12)^2 = 0.349 \text{ ft}^2
\]
\[
A_{10} = 3.14 \times (5/12)^2 = 0.545 \text{ ft}^2
\]
\[\text{m}_\text{inlets} = \text{m}_\text{outlets}\]
\[\text{m}_8 = \text{m}_{12} - \text{m}_{10}\]
\[(p \text{ Av})_8 = \text{m}_8 \times (p \text{ Av})_{10}\]
\[(p \text{ Av})_8 = 500 - (62.4 \times 0.545 \times 10) = 159.9 \text{ lbm/sec}\]
\[v_8 = 159.9 / (0.349 \times 62.4) = 7.3 \text{ ft/sec}\]
1.11 Average pipeline pressure
The gas compressibility factor Z used in the General Flow equation is based upon the flowing temperature and the average pipe pressure. The average pressure may be approximated as the arithmetic average

\[ P_{avg} = \frac{P_1 + P_2}{2} \]

of the upstream and downstream pressures \( P_1 \) and \( P_2 \). However, a more accurate average pipe pressure is usually calculated as follows

\[ P_{avg} = \frac{2}{3} \left( P_1 + P_2 - \frac{P_1 P_2}{P_1 + P_2} \right) \quad \text{Eq. 1.42} \]

Where
\( P_1, P_2, P_{avg} \) = pressure, psia

Another form of the average pressure in a pipe segment is

\[ P_{avg} = \frac{2}{3} \left( P_1^3 - P_2^3 \right) / \left( 3 \left( P_1^2 - P_2^2 \right) \right) \quad \text{Eq. 1.43} \]

1.12 Fluid head, friction losses, and Bernoulli’s equation

1.12.1 Fluid head
The term “head” is used to represent the vertical height of a static column of a liquid corresponding to the mechanical energy contained in the liquid per unit mass. Head can also be considered as the amount of work necessary to move a liquid from its original position to its final required delivery position. Accordingly, this includes the additional work necessary to overcome the resistance to flow in the line.

A liquid at any point consists of the following three kinds of head:

1.12.1.1 Pressure head
Static pressure head represents the energy contained in the liquid due to its pressure. The pressure head has dimensions of length and is calculated in units of feet or meters.

1.12.1.2 Velocity head
Velocity head represents the kinetic energy contained in the liquid due to its velocity. This head is usually very small, and while it would be technically incorrect to disregard it, for practical piping problems, its contribution is often negligible. The velocity head is particularly important when calculating friction head losses through fittings.

1.12.1.3 Elevation head and Relationship between Depth and Pressure
Elevation head represents the energy contained in the liquid due to its position measured by the vertical height above some plane of reference. This is the head due to potential energy and accounts for that portion of the head required to move the fluid from one elevation to another. The pressure of a liquid is directly proportional to the depth, and for a given depth the liquid exerts the same pressure in all directions.

As shown in Figure 1-12 the pressure at different levels in the tank varies and this causes the fluid to leave the tank at varying velocities. Pressure was defined to be force per unit area. In the case of this tank, the force is due to the weight of the water above the point where the pressure is being determined.

Example:
Pressure = Force / Area
= Weight / Area
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\[
P = \frac{m \, g}{A g_c} \quad \text{Eq. 1.44}
\]

\[
P = \rho \, V \, g / A g_c \quad \text{Eq. 1.45}
\]

where:
- \( m = \) mass in lbm
- \( g = \) acceleration due to earth's gravity 32.17 ft/sec²
- \( g_c = 32.17 \text{ lbm-ft/lbf-sec}^2 \)
- \( A = \) area in ft²
- \( V = \) volume in ft³
- \( \rho = \) density of fluid in lbm/ft³

The volume is equal to the cross-sectional area times the height (h) of liquid. Substituting this into the above equation yields:

\[
P = \rho \, A \, h \, g / A g_c
\]

\[
P = \rho \, h \, g / g_c \quad \text{Eq. 1.46}
\]

![Figure 1.12 Pressure versus Depth](image)

Equation 1.46, tells us that the pressure exerted by a column of water is directly proportional to the height of the column and the density of the water and is independent of the cross-sectional area of the column. The pressure thirty feet below the surface of a one inch diameter standpipe is the same as the pressure thirty feet below the surface of a large lake.

Example 1-19:
If the tank in Figure 1.12 is filled with water that has a density of 62.4 lbm/ft³, calculate the pressures at depths of 10, 20, and 30 feet.

Solution:
\[ P_{10 \text{ feet}} = \rho \ h \ g / g_c \]
\[ P = 10 \times 62.4 \]
\[ = 624 \text{ lbf/ft}^2 \]
\[ = 624/144 = 4.33 \text{ lbf/in}^2 \ (\text{psi}) \]
\[ P_{10 \text{ feet}} = 4.33 \text{ psi} \]
\[ P_{20 \text{ feet}} = 20 \times 62.4 / 144 = 8.67 \text{ psi} \]
\[ P_{30 \text{ feet}} = 30 \times 62.4 / 144 = 13 \text{ psi} \]

Therefore, each feet (height) of water will result in 0.433 psi, which is call water pressure gradient. Consequently, each feet of fluid with specific gravity 0.8 (for example), will result in (0.433 x 0.8) psi.

Example 1-20:
A cylindrical water tank 40 ft high and 20 ft in diameter is filled with water that has a density of 61.9 lbm/ft³.
(a) What is the water pressure on the bottom of the tank?
(b) What is the average force on the bottom?
Solution:
(a)
\[ P = \rho \ h \ g / g_c \]
\[ P = 61.9 \times 40 / 144 = 17.2 \text{ psi (lbf/in}^2) \]

(b)
\[ \text{Pressure} = \text{Force} / \text{Area} \]
\[ \text{Force} = (\text{Pressure}) \times (\text{Area}) \]
\[ \text{Area} = \pi r^2 \]
\[ \text{Force} = 17.2 \times 3.14 \times 10^2 \times 144 = 7.78 \times 10^5 \text{ lbf} \]
\[ N.B. \ (10^2 \times 144) \text{ same as } (10 \times 12)^2. \]

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ممكن تفتح الملف وتخفظه بعد ذلك, ومن الممكن أن تختار من يمين الماوس وهو على الرابط (احفظ الملف أو الهدف ك... ) وتقوم بتوزيله مباشرة على اللاب توب أو الكمبيوتر.
Friction Losses & Pressure Drop Equations

2.1 Friction losses

Although there are a number of software programs available to compute friction losses, the ability to perform hand calculations should not be underestimated. Friction losses depend on the following variables:

- Flow rate
- Diameter
- Type of pipe (relative roughness)
- Length of pipe (major losses)
- Number and sizes of fittings, valves, and accessories (minor losses)
- Entrance and exit losses (minor losses)

The inside diameter of the pipe is determined by fluid velocity. The velocity is usually based on recommended rules of thumb that have been used with success over the years. Refer to Chapter 2 and API RP 14E for suggested velocities for liquids, gases, and two-phase flow.

Depending on the particular application, liquid velocities should be between 3 and 15 fps (1-5 m/s). The danger of creating water hammer increases above 15 fps. Pump suction lines should always be sized for low velocities between 2 and 5 fps (0.6-1.5 m/s), thus reducing the likelihood of creating cavitation in the pump.

Roughness of the pipe is a function of the pipe material and its age. Ferrous pipes that have been in service many years may be corroded or scaled, and the relative roughness for high flow rates will have increased due to these surface irregularities. On the other hand, for low flow rates (laminar flow), it turns out that the roughness does not matter at all.

The number of fittings and other factors such as entrance and exit losses that contribute to the minor losses may be converted into an "equivalent length," which is then added to the length of straight pipe for a total length. This exercise must be performed for each segment that has a different diameter in the system.

2.2 Bernoulli’s equation

Bernoulli’s equation states that: as a fluid flows from one point to another point in a piping system, the total of the static pressure, potential, and velocity heads at the upstream point (subscript 1) equals the total of the three heads at the downstream point (subscript 2) plus the frictional head loss between points 1 and 2, and is expressed as follows:

\[
(H_s)_1 + (H_p)_1 + (H_v)_1 = (H_s)_2 + (H_p)_2 + (H_v)_2 + H_L \quad \text{Eq. 2.1}
\]

where
- \(H_s\) = static pressure head (ft.),
- \(H_p\) = pressure head (ft.),
- \(H_v\) = velocity head (ft.).

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\[ H_L = \text{frictional head loss (ft.)}. \]

When Bernoulli’s equation is written for a section of pipe carrying a fluid of constant fluid properties, the resulting equation is expressed as

\[ Z_1 + \left(\frac{144P_1}{\rho_1}\right) + \left(\frac{V_1^2}{2g}\right) = Z_2 + \left(\frac{144P_2}{\rho_2}\right) + \left(\frac{V_2^2}{2g}\right) + H_L \quad \text{Eq. 2.2} \]

where
\[ V_1 = \text{upstream fluid velocity (ft./s)}, \]  
\[ V_2 = \text{downstream fluid velocity (ft./s)}, \]  
\[ P_1 = \text{upstream pressure (psi)}, \]  
\[ P_2 = \text{downstream pressure (psi)}, \]  
\[ Z_1 = \text{upstream elevation (ft.)}, \]  
\[ Z_2 = \text{downstream elevation (ft.)}, \]  
\[ H_L = \text{frictional head loss (ft.)}, \]  
\[ \rho = \text{fluid density (lbf/ft.}^3\text{)}, \]  
\[ g = \text{acceleration of gravity (32.17 ft./s}^2\text{)}. \]

Figure 2.1 illustrates the condition for flow and pressure drop equations that will be presented in the next section.

Velocity, as used herein, refers to the average velocity of a fluid at a given cross section and is determined by the steady-state flow equation:

\[ V = \frac{Q}{A} = \frac{W_\text{s}}{A_p} \quad \text{Eq. 2.3} \]

where
\[ V = \text{average velocity (ft./s)}, \]  
\[ Q = \text{rate of flow (ft.}^3\text{/s)}, \]  
\[ A_p = \text{cross-sectional area of the pipe (ft.}^2\text{)}, \]  
\[ W_\text{s} = \text{rate of flow (lb/s)}. \]

2.3 Pressure drop equations

A lot of attention has been given to pressure drop calculations throughout the years, and the result is that there are several good methods to compute pressure drops.
Bernoulli’s equation has the frictional effects included in the $H_f$ term, and the method for calculating frictional effects is unspecified. The Darcy-Weisbach and Fanning equations provide a means of calculating these friction effects.

2.3.1 Darcy-Weisbach equation

The head loss due to friction is given by the Darcy-Weisbach equation as follows:

$$H_f = f_m L V^2 / D 2g \quad \text{Eq. 2.4}$$

The velocity of the fluid is understood to be the average velocity and is best described as the volumetric flow rate divided by the cross-sectional wetted area of the pipe.

2.3.2 Fanning equation

The head loss due to friction is given by the Fanning equation as follows:

$$H_f = 4f_f L V^2 / D 2g \quad \text{Eq. 2.5}$$

where
- $H_f$ = frictional head loss (ft.),
- $f_m$ = Moody friction factor, dimensionless, (Darcy friction factor “f” or “$f_m$”)
- $f_f$ = Fanning friction factor, dimensionless,
- L = pipe length (ft.),
- D = pipe diameter (ft.),
- $V$ = fluid velocity (ft./s),
- $g$ = acceleration of gravity (32.17 ft./s (9.81 m/s²)).

Equations 2.2 and 2.4 or 2.5 can be used to calculate the pressure at any point in a piping system if the pressure, average flow velocity, pipe inside diameter, and elevation are known at any other point. If the pressures, pipe inside diameter, and elevations are known at two points, the flow velocity can be calculated. Neglecting the head differences due to elevation and velocity changes between the two points, Equation 2.2 can be reduced to

$$P_1 - P_2 = \Delta P = \rho H_f / 144 \quad \text{Eq. 2.6}$$

where
- $\Delta P$ = pressure drop (psi),
- $H_f$ = pipe friction head loss (ft.),
- $\rho$ = density of liquid (lb/ft.

Substituting Equation 2.4 to Equation 2.6 and expressing pipe inside diameter in inches,

$$\Delta P = 0.0013 f \rho L V^2 / d \quad \text{Eq. 2.7}$$

Where
- $d$ = pipe inside diameter (in.),
- $f$ = Moody friction factor, dimensionless,
- $\rho$ = density of liquid (lb/ft.³),

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L = length of pipe (ft.),
V = average velocity (ft./s),
ΔP = pressure drop (psi).

2.3.3 Moody and fanning friction factor comparison

The Darcy-Weisbach equation can be derived by dimensional analysis, except for the friction factor (f), which must be determined experimentally. There has been a considerable amount of research in reference to pipe roughness and friction factors. The Moody friction factor is generally accepted and used in pressure drop calculations.

Some texts, including API RP 14E, use the “Fanning” friction factor, which is one fourth the value of the Moody friction factor.

Confusion between Moody and Fanning friction factors is a common source of error. In this text, the Moody friction factor will be used exclusively. Many references are not clear as to which friction factor is being used. Therefore, the reader is cautioned to always note which friction factor, Moody or Fanning, is used in the applicable equations and which friction factor diagram is used as a source when calculating pressure drops.

Comparison of the two pressure drop equations reveals that the Darcy-Weisbach and Fanning equations differ only by a factor of 4. This occurs because of the differences in friction factors:

\[ f_{\text{Fanning}} = f / 4 \quad \text{Eq. 2.8} \]

The friction factor for fluids in laminar flow is directly related to the Reynolds number (Re<2000) and is expressed:

\[ f = \frac{64}{Re} = 0.52 \frac{\mu}{d V \rho} \quad \text{Eq. 2.9} \]

Where
- f = Moody friction factor, “Darcy friction factor, f, or f_m”
- \( Re \) = the Reynolds number,
- \( \mu \) = absolute viscosity (cP),
- d = pipe inside diameter (in.),
- V = average velocity (ft./s),
- \( \rho \) = density of fluid (lb/ft.\(^3\)).

If this quantity is substituted into Equation 2.7, pressure drop in psi for fluids in laminar flow becomes

\[ \Delta P = 0.000676 \frac{\mu LV}{d^2} \quad \text{Eq. 2.10} \]

The friction factor for fluids in turbulent flow (Re>4000) depends on the Reynolds number and the relative roughness of the pipe. Relative roughness is the ratio of the pipe absolute roughness, \( \varepsilon \), to pipe inside diameter. Roughness is a measure of the smoothness of the pipe’s inner surface. Table 2.1 shows the absolute roughness, \( \varepsilon \), for various types of new, clean pipe. For pipe that has been in service for some time, it is often recommended that the absolute roughness to be used for calculations should be up to four times as much as the values in Table 2.1.

For the Reynolds numbers less than 2000 (laminar flow):
- Characterized by little mixing of the flowing fluids and a parabolic velocity profile
• All the friction factor curves collapse onto a straight line

For the Reynolds number greater than 4000 (turbulent flow):
• Characterized by complete mixing of the flowing fluid and a more uniform velocity profile.
• Friction factor curves are widely spread and almost horizontal, indicating that the friction factor is almost totally a function of relative roughness.

For the Reynolds numbers between 2000 and 4000 (transition flow):
• Unstable region between laminar and turbulent flow.
• Friction factor is undetermined.

The friction factor, f, can be determined from the Moody diagram, Figure 2.2.
The pressure drop between any two points in a piping system can be determined from Equation 2.10 for laminar flow or Equation 2.7 for turbulent flow using the friction factor from Figure 2.2.

<table>
<thead>
<tr>
<th>Pipe roughness</th>
</tr>
</thead>
<tbody>
<tr>
<td>Type of pipe</td>
</tr>
<tr>
<td>New, clean condition</td>
</tr>
<tr>
<td>Unlined concrete</td>
</tr>
<tr>
<td>Cast iron—uncoated</td>
</tr>
<tr>
<td>Galvanized iron</td>
</tr>
<tr>
<td>Carbon steel</td>
</tr>
<tr>
<td>Fiberglass epoxy</td>
</tr>
</tbody>
</table>

Table 2.1 Absolute roughness “ε” for various types of new, clean pipe

2.3.4 Calculation procedure
Once the pipe inside diameter and material have been selected, the following procedure should be followed:
• Determine the relative roughness from Table 2.1.
• Calculate the Reynolds number using the design flow rate, fluid properties, and pipe diameter.
• Using the calculated Reynolds number and determined relative roughness, determine the Moody friction factor from Figure 2.2. (Or the friction factor can be calculated using Colebrook-White, or AGA equation which will be presented later).
• This friction factor is used in the Darcy-Weisbach equation to calculate frictional head loss.
• This frictional head loss is inserted into the momentum equation to determine the overall pressure drop.
Figure 2.2. Friction factor as a function of the Reynolds number and pipe roughness. Courtesy of API.

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2.3.5 Effects of elevation changes
In single-phase gas or liquid flow, the pressure change between two points in a line will be affected by the relative elevations of those points, but not by intermediate elevation changes. The reason for this is because the density of the flowing fluid is nearly constant, and the pressure increase caused by any decrease in elevations is balanced by the pressure decrease caused by an identical increase in elevation.
Thus, for a single-phase flow, the pressure drop due to elevation changes is determined solely by the elevation change of the end points.

\[ \Delta P_z = 0.433 \ (SG) \Delta Z \quad \text{Eq. 2.11} \]

where
\[ \Delta P_z = \text{pressure drop due to elevation increase (psi)} \]
\[ \Delta Z = \text{total increase in elevation (ft.)} \]
\[ SG = \text{specific gravity relative to water} \]

In two-phase flow, the density of the fluids in the uphill runs is higher than the density of the fluids in the downhill runs. In downhill lines, flow is stratified with liquid flowing faster than gas. The depth of the liquid layer adjusts to the depth where the static head advantage is equal to the pressure drop due to friction, and thus, the average density of the mixture approaches that of the gas phase.
The uphill segments at low gas rates are liquid full, and the density of the mixture approaches that of the liquid phase. As a worst-case condition, it can be assumed that the downhill segments are filled with gas and the uphill segments are filled with liquid.
Thus, the maximum pressure drop due to elevation changes in two-phase lines can be estimated from

\[ \Delta P_z \leq 0.433 \ (SG) \sum Z_e \quad \text{Eq. 2.12} \]

where
\[ \sum Z_e = \text{sum of vertical elevation rises only (ft.)} \]
\[ \Delta P_z = \text{pressure drop due to elevation changes (psi)} \]
\[ (SG) = \text{specific gravity of liquid relative to water} \]

With increasing gas flow, the total pressure drop decrease as liquid is removed from uphill segments. More accurate prediction of the pressure drop due to elevation changes requires complete two-phase flow models that are beyond the scope of this text. There are a number of proprietary computer programs available that take into account fluid property changes and liquid holdup in small line segments; they model pressure drop due to elevation changes in two-phase flow more accurately.

2.3.6 Effects of viscosity
As viscosity increases, the Reynolds number decreases, and the frictional factor becomes more sensitive to changes in viscosity.
Accurate determination of viscosity is often difficult, especially for heavy liquids.
Since liquid viscosities tend to increase exponentially as temperature decreases, accurate knowledge of temperature profiles is necessary. If data are important for facility design, the liquid can be analyzed in a laboratory. If not critical, data of a similar liquid can be used or data from Figures 1.4, and 1.5, can be used.
The effective viscosity of oil and water is also difficult to determine. Oil and water may or may not form an emulsion, depending on the surface tension characteristics of the oil and the presence of
an emulsifying agent. Emulsifying tendencies are dependent on liquid composition. If an emulsion
does not form, the water merely acts as a diluent, reducing the effective viscosity of the oil in
proportion to the relative volumes of oil and water present. If an emulsion is formed, the emulsion
viscosity can be 10-50 times higher than the dewatered oil viscosity.
At water cuts from 40% to 80%, highly viscous oil emulsion will form an inverse (water-external)
emulsion. When this inverse emulsion occurs, the effective liquid viscosity falls from about 50
times the dewatered oil viscosity to the water viscosity. Thus, the pressure drop in a pipe is thus
extremely difficult to predict. Not even laboratory measurements of viscosities of field oil-water
mixtures will determine the mixture viscosity since emulsion viscosities are dependent on shear
history. The shear history in the pipeline is virtually impossible to recreate in the laboratory. (Refer
to paragraph 1.5.1.5.)

2.3.7 Hazen-Williams equation
Using the general equation for calculating pressure drop-flow rate relations in liquid lines is
troublesome and time-consuming. For approximate hand calculations used for screening studies
and preliminary cost estimates, a simpler and quicker method may be used. The Hazen-Williams
equation is such a method, which is expressed as:

\[
H_L = 0.00208 \left(\frac{100}{C}\right)^{1.85} \left(\frac{\text{gpm}}{d^{2.63}}\right)^{1.85} L \quad \text{Eq. 2.13}
\]

or

\[
H_L = 0.015 \left(\frac{Q_L^{1.85}}{d^{4.87}} C^{1.85}\right) \quad \text{Eq. 2.14}
\]

Or

\[
H_L = 79.309 Q_L^{1.85} L_m^{1.85} / d^{4.87} C^{1.85} \quad \text{Eq. 2.15}
\]

where
\[
H_L = \text{frictional head loss (ft.),}
Q_L = \text{fluid flow rate (BPD),}
\text{Gpm = fluid flow rate (gpm),}
L = \text{pipeline length (ft.),}
L_m = \text{pipeline length (miles),}
d = \text{pipeline inside diameter (in.),}
C = \text{Hazen-Williams “C” factors, dimensionless,}
= 140 \text{ for new steel pipe,}
= 130 \text{ for new cast iron pipe,}
= 100 \text{ for riveted pipe,}
= 90 \text{ for old steel pipe.}
\]

The Hazen-Williams equation is based on the following assumptions:
\begin{itemize}
\item Fluid is water at 60 °F.
\item Flow regime is turbulent.
\item C and f_D are not directly related.
\end{itemize}

Once the frictional head loss is calculated using the Hazen-Williams equation, the head loss can
be inserted into the momentum equation to determine pressure drop.
The Hazen-Williams equation is easily inverted so that flow rate can be determined if frictional
head loss is known.
In its inverted form, the Hazen-Williams equation is

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\[ Q_L = 0.094286 \, Cd^{2.63} \, (H_L/L_m)^{0.54} \]  
Eq. 2.16

where
\( Q_L \) = fluid flow rate (BPD),
\( d \) = pipeline inside diameter (in.),
\( H_L \) = frictional head loss (in.),
\( L_m \) = pipeline length (miles),
\( C \) = Hazen-Williams "C" factor, dimensionless.

\[ \Delta P = SG \times (62.4) \times \frac{H_L}{144} \]  
Eq. 2.17

Where
\( \Delta P \) = pressure drop, psi

2.3.7.1 “C” factors
With all the precision and sophistication of the friction factor, the Reynolds number plot (Figure 2.2) has been collected into the single Hazen-Williams “C” factor. A tabulation of recommended “C” factors is included in Table 2.2. The spread in recommended “C” factors indicates that the Hazen-Williams equation is a fairly approximate method. The best way to use the equation is with “C” factors that have been empirically determined by field experience in conditions similar to those under consideration. Since the effects of relative roughness are included in the “C” factor, field data from pipe in similar service and condition to that being calculated are desirable.

2.3.7.2 Applicability
Do not use the Hazen-Williams equation for any pressure drop-flow rate calculations having odd or unusual characteristics. Do not use the equation for viscous fluids. Recommended “C” factors do not take into account the effect of a low Reynolds number and are particularly inappropriate for laminar flows. Do not use the equation for pipe in poor conditions. Recommended “C” factors do not account for the unusually high roughness ratio of deteriorated pipe. If in doubt, use the general equation. This method is much more cumbersome but it is more versatile and more accurate.

<table>
<thead>
<tr>
<th>Type of pipe</th>
<th>“C” values</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Range*</td>
</tr>
<tr>
<td>Welded and seamless steel</td>
<td>150-80</td>
</tr>
<tr>
<td>Cement asbestos</td>
<td>160-140</td>
</tr>
<tr>
<td>Fiber</td>
<td>-</td>
</tr>
<tr>
<td>Cement-lined iron</td>
<td>-</td>
</tr>
<tr>
<td>Bitumastic-enamel-lined iron</td>
<td>160-130</td>
</tr>
<tr>
<td>Copper, brass, lead, tin, or glass</td>
<td>150-120</td>
</tr>
</tbody>
</table>

Remarks.
* Range, high best, smooth, well laid; low-poor or corroded.
* Average, value for good, clean, new pipe.
* Common, used value for design purposes

Table 2.2. Hazen-Williams “C” factors.

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2.4 Gas flow
Various formulas were developed over the years to predict the performance of a pipeline transporting gas. These formulas are intended to show the relationship between the gas properties, such as gravity and compressibility factor, with the flow rate, pipe diameter and length, and the pressures along the pipeline.
Thus, for a given pipe size and length, we can predict the flow rate possible through a pipeline based upon an inlet pressure and an outlet pressure of a pipe segment. Simplifications are sometimes introduced, such as uniform gas temperature and no heat transfer between the gas and the surrounding soil in a buried pipeline, in order to adopt these equations for manual calculations. With the advent of microcomputers, we are able to introduce heat transfer effects and, therefore, more accurately model gas pipelines, taking into consideration gas flow temperatures, soil temperatures, and thermal conductivities of pipe material, insulation, and soil.
In this chapter we will concentrate on steady-state isothermal flow of gas in pipelines.

2.4.1 Gas Flow Equations
Several equations are available that relate the gas flow rate with gas properties, pipe diameter and length, and upstream and downstream pressures. These equations are listed as follows:
1. General Flow equation
2. Colebrook-White equation
3. Modified Colebrook-White equation
4. AGA equation
5. Oliphant equation
6. Spitzglass equation
7. Weymouth equation
8. Panhandle A equation
9. Panhandle B equation
10. IGT equation

We will discuss each of these equations, their limitations, and their applicability to compressible fluids, such as natural gas, using illustrated examples. A comparison of these equations will also be discussed using an example pipeline.

2.4.1.1 General Flow Equation
The General Flow equation, also called the Fundamental Flow equation, for the steady-state isothermal flow in a gas pipeline is the basic equation for relating the pressure drop with flow rate. The most common form of this equation in the U.S. Customary System (USCS) of units is given in terms of the pipe diameter, gas properties, pressures, temperatures, and flow rate as follows.
Refer to Figure 2.1 for an explanation of symbols used.

\[ Q = 77.54 \left( \frac{T_0}{P_0} \right) d^{2.5} \left( \frac{(P_1^2 - P_2^2)}{(G T_f L_m Z_a f)} \right)^{0.5} \]

Eq. 2.18

where
- \( Q \) = gas flow rate, measured at standard conditions, ft\(^3\)/day (SCFD)
- \( f \) = friction factor, “Darcy from Moody diagram”, dimensionless
- \( P_0 \) = base pressure, psia
- \( T_0 \) = base temperature, °R (460 + °F)
- \( P_1 \) = upstream pressure, psia
- \( P_2 \) = downstream pressure, psia

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G = gas gravity (air = 1.00)
T_f = average gas flowing temperature, °R (460 + °F)
L_m = pipe segment length, mile
Z = gas compressibility factor at the flowing temperature, dimensionless
d = pipe inside diameter, in.

It must be noted that for the pipe segment from section 1 to section 2, the gas temperature \( T_f \) is assumed to be constant (isothermal flow).

Equation 2.20 relates the capacity (flow rate or throughput) of a pipe segment of length \( L \), based on an upstream pressure of \( P_1 \) and a downstream pressure of \( P_2 \) as shown in Figure 2.3. It is assumed that there is no elevation difference between the upstream and downstream points; therefore, the pipe segment is horizontal.

Upon examining the General Flow Equation 2.18, we see that for a pipe segment of length \( L \) and diameter \( d \), the gas flow rate \( Q \) (at standard conditions) depends on several factors. \( Q \) depends on gas properties represented by the gravity \( G \) and the compressibility factor \( Z \). If the gas gravity is increased (heavier gas), the flow rate will decrease. Similarly, as the compressibility factor \( Z \) increases, the flow rate will decrease. Also, as the gas flowing temperature \( T_f \) increases, throughput will decrease.

Thus, the hotter the gas, the lower the flow rate will be. Therefore, to increase the flow rate, it helps to keep the gas temperature low. The impact of pipe length and inside diameter is also clear. As the pipe segment length increases for given pressure \( P_1 \) and \( P_2 \), the flow rate will decrease. On the other hand, the larger the diameter, the larger the flow rate will be. The term \( (P_1^2 - P_2^2) \) represents the driving force that causes the flow rate from the upstream end to the downstream end. As the downstream pressure \( P_2 \) is reduced, keeping the upstream pressure \( P_1 \) constant, the flow rate will increase. It is obvious that when there is no flow rate, \( P_1 \) is equal to \( P_2 \). It is due to friction between the gas and pipe walls that the pressure drop \( (P_1 - P_2) \) occurs from the upstream point 1 to downstream point 2. The friction factor \( f \) depends on the internal condition of the pipe as well as the type of flow (laminar or turbulent) and will be discussed in detail.

Sometimes the General Flow equation is represented in terms of the transmission factor \( F \) instead of the friction factor \( f \). This form of the equation is as follows.

\[
Q = 38.77 \ F \ (T_0/P_0) \ d^{2.5} \ [(P_1^2 - e^4P_2^2) \ / \ (G \ T_f \ L \ e \ Z_a)]^{0.5}
\]

Eq. 2.19

where

\( F \) = the transmission factor \( F \) and friction factor \( f \) are related by
F = 2/(f)\(^{0.5}\)

L\(_e\) = Equivalent length, mile.

\[
L_e = L \left( e^s - 1 \right) /s \quad \text{Eq. 2.20}
\]

s = elevation adjustment parameter, dimensionless

\(\Delta H\) = elevation difference, ft

\(e\) = base of natural logarithms ( \(e = 2.718\ldots\) )

\[
s = 0.0375 \, G \left( \Delta H / T_f Z_a \right) \quad \text{Eq. 2.21}
\]

In the calculation of \(L_e\) in Equation 2.20, we have assumed that there is a single slope between the upstream point 1 and the downstream point 2 in Figure 2.3. If, however, the pipe segment of length \(L\) has a series of slopes, then we introduce a parameter \(j\) as follows for each individual pipe sub-segment that constitutes the pipe length from point 1 to point 2.

\[
J = (e^s - 1) /s \quad \text{Eq. 2.22}
\]

The parameter \(j\) is calculated for each slope of each pipe sub-segment of length \(L_1, L_2, \text{ etc.}\) that make up the total length \(L\). The equivalent length term \(L_e\) in Equation 2.21 is calculated by summing the individual slopes as defined below.

\[
L_e = j_1 L_1 \, s^1 + j_2 L_2 \, e^{s2} + j_3 L_3 \, e^{s3} + \ldots \quad \text{Eq. 2.23}
\]

The terms \(j_1, j_2, \text{ etc.}\) for each rise or fall in the elevations of individual pipe Sub-segments are calculated for the parameters \(s^1, s^2, \text{ etc.}\) for each segment in accordance with Equation 2.22, from the pipeline inlet to the end of each segment.

In the subsequent sections of this chapter, we will discuss how the friction factor and transmission factor are calculated using various equations such as Colebrook-White and AGA. It is important to note that the General Flow equation is the most commonly used equation to calculate the flow rate and pressure in a gas pipeline.

In order to apply it correctly, we must use the correct friction factor or transmission factor. The Colebrook equation, AGA equation, and other empirical equations are used to calculate the friction factor to be used in the General Flow equation. Several other equations, such as Panhandle A, Panhandle B, and Weymouth, calculate the flow rate for a given pressure without using a friction factor or transmission factor.

However, an equivalent friction factor (or transmission factor) can be calculated using these methods as well.

### 2.4.1.2 Friction Factor

In order to calculate the pressure drop in a pipeline at a given flow rate, we must first understand the concept of friction factor. The term friction factor is a dimensionless parameter that depends upon the Reynolds number of flow. In engineering literature, we find two different friction factors mentioned. The Darcy friction factor is more common and will be used throughout this book.

Another friction factor known as, the Fanning friction factor, which is preferred by some engineers. The Fanning friction factor is numerically equal to one-fourth the Darcy friction factor "Moody friction factor" as below.


\[ f_t = \frac{f_m}{4} \quad \text{or} \quad f_t = \frac{f}{4} \quad \text{Eq. 2.24} \]

where

\( f_t \) = Fanning friction factor

\( f_m \) or \( f_d \) or \( f \) = Darcy friction factor "Moody"

To avoid confusion, in subsequent discussions, the Darcy friction factor is used and will be represented by the symbol \( f \).

For laminar flow, the friction factor is inversely proportional to the Reynolds number, as indicated below.

\[ f = \frac{64}{Re} \quad \text{Eq. 2.25} \]

For turbulent flow, the friction factor is a function of the Reynolds number, pipe inside diameter, and internal roughness of the pipe. Many empirical relationships for calculating “\( f \)” have been put forth by researchers. The more popular correlations include the Colebrook-White and AGA equations.

Before we discuss the equations for calculating the friction factor in turbulent flow, it is appropriate to analyze the turbulent flow regime. Turbulent flow in pipes \((Re > 4000)\) is subdivided into three separate regions as follows:

1. Turbulent flow in smooth pipes
2. Turbulent flow in fully rough pipes
3. Transition flow between smooth pipes and rough pipes

For turbulent flow in smooth pipes, the friction factor \( f \) depends only on the Reynolds number. For fully rough pipes, \( f \) depends more on the pipe internal roughness and less on the Reynolds number. In the transition zone between smooth pipe flow and flow in fully rough pipes, \( f \) depends on the pipe roughness, pipe inside diameter, and the Reynolds number. The various flow regimes are depicted in the Moody diagram, shown in Figures 2.4, and 2.2.

The Moody diagram is a graphic plot of the variation of the friction factor with the Reynolds number for various values of relative pipe roughness. The latter term is simply a dimensionless parameter obtained by dividing the absolute (or internal) pipe roughness by the pipe inside diameter as follows:

\[ \text{Relative roughness} = \frac{\varepsilon}{d} \quad \text{Eq. 2.26} \]

where

\( \varepsilon \) = absolute or internal roughness of pipe, in.

\( d \) = pipe inside diameter, in.

The terms absolute pipe roughness and internal pipe roughness are equivalent. Generally, the internal pipe roughness is expressed in microinches (one-millionth of an inch). For example, an internal roughness of 0.0006 in. is referred to as 600 microinches or 600 u in. If the pipe inside diameter is 15.5 in., the relative roughness is, in this case,

Relative roughness \( = 0.0006 / 15.5 = 0.0000387 = 3.87 \times 10^{-5} \)

For example, from the Moody diagram in Figure 2.4, for \( Re = 10 \) million and \( \varepsilon/D = 0.0001 \), we find that \( f = 0.012 \).
2.4.1.3 Colebrook-White Equation

The Colebrook-White equation, sometimes referred to simply as the Colebrook equation, is a relationship between the friction factor and the Reynolds number, pipe roughness, and inside diameter of pipe. The following form of the Colebrook equation is used to calculate the friction factor in gas pipelines in turbulent flow.

\[
\frac{1}{f^{0.5}} = -\log_{10} \left[ \left( \frac{\epsilon}{3.7d} \right) + \left( \frac{2.51}{Re f^{0.5}} \right) \right] \quad \text{for } Re > 4000 \quad \text{Eq. 2.27}
\]

where
- \( f \) = friction factor, dimensionless
- \( d \) = pipe inside diameter, in.
- \( \epsilon \) = absolute pipe roughness, in.
- \( Re \) = Reynolds number of flow, dimensionless

It can be seen from Equation above 2.27 that in order to calculate the friction factor \( f \), we must use a trial-and-error approach. It is an implicit equation in \( f \), since \( f \) appears on both sides of the equation. We first assume a value of \( f \) (such as 0.01) and substitute it in the right-hand side of the equation. This will yield a second approximation for \( f \), which can then be used to calculate a better value of \( f \), and so on. Generally, 3 to 4 iterations are sufficient to converge on a reasonably good value of the friction factor.
It can be seen from the Colebrook Equation 2.27, for turbulent flow in smooth pipes, the first term within the square brackets \((\varepsilon/3.7d)\) is negligible compared to the second term, since pipe roughness \(\varepsilon\) is very small. Therefore, for smooth pipe flow, the friction factor equation reduces to

\[
1/(f)^{0.5} = -2\log_{10} (2.51/ Re f^{0.5}) \quad \text{for turbulent flow in smooth pipes} \quad \text{Eq. 2.28}
\]

Similarly, for turbulent flow in fully rough pipes, with \(Re\) being a large number, "\(f\) " depends mostly on the roughness \(\varepsilon\) and, therefore, the friction factor equation reduces to

\[
1/(f)^{0.5} = -2\log_{10} (\varepsilon/3.7d) \quad \text{for turbulent flow in fully rough pipes} \quad \text{Eq. 2.29}
\]

Table 2.3 lists typical values of pipe internal roughness used to calculate the friction factor. or, \(\varepsilon/d\) = relative roughness can be determined from Figure 2.5
As an example, if \(Re = 100\) million or larger and \(\varepsilon /d = 0.0002\), the friction factor from Equation 2.31 is

\[
1/(f)^{0.5} = 2\log_{10} (0.0002/3.7) = 0.0137,
\]

which correlates well with the friction factor obtained from the Moody diagram in Figure 2.4 or 2.2.

<table>
<thead>
<tr>
<th>Pipe Material</th>
<th>Roughness, in.</th>
<th>Pipe Material</th>
<th>Roughness, in.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Riveted steel</td>
<td>0.0354 to 0.354</td>
<td>Commercial steel/welded steel</td>
<td>0.0018</td>
</tr>
<tr>
<td>Cast iron</td>
<td>0.0102</td>
<td>Galvanized iron</td>
<td>0.0059</td>
</tr>
<tr>
<td>Asphalated cast iron</td>
<td>0.0047</td>
<td>Wrought iron</td>
<td>0.0018</td>
</tr>
<tr>
<td>PVC, drawn tubing, glass</td>
<td>0.000059</td>
<td>Concrete</td>
<td>0.0118 to 0.118</td>
</tr>
</tbody>
</table>

Table 2.3 Pipe Internal Roughness,

Example 2.1
A natural gas pipeline, NPS 20 with 0.500 in. wall thickness, transports 200 MMscfd. The specific gravity of gas is 0.6 and viscosity is 0.000008 lb/ft-s. Calculate the friction factor using the Colebrook equation. Assume absolute pipe roughness = 600 \(\mu\) in.

Solution
Pipe inside diameter = 20 \(- (2 x 0.5) = 19.0\) in.
Absolute pipe roughness = 600 \(\mu\) in. = 0.0006 in.

First, we calculate the Reynolds number using Equation 1.34:

\[
Re = 1.35 \times 10^{-5} \frac{(G Q)}{\mu d} \quad \text{Eq. 1.34}
\]

\[
Re = 1.35 \times 10^{-5} \frac{(G \times 200 \times 10^6/0.000008 \times 19)}{10,663,452}
\]

Using Equation 2.27,

\[
1/(f)^{0.5} = -2\log_{10} \left[ (\varepsilon/3.7) + (2.51/ Re f^{0.5}) \right]
\]

This equation will be solved by successive iteration.
Assume \(f = 0.01\) initially

\[
1/(f)^{0.5} = -2\log_{10} \left[ (0.0006/3.7 \times 19) + (2.51/ 10663 \times (0.01)^{0.5}) \right]
\]

We get a better approximation as \(f = 0.0101\). Repeating the iteration, we get the final value as \(f = 0.0101\). Therefore, the friction factor is 0.0101.
2.4.1.4 Transmission Factor
The transmission factor "F" is considered the opposite of the friction factor "f". Whereas the friction factor indicates how difficult it is to move a certain quantity of gas through a pipeline, the transmission factor is a direct measure of how much gas can be transported through the pipeline. As the friction factor increases, the transmission factor decreases and, therefore, the gas flow...
rate also decreases. Conversely, the higher the transmission factor, the lower the friction factor and, therefore, the higher the flow rate will be.

The transmission factor $F$ is related to the friction factor $f$ as follows:

$$ F = 2 / (f)^{0.5} \quad \text{Eq. 2.30} $$

Therefore

$$ f = 4 / F^2 \quad \text{Eq. 2.31} $$

where

- $f$ = friction factor
- $F$ = transmission factor

It must be noted that the friction factor $f$ in the above equation is the Darcy friction factor. Since some engineers prefer to use the Fanning friction factor, the relationship between the transmission factor $F$ and the Fanning friction factor is given below for reference.

$$ F = 1 / (f_f)^{0.5} \quad \text{Eq. 2.32} $$

where $f_f$ is the Fanning friction factor.

For example, if the Darcy friction factor is 0.025, the transmission factor $F = 2 / (0.025)^{0.5} = 12.65$

The Fanning friction factor in this case will be $0.025/4 = 0.00625$. (Eqn. 2.24), Therefore, the transmission factor using Equation 2.32 is $1/(0.00625)^{0.5} = 12.65$, which is the same as calculated using the Darcy friction factor.

Thus, it must be noted that there is only one transmission factor, whereas there are two different friction factors.

Having defined a transmission factor, we can rewrite the Colebrook Equation 2.27 in terms of the transmission factor using Equation 2.30 as follows:

$$ F = -4\log_{10} \left[ (\varepsilon/3.7d) + (1.255 F / Re) \right] \quad \text{Eq. 2.33} $$

Similar to the calculation of the friction factor $f$ from Equation 2.27, to calculate the transmission factor $F$ from Equation 2.33, an iterative approach must be used.

This will be illustrated using an example.

Example 2.2

For a gas pipeline, flowing 100 MMscfd gas of specific gravity 0.6 and viscosity of 0.000008 lb/ft-s, calculate the friction factor and transmission factor considering an NPS 20 pipeline, 0.500-in. wall thickness, and an internal roughness of 600 microinches, calculate the friction factor and transmission factor?

Solution

we calculate the Reynolds number as

$$ Re = 1.35 \times 10^{-5} (GQ/\mu \cdot d) \quad \text{Eq. 1.34} $$

$$ Re = 1.35 \times 10^{-5} (0.6 \times 100 \times 10^6/0.000008 \times 19) $$

$$ Re = 5,331,726 $$

$$ 1/(f)^{0.5} = -2\log_{10} \left[ (\varepsilon/3.7d) + (2.51/ \text{Re f}^{0.5}) \right] $$

$$ 1/(f)^{0.5} = -2\log_{10} \left[ (0.0006/3.7 \times 19) + (2.51/ 5331726 \times f^{0.5}) \right] $$

Books are available [here](http://www.oilprocessing.net/oil/)
By solving we will get
\[ f = 0.0105 \]
\[ F = \frac{2}{(0.0105)^{0.5}} = 19.53 \]

2.4.1.5 Modified Colebrook-White Equation
The Colebrook-White equation discussed in the preceding section has been in use for many years in both liquid flow and gas flow. The U.S. Bureau of Mines, in 1956, published a report that introduced a modified form of the Colebrook-White equation.

The modification results in a higher friction factor and, hence, a smaller value of the transmission factor. Because of this, a conservative value of flow rate is obtained due to the higher friction and pressure drop. The modified version of the Colebrook-White equation for turbulent flow is as follows:

\[ \frac{1}{(f)^{0.5}} = -2 \log_{10} \left( \frac{\varepsilon}{3.7d} + \frac{2.825}{Re f^{0.5}} \right) \]  
Eq. 2.34

Rewriting Equation 2.34 in terms of the transmission factor, we get the following version of the modified Colebrook-White equation:

\[ F = -4 \log_{10} \left( \frac{\varepsilon}{3.7d} + \frac{1.4125 F}{Re} \right) \]  
Eq. 2.35

It can be seen that the difference between the Colebrook equation and the modified Colebrook equation lies in the second constant term within the square brackets. The constant 2.51 in Equation 2.27 is replaced with the constant 2.825 in Equation 2.34. Similarly, in the transmission factor equations, the modified equation has 1.4125 instead of 1.255 in the original Colebrook-White equation. Many commercial hydraulic simulation programs list both Colebrook-White equations. Some use only the original Colebrook-White equation.

Example 2.3
For a gas pipeline, flowing 100 MMSCFD gas of specific gravity 0.6 and viscosity of 0.000008 lb/ft-s, calculate, using the modified Colebrook-White equation, the friction factor and transmission factor assuming an NPS 20 pipeline, 0.500 in. wall thickness, and an internal roughness of 600 μ in. How do these numbers compare with those calculated, using the original Colebrook equation?

Solution
Pipe inside diameter = 20 – (2 x 0.500) = 19.0 in.
\[ Re = 1.35 \times 10^5 \ (\text{GQ}/\mu \ \text{d}) \]  
Eq. 1.34

Reynolds number = 5,331,726
Using the modified Colebrook equation
\[ \frac{1}{(f)^{0.5}} = -2 \log_{10} \left( \frac{\varepsilon}{3.7d} + \frac{2.825}{Re f^{0.5}} \right) \]
Solving by successive iteration, we get
\[ f = 0.0106 \]

Therefore, the transmission factor \[ F = \frac{2}{(0.0106)^{0.5}} = 19.43 \]

By comparing these results with the friction factor and the transmission factor calculated in Example 2.2 using the unmodified Colebrook equation, it can be seen that the modified friction factor is approximately 1% higher than that calculated using the original Colebrook-White equation, whereas the transmission factor is approximately 0.5% lower than that calculated using the original Colebrook-White equation.

http://www.oilprocessing.net/oil/
2.4.1.6 American Gas Association (AgA) Equation
In 1964 and 1965, the American Gas Association (AGA) published a report on how to calculate the transmission factor for gas pipelines to be used in the General Flow equation. This is sometimes referred to as the AGA NB-13 method. Using the method outlined in this report, the transmission factor \( F \) is calculated using two different equations. First, \( F \) is calculated for the rough pipe law (referred to as the fully turbulent zone). Next, \( F \) is calculated based on the smooth pipe law (referred to as the partially turbulent zone). Finally, the smaller of the two values of the transmission factor is used in the General Flow Equation 2.18, for pressure drop calculation. Even though the AGA method uses the transmission factor \( F \) instead of the friction factor \( f \), we can still calculate the friction factor using the relationship shown in Equation 2.19.

For the fully turbulent zone, AGA recommends using the following formula for \( F \), based on relative roughness \( \varepsilon/D \) and independent of the Reynolds number:

\[
F = 4\log_{10}(3.7d/\varepsilon) \quad \text{Eq. 2.36}
\]

Equation 2.36 is also known as the Von Karman rough pipe flow equation. For the partially turbulent zone, \( F \) is calculated from the following equations using the Reynolds number, a parameter \( D_f \) known as the pipe drag factor, and the Von Karman smooth pipe transmission factor \( F_t \):

\[
F = 4D_f \log_{10}(Re/1.4125 F_t) \quad \text{Eq. 2.37}
\]

And

\[
F_t = 4\log_{10}(Re/F_t) - 0.6 \quad \text{Eq. 2.38}
\]

where
- \( F_t \) = Von Karman smooth pipe transmission factor
- \( D_f \) = pipe drag factor that depends on the Bend Index (BI) of the pipe

The pipe drag factor \( D_f \) is a parameter that takes into account the number of bends, fittings, etc. Its value ranges from 0.90 to 0.99. The Bend index is the sum of all the angles and bends in the pipe segment, divided by the total length of the pipe section under consideration.

\( BI = \frac{\text{total degrees of all bends in pipe section}}{\text{total length of pipe section}} \)

The value of \( D_f \) is generally chosen from Table 2.4.

<table>
<thead>
<tr>
<th>Bend Index</th>
<th>Extremely Low 5° to 10°</th>
<th>Average 60° to 80°</th>
<th>Extremely High 200° to 300°</th>
</tr>
</thead>
<tbody>
<tr>
<td>Bare steel</td>
<td>0.975–0.973</td>
<td>0.960–0.956</td>
<td>0.930–0.900</td>
</tr>
<tr>
<td>Plastic lined</td>
<td>0.979–0.976</td>
<td>0.964–0.960</td>
<td>0.936–0.910</td>
</tr>
<tr>
<td>Pig burnished</td>
<td>0.982–0.980</td>
<td>0.968–0.965</td>
<td>0.944–0.920</td>
</tr>
<tr>
<td>Sand blasted</td>
<td>0.985–0.983</td>
<td>0.976–0.970</td>
<td>0.951–0.930</td>
</tr>
</tbody>
</table>

Note: The drag factors above are based on 40-ft joints of pipelines and mainline valves at 10-mile spacing.

Table 2.4 Bend Index and Drag Factor.

Example 2.4
Using the AGA method, calculate the transmission factor and friction factor for gas flow in an NPS 20 pipeline with 0.500 in. wall thickness. The flow rate is 200 MMscfd, gas gravity = 0.6, and viscosity = 0.000008 lb/ft·sec. The absolute pipe roughness is 700 u in.

Assume a bend index of 60°

Solution

Inside diameter of pipe = 20 – (2 x 0.5) = 19.0 in.

\[
Re = 1.35 \times 10^{-5} \quad (\text{GQ/\mu d}) \quad \text{Eq. 1.34}
\]

Reynolds number = 10,685,214
Next, we will calculate the two transmission factors.

1. The fully turbulent transmission factor, using Equation 2.36, is

\[ F = 4\log_{10}(3.7d / \varepsilon) \]

\[ F = 4\log_{10}(3.7 \times 19 / 0.0007) = 20.01 \]

For the smooth pipe zone, using Equation 2.38, the Von Karman transmission factor is

\[ F_t = 4\log_{10}(10685214/F) - 0.6 \]

Solving this equation by trial and error, we get \( F_t = 22.13 \).

From Table 2.4, for a bend index of 60°, the drag factor \( D_t \) is 0.96.

Therefore, for the partially turbulent flow zone, using Equation 2.37, the transmission factor is

\[ F = 4D_t \log_{10}\left(\frac{10685214}{1.4125 F_t}\right) - 0.6 \]

\[ F = 4 \times 0.96 \log_{10}\left(10685214/1.4125 \times 22.13\right) = 21.25 \]

From the above two values of \( F \), using the smaller number, we get the AGA transmission factor as \( F = 20.01 \).

Therefore, the corresponding friction factor \( f \) is found from Eq. 2.31 \( (f = 4 / F^2) \)

\[ f = 0.0100 \]

2.4.2 Low Pressure Gas Flow

Gas gathering often involves operating pressures below 100 psi. Some systems flow under vacuum conditions. For these low pressure conditions, equations have been developed that give a better fit than the Weymouth or Panhandle equations.

Two such formulas are:

The Oliphant equation and the Spitzglass equation

2.4.2.1 Oliphant equation

The Oliphant equation was derived for use in gas distribution mains prevalent in the early 1900s. Use is restricted to pipes from vacuum to 35 psig. It is most reliable for gases with specific gravities around 0.75. The Oliphant equation is of the form

\[ Q = \frac{(314496/GT)}{(d^{2.5}+d^{3/30})} \left(14.4/P_0\right) \left(T_0/520\right) \left[(P_1^2 - P_2^2)/L_m\right]^{0.5} \quad \text{Eq. 2.39} \]

where

\[ Q = \text{volumetric gas flow rate, scfd} \]
\[ T_0 = \text{temperature base (°R)} \]
\[ P_0 = \text{pressure base (psia)} \]
\[ d = \text{pipe inside diameter (in.)} \]
\[ G = \text{gas gravity (air = 1.0)} \]
\[ T = \text{gas flowing temperature (°R)} \]
\[ L_m = \text{pipeline length (miles)} \]
\[ P_1 = \text{pipeline inlet pressure (psia)} \]
\[ P_2 = \text{pipeline discharge pressure (psia)} \]
2.4.2.2 Spitzglass Equation
The Spitzglass equation has been around for many years and originally was used in fuel gas piping calculations. There are two versions of the Spitzglass equation. One equation is for low pressure (less than or equal to 1 psig) and another is for high pressure (more than 1 psig). These equations have been modified to include a pipeline efficiency and compressibility factor. The low-pressure (less than or equal to 1 psig) version of the Spitzglass equation is

\[ Q = 3.839 \times 10^3 \left( \frac{T_b}{P_b} \right) E d^{2.5} \left[ \frac{(P_1 - P_2)}{G T_f L_e Z (1+(3.6/d)+0.03d)} \right]^{0.5} \] Eq. 2.40

The high-pressure (more than 1 psig) version is as follows.

\[ Q = 729.6087 \left( \frac{T_b}{P_b} \right) E d^{2.5} \left[ \frac{(P_1^2 - e^s P_2^2)}{G T_f L_e Z (1+(3.6/d)+0.03d)} \right]^{0.5} \] Eq. 2.41

Where
\[ L_e = \text{pipeline equivalent length (miles)}, \]
\[ E = \text{pipeline efficiency, a decimal value less than or equal to 1.0} \]
\[ = 1.0 \text{ for brand new pipe}, \]
\[ = 0.95 \text{ for good operating conditions}, \]
\[ = 0.92 \text{ for average operating conditions}, \]
\[ = 0.85 \text{ for unfavorable operating conditions}. \]

Example 2.5
Calculate the fuel gas capacity of an NPS 6 pipe, with an inside diameter of 6 in. and a total equivalent length of 180 ft. The flowing temperature of fuel gas is 60°F, and the inlet pressure is 1.0 psig. Consider a pressure drop of 0.1 psi and the specific gravity of gas = 0.6. Assume pipeline efficiency \(E = 1.0\) and compressibility factor \(Z = 1.0\). The base pressure and base temperature are 14.7 psia and 60°F, respectively.

Solution
Base temperature = 60 + 460 = 520°F
Gas flowing temperature = 60 + 460 = 520°F
Pressure drop \((P_1 - P_2) = 0.9 \text{ psi} \) (1 in. of water = 0.03609 psi)
Since this is low pressure, using Spitzglass Equation 2.40, we get
Length = 180/5280 mile.

\[ Q = 3.839 \times 10^3 \left( \frac{520}{14.7} \right) \times 1 \times 6^{2.5} \left[ (0.9) / (0.6 \times 520 \times (180/5280)) \times (1+(3.6/6)+0.03 \times 6) \right]^{0.5} \]
\[ Q = 3.4 \text{ MMSCFD} \]

2.4.3 Empirical gas flow equations
Due to the tedious nature of solving the general gas equation, a number of empirical equations were developed. The Weymouth, Panhandle “A,” and Panhandle “B” were developed to simulate compressible gas flow in pipelines. The equations were developed from the fundamental energy equation for compressible flow, but each has a special representation of the friction factor to allow the equations to be solved analytically.

2.4.3.1 Weymouth equation
The Weymouth equation is most accurate for small-diameter, less than 12 in., short lengths of pipe where high pressure drops are likely, that is, turbulent flow. For larger pipe, the Weymouth

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equation becomes increasingly conservative, that is, predicted flow capacities become increasingly less compared with actual flow capacities. The Weymouth equation is based on the following assumptions:

- Turbulent flow.
- High Reynolds number exists.
- Friction factor is dependent upon relative roughness, \( \varepsilon/D \).

Experience indicates that Weymouth’s equation is suitable for most gas piping within the facility. However, the friction factor used by Weymouth is generally too low for large-diameter or low-velocity lines, where the flow regime is more properly characterized by the sloped portion of the Moody diagram.

For steel pipe with a fixed absolute roughness, \( \varepsilon \), the friction factor was assumed as

\[
f = 0.032 / d^{0.333}
\]

Substituting the above friction factor, into the general equation, Equation 2.19, for a gas yields the Weymouth equation as shown below:

\[
Q = 433.5 \times E \left( \frac{T_b}{P_b} \right) d^{2.667} \left[ \frac{(P_1^2 - \varepsilon^2 P_2^2)}{Z T_s Le G} \right]^{0.5} \quad \text{Eq. 2.42}
\]

Where

- \( E = \) pipeline efficiency factor,

Example 2.6

Given:
- \( Q = 100 \) MMSCFD
- \( P_1 = 1214.7 \) psia
- \( P_2 = 1014.7 \) psia
- \( G = 0.621 \)
- \( T = 70^\circ \text{F} \) or 530 °R
- \( Le = 79,200 \) ft. = 15 mile
- \( Z = 0.863 \)
- \( E = \) pipeline efficiency factor, = 0.95

Use the Weymouth equation to determine the diameter of the pipe.

\[
Q = 433.5 \times E \left( \frac{T_b}{P_b} \right) d^{2.667} \left[ \frac{(P_1^2 - \varepsilon^2 P_2^2)}{Z T_s Le G} \right]^{0.5}
\]

\[
100 \times 10^6 = 433.5 \times 0.95 \times 0.000008 \times \frac{520}{14.7} \times d^{2.667} \left[ \frac{(445880 - 0.863 \times 530 \times 15 \times 0.621)}{0.863 \times 530 \times 15 \times 0.621} \right]^{0.5}
\]

\[
= 14568 \times 10^{2.667} \times d^{2.667} \times 149030 = 671 \]

\[
d = 11.5
\]

Example 2.7

Calculate the flow rate using the Weymouth equation in a gas pipeline system, 15 miles long pipe with internal diameter 12. 50 in. at an efficiency of 0.95. The upstream pressure is 1200 psia, and the delivery pressure required at the end of the pipe segment is 750 psia. Use gas gravity = 0.59 and viscosity = 0.000008 lb/ft-sec. The flowing temperature of gas = 75°F, base pressure = 14.7 psia, and base temperature = 60°F.

Assume compressibility factor to be 0.94.

Neglect elevation difference along the pipe. How does this compare with the flow rate calculated using the General Flow equation with the Colebrook friction factor?

Assume a pipe roughness of 700 u in.
Solution
Using Equation 2.44, we get the flow rate for the Weymouth equation as follows:
\[ Q = 433.5 \times (T_b/P_b)^{2.667} \left( (P_1^2 - e^8P_2^2) / (L T_f L_e G) \right)^{0.5} \] Eq. 2.44
\[ Q = 433.5 \times (0.95) \times (520/14.7)^{2.667} \left( (1200^2 - 750^2) / (0.94 \times 535 \times 15 \times 0.59) \right)^{0.5} \]
\[ Q = 163,255,858 \text{ SCFD} \]
or
\[ Q = 163.26 \text{ MMSCFD} \]

For General equation, we will calculate the Reynolds number using Equation 1.34
\[ Re = 1.35 \times 10^{-5} (SQ/\mu \ d) \] Eq. 1.34
\[ Re = 0.08 Q \ \text{ result (a)} \]
Since \( Q \) is unknown, we will first assume a transmission factor \( F = 20 \) and calculate the flow rate from the General Flow Equation 2.19.
\[ Q = 38.77 \times F \times (T_0/P_0)^{0.8539} \left( (P_1^2 - e^8P_2^2) / (L T_f L_e G) \right)^{0.5394} \] Eq. 2.19
\[ Q = 38.77 \times 20 \times (520/14.7)^{2.5} \left( (1200^2 - 750^2) / (0.59 \times 535 \times 15 \times 0.94) \right)^{0.5} \]
\[ 15152588 \text{ (14)} \]
\[ Q = 212.1 \text{ MMscfd} \]

Next, we will calculate the Reynolds number and the transmission factor based on this flow rate from result (a) above, as follows:
\[ Re = 0.08 \times 212.1 \times 10^6 = 16.45 \times 10^6 \]
and, using Equation 2.33,
\[ F = -4 \log_{10} \left( \left( \epsilon/3.7d \right) + (1.255 F/Re) \right) \] Eq. 2.33
\[ F = -4 \log_{10} \left( \left( 700 \times 10^6 / 3.7 \times 12.5 \right) + (1.255 F/16.45 \times 10^6) \right) \]
Solving for \( F \), we get: \( F = 19.09 \)
Using this value, the revised flow rate is found by proportion as \( 202.4 \) MMscfd
Repeating the calculation of \( Re \) and \( F \), we get
\[ Re = 16.2 \times 10^6 \]
Therefore, \( F = 19.08 \).
This is fairly close to the previous value of \( F = 19.09 \); therefore, we will use this value and calculate the flow rate as \( 202.4 \) MMscfd.
Comparing this result using the General Flow equation with that calculated using the Weymouth equation, we see that the latter equation is quite conservative.

2.4.5.2 Panhandle “A” equation
Two gas flow equations were developed by Panhandle Eastern for calculating flow rates in long, larger diameter (12 in. and above) cross-country gas transmission pipelines.
Today, only the Panhandle “B” equation is used. To help the reader understand the basis of the Panhandle “B” equation, the discussion below reviews the basis of the original Panhandle “A” equation.

The first equation is frequently called the “Panhandle A” equation. It is expressed as
\[ Q = 435.87 \times (T_b/P_b)^{1.0788} \times d^{2.6182} \left( (P_1^2 - e^8P_2^2) / (L T L G)^{0.8539} \right)^{0.5394} \] Eq. 2.43
where
\( Q \) = volumetric gas flow rate, scfd (ft.\(^3\)/d at \( P_0 \) and \( T_0 \)),
\( E \) = pipeline efficiency, dimensionless,
\( T_0 \) = temperature base (°R),
\( P_0 \) = pressure base (psia),
\( d \) = pipeline inside diameter (in.),
\( P_1 \) = inlet pressure (psia),
\( P_2 \) = discharge pressure (psia),
\( G \) = gas gravity (air = 1),
\( T \) = gas flowing temperature (°R),
\( L \) = pipeline length (miles).

Panhandle Eastern originally recommended an efficiency factor of 0.92. Subsequent field experience indicated that the efficiency factor should be calculated according to the equation

\[ E = 0.92C \]

Eq. 2.44

where
\( C \) = experience factor given in Table 2.3.

<table>
<thead>
<tr>
<th>OD (in.)</th>
<th>ID (in.)</th>
<th>C</th>
<th>OD (in.)</th>
<th>ID (in.)</th>
<th>C</th>
</tr>
</thead>
<tbody>
<tr>
<td>2.375</td>
<td>2.067</td>
<td>0.900</td>
<td>2.875</td>
<td>2.469</td>
<td>0.900</td>
</tr>
<tr>
<td>3.500</td>
<td>3.048</td>
<td>0.900</td>
<td>4.500</td>
<td>4.026</td>
<td>0.900</td>
</tr>
<tr>
<td>6.625</td>
<td>6.125</td>
<td>0.900</td>
<td>8.625</td>
<td>8.125</td>
<td>0.900</td>
</tr>
<tr>
<td>10.750</td>
<td>10.250</td>
<td>0.900</td>
<td>12.750</td>
<td>12.250</td>
<td>0.900</td>
</tr>
<tr>
<td>14.000</td>
<td>13.376</td>
<td>0.900</td>
<td>16.000</td>
<td>15.376</td>
<td>0.917</td>
</tr>
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<td>18.000</td>
<td>17.512</td>
<td>0.937</td>
<td>20.000</td>
<td>19.512</td>
<td>0.951</td>
</tr>
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<td>22.000</td>
<td>19.512</td>
<td>0.951</td>
<td>24.000</td>
<td>23.250</td>
<td>0.989</td>
</tr>
<tr>
<td>26.000</td>
<td>25.250</td>
<td>1.000</td>
<td>30.000</td>
<td>29.188</td>
<td>1.031</td>
</tr>
<tr>
<td>36.000</td>
<td>35.188</td>
<td>1.054</td>
<td>42.000</td>
<td>40.876</td>
<td>1.070</td>
</tr>
</tbody>
</table>

Table 2.3. Experience factor "C" for use in the Panhandle "A" equation

2.4.5.3 Panhandle “B” equation
Panhandle Eastern subsequently issued a revised version of the Panhandle “A” equation; it too is best used for long, larger diameter (greater than 12 in.) cross-country gas pipelines. This equation has revised exponents and includes the gas compressibility factor. It assumes that the friction factor can be represented by a straight line of constant negative slope \( (f_{m}=C/Re^0) \) in the moderate Reynolds number region of the Moody friction factor diagram.

\[ Q = 737 \left( \frac{T_b}{P_b} \right)^{0.02} E d^{2.53} \left[ (P_1^2 - e^8 P_2^2) / G^{0.961} Z_{avg} T_f L_e \right]^{0.51} \]

Eq. 2.45

where
\( Q \) = volumetric flow rate,(MMscfd),
\( d \) = pipeline inside diameter (in.),
\[ P_1 = \text{pipeline inlet pressure (psia)}, \]
\[ P_2 = \text{pipeline outlet pressure (psia)}, \]
\[ L_n = \text{pipeline length (miles)}, \]
\[ G = \text{gas specific gravity (air = 1)}, \]
\[ T = \text{gas flowing temperature (°R)}, \]
\[ Z = \text{compressibility factor of gas, dimensionless}, \]
\[ E = \text{efficiency factor, dimensionless}, \]
\[ = 1.0 \text{ for brand new pipe,} \]
\[ = 0.95 \text{ for good operating conditions,} \]
\[ = 0.92 \text{ for average operating conditions,} \]
\[ = 0.85 \text{ for unfavorable operating conditions.} \]

In practice, the Panhandle “B” equation is commonly used for large-diameter long cross-country gas pipelines where the Reynolds number is on the straight-line portion of the Moody diagram. Neither the Weymouth nor the Panhandle “B” equation represents a “conservative” assumption that can always be used to overstate pressure drop. If the Weymouth equation is used and the flow is in the moderate Reynolds number range, the friction factor will in reality be higher than assumed (because the sloped line portion is higher than the horizontal portion of the Moody diagram), and the actual pressure drop will be higher than calculated. If the Panhandle “B” equation is used and the flow is actually in a high Reynolds number regime, the friction factor also will be higher than assumed (because the equation assumes the friction factor continues to decline with increased Reynolds number beyond the horizontal portion of the curve), and the actual pressure drop will again be higher than calculated.

Example 2.8
Application of the Panhandle “B” equation
Given data from Examples 2.6
Since the pipe diameter calculated in Examples 2.6 using the Weymouth equation approaches the upper limit of the equation, it is desirable to check the diameter with the “Panhandle B” equation. Assume that the efficiency factor \( E = 0.90 \).

Solution:
(1) From Example 2.6
\[ Q_g = 100 \text{ MMSCFD} \]
\[ P_1 = 1214.7 \text{ psia} \]
\[ P_2 = 1014.7 \text{ psia} \]
\[ G = 0.621 \]
\[ T = 70^\circ \text{ F or 530 °R} \]
\[ L = (79,200 \text{ ft.} / 5280) = 15 \text{ mile.} \]
\[ Z = 0.863 \]

(2) Solving for diameter, the Panhandle “B” equation becomes
\[ Q = 737 \left( \frac{T_b}{P_b} \right)^{1.02} E d^{2.53} \left[ \left( \frac{P_1^2 - P_2^2}{G} \right)^{0.961} Z_{\text{avg}} T_f L_e \right]^{0.51} \quad \text{Eq. 2.45} \]
\[ 100 \times 10^6 = 737 \left( \frac{T_b}{P_b} \right)^{1.02} (0.9) d^{2.53} \left[ (1214.7^2 - 1014.7^2) / (0.621)^{0.961} x 0.863 x 530 x 15 \right]^{0.51} \]
\[ = 25198 d^{2.53} [(445880)/(4340)]^{0.51} \]
\[ = 25198 d^{2.53} [(445880)/(4340)]^{0.51} \]
\[ d^{2.53} = 100 \times 10^6/267515 \approx 374 \]
Books are available [here](http://www.oilprocessing.net/oil/)

d = 10.40 in.
For a pipe diameter at the upper limit of the Weymouth equation and at the lower limit of the Panhandle "B" equation, the Panhandle "B" gives somewhat smaller pipe size.

2.4.5.4 Institute of Gas Technology (IGT) Equation
The IGT equation proposed by the Institute of Gas Technology is also known as the IGT distribution equation and is stated as follows:

\[
Q = 136.9 \ E \ (T_b/P_b) \ d^{2.667} \ [(P_1^2 - e^sP_2^2)/G^{0.8} \ T_f \ L_e \ u^{0.2}]^{0.5} \quad \text{Eq. 2.46}
\]

where
\[
\begin{align*}
Q & = \text{volume flow rate, standard ft}^3/\text{day (SCFD)} \\
E & = \text{pipeline efficiency, a decimal value less than 1.0} \\
P_b & = \text{base pressure, psia} \\
T_b & = \text{base temperature, °R (460 + °F)} \\
P_1 & = \text{upstream pressure, psia} \\
P_2 & = \text{downstream pressure, psia} \\
G & = \text{gas gravity (air = 1.00)} \\
T_f & = \text{average gas flow temperature, °R (460 + °F)} \\
L_e & = \text{equivalent length of pipe segment, miles} \\
Z & = \text{gas compressibility factor, dimensionless} \\
D & = \text{pipe inside diameter, in.} \\
u & = \text{gas viscosity, lb/ft-s}
\end{align*}
\]

Example 2.9
Using the IGT equation, calculate the flow rate in a natural gas pipeline, NPS 16 with 0.250 in. wall thickness, 15 mi long. The inlet and outlet pressure are 1000 psig and 800 psig, respectively. The gas gravity \( G = 0.6 \) and viscosity \( u = 0.000008 \) lb/ft-s.
The average gas temperature is 80°F, base pressure is 14.7 psia, and base temperature is 60°F. The compressibility factor \( Z = 0.90 \), and the pipeline efficiency is 0.95.

Solution
Inside diameter of pipe \( = 16 - (2 \times 0.25) = 15.5 \) in.
The pressures given are in psig, and they must be converted to absolute pressures.
Therefore,
\[
\begin{align*}
P_1 & = 1000 + 14.7 = 1014.7 \text{ psia} \\
P_2 & = 800 + 14.7 = 814.7 \text{ psia} \\
T_b & = 60 + 460 = 520 \text{ °R} \\
T_f & = 80 + 460 = 540 \text{ °R}
\end{align*}
\]

Substituting in IGT Equation 2.46, we get
\[
Q = 136.9 \ (520/14.7) \ (0.95) \ (15.5)^{2.667} \ [(1014.7^2 - 814.7^2)/0.6^{0.8} \times 540 \times 15 \times (0.000008)^{0.2}]^{0.555}
\]
\[
\begin{align*}
520/14.7 & = 36.4 \\
Z & = 7538748 \\
Q & = 288 \text{ MMSCFD}
\end{align*}
\]
2.4.6 Selection of gas flow equations

2.4.6.1 General equation
General Equation—Use for most general usage. If it is inconvenient to use the iterative procedure of the general equation but high accuracy is required, compute the results using both the Weymouth and Panhandle “B” equations and use the higher calculated pressure drop.

2.4.6.2 Weymouth equation
The Weymouth Equation—Use the Weymouth equation only for small-diameter, short run pipe within the production facility where the Reynolds number is expected to be high. The use of the Weymouth equation for pipe greater than 20 in. in diameter or in excess of 15,000 ft. (4600 m) long is not recommended.

2.4.6.3 Panhandle “B” equation
The Panhandle “B” Equation—Use the Panhandle “B” equation only for large diameter, long run pipelines where the Reynolds number is expected to be moderate.

2.4.6.4 Spitzglass equation
Spitzglass Equation—Use the Spitzglass equation for low-pressure vent lines less than 12 in. in diameter.

2.4.6.5 Oliphant equation
Oliphant Equation—Use the Oliphant equation for lines with pressures less than 35 psig (2 barg) and gases with gravities between 0.7 and 0.8.

2.4.6.6 IGT equation
Used on gas distribution piping

2.4.7 Steam equations - Babcock equation

\[ \Delta P = 3.63 \times 10^{-8} \left[ \frac{(d+3.6)^6}{d^6} \right] \left[ \frac{W^2L}{\rho} \right] \]  Eq. 2.47

where
\[ \Delta P = \text{differential pressure (psig)}, \]
\[ W = \text{rate of flow (lbs/h)}, \]
\[ \rho = \text{specific volume of fluid (ft.}^3/{\text{lb}}), \]
\[ L = \text{length of pipe (ft.)}, \]
\[ d = \text{pipe inside diameter (in.)}. \]
Chapter Three

Two-phase flow

3.1 Introduction
There are situations where more than one fluid phase is transported simultaneously through a pipeline. Examples of two-phase flow are found in pipelines transporting natural gas and condensate and crude oil and associated gas. These conditions occur when fluids are transported in a pipeline such as

- crude oil produced below the bubble point,
- natural gas produced above the dew point.

Two-phase flow is further complicated by the presence of a second liquid phase, such as

- water with oil,
- gas or glycol with gas and condensate stream.

Designing a pipeline for two-phase or multiphase flow is complex because the properties and flow patterns of two or more fluids must be covered in the design.

3.2 Factors that affect two-phase flow
When designing a two-phase pipeline, the following factors must be considered:

1. Liquid volume fraction
2. Pipeline profile
3. Flow regime
4. Liquid holdup
5. Two-phase pressure loss

Knowing how much gas and liquid is present in a hydrocarbon mixture is essential in designing a two-phase flow pipeline. The two-phase gas-liquid composition of a hydrocarbon mixture is best illustrated by a phase envelope. The phase envelope is a graphic representation of the relative quantities of gas and liquid present in a mixture at various pressures and temperatures. Figure 3.1 is an example of a typical phase envelope of a gas-condensate reservoir.
Figure 3.1. Typical phase envelope of a gas-condensate reservoir.

3.2.1 Liquid volume fraction
The liquid volume fraction is the fraction of fluid flowing as a liquid present at a local point in the pipeline. The phase envelope illustrated in Figure 3.1 graphically represents this. It is determined by assuming that thermodynamic equilibrium exists between the two phases at the local pressure and temperature.

3.2.2 Pipeline profile
Changes in elevation of the pipeline route play an important role in two-phase flow pipeline design and operation. The elevation profile influences the liquid holdup and determines the flow regime. This in turn impacts the pressure loss incurred during the two-phase flow.

3.2.3 Liquid holdup
The liquid holdup is the accumulation of liquid in the pipeline due to the difference in velocity, "slip," between the phases and the influence of gravity. This is illustrated in Figure 3.2. Liquid holdup is usually larger than the liquid volume fraction. The only exception might be during a downward sloping flow at low velocity.

3.2.4 Two-phase flow regimes
The way in which gas and liquid simultaneously flow within a pipeline is characterized by the flow regime. Flow regimes indicate the distribution of the phases over the cross-sectional area of the pipeline. The type of flow pattern depends primarily on the gas and liquid flow rates. Some of the most commonly occurring two-phase flow regimes in a pipeline are the following:

1. Bubble flow
2. Annular dispersed flow
3. Stratified or laminar flow
4. Plug or churn flow
5. Slug flow

The relative positions of the liquid and gas as they move along the pipeline depend on whether the flow is horizontal or vertical. When a gas-liquid mixture enters a horizontal pipeline, the two phases tend to separate with the heavier liquid settling to the bottom. The type of flow pattern depends primarily on the gas and liquid flow rates. Figure 3.3 shows typical horizontal two-phase pipe flow. Horizontal flow regimes can be described as follows:

\[
\frac{A_L}{A} = H_L = \text{Liquid Hold-up}
\]

where

\[
A = \text{Total cross-sectional area}
\]
\[
A_L = \text{Liquid cross-sectional area}
\]

Figure 3.2 Liquid holdup in a pipeline.

3.2.4.1 Horizontal flow regimes
3.2.4.1.1 Dispersed bubble flow:
In this type of flow, bubbles of gas flow along with the liquid at about the same velocity. Gas
bubbles move along the upper part of the pipe.

3.2.4.1.2 Elongated bubble (plug flow):
It is similar to bubble flow, except the gas bubbles coalesce into large bubbles or plugs. Gas bubbles become larger and form moderate-size gas plugs.

3.2.4.1.3 Stratified smooth flow:
In stratified flow, both the gas plugs become longer until the gas and liquid flow separately with a distinct and relatively smooth interface between them. Gas flows in the upper part and liquid moves in the lower part of the pipe. Sometimes, a wave is formed in which case it is called stratified wavy flow. This type of flow is commonly seen in downhill lines.

3.2.4.1.4 Slug flow;
Slug flow is similar to stratified wavy flow, except that the top of some waves touches the top of the pipe wall, trapping gas slugs between wave crests. Flow tends to be alternative slugs of liquid and gas. The length of these slugs can be several hundred feet long in some cases. This is probably the most commonly encountered regime in level and uphill lines. The intermittent nature of the liquid slugs makes downstream separator sizing more difficult, which may require larger separation facilities.

3.2.4.1.5 Annular-mist flow;
At extremely high gas-liquid ratios, the liquid slugs become more and more aerated until they are indistinguishable from the intervening bubbles. In this condition, the liquid coats the pipe wall in an annular layer and the gas flows in the core, suspending droplets of liquid.

Figure 3.3. Two-phase flow regimes in a horizontal pipe.

Figure 3.4 can be used to approximate the type of flow regime expected for any flow condition. Coordinates of this map are superficial liquid velocity (VSL) and superficial gas velocity (VSG). A superficial velocity is the volumetric flow rate of that phase divided by the total pipe cross-sectional area. *(The superficial liquid velocity is the velocity the liquid would have in the pipeline, if only the liquid phase was present. Likewise, the superficial gas velocity is the velocity the gas would have, if only the gas phase was present.)* The use of superficial liquid and gas velocities is an attempt to normalize fluid flow rates on pipe size. This flow regime map should be used only
for very rough regime estimation, because it was developed for horizontal pipe flowing air and water at low pressure.

Fluid properties alter the map, and pipe inclination has a profound effect on the map. Inclination changes from horizontal on the order of 0.1-1.0° effect large changes in the map. Upward inclinations cause the growth of the slug flow regime at the expense of the stratified flow regime, while downward inclinations cause the growth of the stratified flow regime at the expense of the slug flow regime.

![Horizontal multiphase flow regime map.](http://www.oilprocessing.net/oil/figure-3.4.png)

**Figure 3.4.** Horizontal multiphase flow regime map.

### 3.2.4.2 Vertical flow regimes

Two-phase flow patterns in vertical flow are somewhat different from those occurring in horizontal flow. Different flow regimes may occur at different segments of pipe; such as flow in a well tubing string where the pressure loss causes gas to come out of solution as the fluid moves up the well. Figure 3.5 shows typical flow regimes in vertical two-phase flow. Vertical flow regimes can be described as follows:

#### 3.2.4.2.1 Bubble flow

In this type of flow, bubbles of gas are present as small, randomly distributed bubbles, whose diameters also vary randomly. The bubbles move at different velocities depending upon their respective diameters. The liquid moves up the pipe at a fairly uniform velocity, and, except for its density, the gas phase has little effect on the pressure gradient.

#### 3.2.4.2.2 Slug flow

In this type of flow, the gas phase is more pronounced. Although the liquid phase is still continuous, the gas bubbles coalesce and form stable bubbles of approximately the same size and shape, which are nearly the diameter of the pipe. These bubbles are separated by slugs of liquid. In vertical flow, alternative slugs of liquid flow in the center of the pipe along with the gas. Both the gas and liquid phases have significant effects on the pressure gradient.

http://www.oilprocessing.net/oil/
3.2.4.2.3 Churn (transition) flow
The change from a continuous liquid phase to a continuous gas phase occurs in this region. The gas phase is predominant and the liquid becomes entrained in the gas. The gas bubbles coalesce into large bubbles or plugs. The gas plugs tend to flow in the center of the pipe. The effects of the liquid are still significant.

![Figure 3.5. Two-phase flow regimes in a vertical pipe.](image)

3.2.4.2.4 Annular-mist flow
In this flow regime, the gas phase is continuous and the bulk of the liquid is entrained in and carried by in the gas. The liquid flows along the wall of the pipe, while the gas flows at high velocity in the center of the pipe. The effects of the liquid film on the pipe wall are secondary. The gas phase is the controlling factor. Normally, flow in oil wells is in the slug or transition flow regime. Flow in gas wells can be in mist flow. Figure 3.6 can be used to determine the type of regime to be expected.

![Figure 3.6 Vertical multiphase flow regime map.](image)
3.3 Two-phase pressure loss
The pressure drop in a two-phase pipeline is higher than that of a single-phase pipeline. The increased pressure loss is essentially caused by the additional friction loss due to the formation of waves and slugs. A number of different correlations have been developed by different authors to predict pressure loss in a pipeline for two-phase flow. Some are more suitable than others for specific fluid types, giving varying levels of uncertainty. Many companies develop their own correlations for two-phase flow that can be generally applied. Specialized computer simulation programs are used to evaluate the behavior of two-phase pipelines. The simulation programs take the input data and use two-phase flow equations and correlations to determine the flow regimes, pressure loss, and liquid holdup for the pipeline.

3.4 Two-phase liquid holdup
In a two-phase pipeline, the liquid holdup varies with throughput and with varying pipeline diameters. Change in temperature and/or pressure also affect the liquid holdup due to expansion or contraction of gas inside pipeline. Therefore, the designer must predict the amount of liquid expected to be present and then manage it. An understanding of how the anticipated liquid holdup in a pipeline will vary with flow rates will greatly help in formulating a suitable liquid management system. This could include a slug catcher, pigging facilities, or operational limits on ramp-up times.

3.5 Two-phase flow correlations
In two-phase piping, pressure drop is caused by the friction developed due to the energy transfer between the two phases, as well as that between each phase and the pipe wall. Pressure drop calculations must take into account the additional friction loss due to the energy transfer between phases.

The detailed calculation of pressure drops in two-phase pipelines requires an evaluation of phase changes due to pressure and temperature changes, evaluation of liquid holdup using empirical formulas, and evaluation of energy transfer between the phases. These are addressed in the many computer programs available. It should be kept in mind that even under the best conditions, small changes from horizontal in piping systems can lead to large errors in calculating pressure drop.

Table 3.1 lists several commercial programs that are available to model pressure drop. Because all are based to some degree on empirical relations, they are limited in accuracy to the data sets from which the relations were designed. It is not unusual for measured pressure drops in the field to differ by +/- 20% from those calculated by any of these models.

<table>
<thead>
<tr>
<th>Multiphase transient simulation</th>
<th>Multiphase steady-state simulation</th>
<th>Sinline-phase gas/liquid transient simulation</th>
</tr>
</thead>
<tbody>
<tr>
<td>OLGA ProFES</td>
<td>Pipe flow, well flow, PIPEPHASE, PIPESIM-SUITE, GENNET-M, Flowsystem, Prosper, Gap, &amp; Perform</td>
<td>Pipeline simulator Winflo, NATASHA PLUS, TLNET, &amp; TGNET</td>
</tr>
</tbody>
</table>

Table 3.1. Examples of hydraulic modeling and flow assurance software

The following three correlations have been found to give reasonable results when used within the limitations inherent in their derivation.

http://www.oilprocessing.net/oil/
3.5.1 Beggs and Brill equation
This correlation was developed by two University of Tulsa engineering students, Dale Beggs and James Brill. Almost all correlations prior to the Beggs and Brill equation could predict pressure drop in two-phase flow for vertical or horizontal flow only. Prior to 1973, however, no correlation existed for predicting the pressure drop in two-phase flow at any angle of inclination. Beggs and Brill therefore set out to develop such a correlation.

Using the best correlations available for pressure drop and liquid holdup, predictions may vary in error by +/- 20% for perfectly horizontal pipe, +/- 50% for inclined and looped systems.

Selection of the appropriate correlation should be based on field data and reservoir phase characteristics. Two commonly used correlations include API RP 14E approximation equation, and the AGA equation.

Both correlations work well under specific conditions.

3.5.2 API RP 14E equation
The following formula, presented in the American Petroleum Institute’s Recommended Practice API RP 14E, is derived from general equation for isothermal flow assuming that the pressure drop is less than 10% of the inlet pressure:

\[ \Delta P = 3.36 \times 10^{-6} fL \frac{W_h^2}{d^5} \rho_m \]  
Eq. 3.1

where
\( \Delta P \) = pressure drop (psi),
\( L \) = length of pipe (ft.),
\( W_h \) = flow rate of liquid and vapor (lb/h),
\( \rho_m \) = mixture density (lb/ft.\(^3\)),
\( d \) = pipe inside diameter (in.),
\( f \) = Moody friction factor.

This equation assumes that there is no energy interchange between the phases, that bubble or mist flow exists so that the fluid can be described by average mixture density, and that there are no elevation changes.

The flow rate of the mixture to use in this equation can be calculated as follows:

\[ W_h = 3180 Q_g G + 14.6 Q_l (SG) \]  
Eq. 3.2

where
\( W_h \) = flow rate of liquid and vapor (lb/h),
\( Q_g \) = gas flow rate (MMSCFD),
\( Q_l \) = liquid flow rate (BPD),
\( G \) = specific gravity of gas relative to air,
\( SG \) = specific gravity of liquid relative to water.

The density of the mixture to use in Equation 3.1 is given by

\[ \rho_m = \frac{(12409 SG P + 2.7 R_a G P)}{(198.7 P + R_a TZ)} \]  
Eq. 3.3

where
\( \rho_m \) = mixture density (lb/ft.\(^3\)),
\( P \) = pressure (psia),

http://www.oilprocessing.net/oil/
SG = specific gravity of the liquid relative to water (use the average gravity for the hydrocarbon and water mixture),
G = specific gravity of gas relative to air,
R_a = gas-liquid ratio (std ft.³/bbl),
T = temperature (°R),
Z = gas compressibility factor, dimensionless.

3.5.3 AGA equations
The American Gas Association method uses a frictional pressure drop calculation originally developed by Dukler and an elevation pressure drop calculation originally developed by Flanigan.

The AGA equation is based on the following conditions:
- High gas-liquid ratios
- Dukler for frictional component pressure drop
- Flanigan for elevation component at pressure drop

The overall two-phase pressure drop is expressed as follows:

$$\Delta P_t = \Delta P_f + \Delta P_e$$  \hspace{1cm} \text{Eq. 3.4}

where
- $\Delta P_t$ = total two-phase pressure drop (psi),
- $\Delta P_f$ = frictional component of pressure drop (psi),
- $\Delta P_e$ = elevation component of pressure drop (psi).

Frictional component of total two-phase pressure drop is

$$\Delta P_f = f_n f_{fr} \rho_k V_m^2 L_m / (0.14623)d$$  \hspace{1cm} \text{Eq. 3.5}

Elevation component of pressure drop is

$$\Delta P_e = (\rho H_{Li}/144) \sum \Sigma Z_e$$  \hspace{1cm} \text{Eq. 3.6}

where
- $f_n$ = single-phase friction factor,
- $f_{fr}$ = friction factor ratio,
- $\rho_k$ = two-phase mixture density (lb/ft.³),
- $\rho$ = density of liquid (lb/ft.³),
- $V_m$ = velocity of mixture (ft./s),
- $L_m$ = pipeline length (miles),
- $d$ = pipe inside diameter (in.),
- $H_{Li}$ = Flanigan liquid holdup fraction,
- $Z_e$ = sum of vertical elevation rises of pipe (no elevation drops are considered) (ft.).

3.5.4 Calculation procedure
(1) Determine the following liquid volume fraction from the following:

$$\lambda = Q_L / (Q_L + Q_g)$$  \hspace{1cm} \text{Eq. 3.7}

where
- $\lambda$ = flowing liquid volume fraction,
- $Q_L$ = liquid flow rate at flowing conditions (ft.³/s),
- $Q_g$ = gas flow rate at flowing conditions (ft.³/s).

http://www.oilprocessing.net/oil/
(2) Determine the mixture viscosity for the Dukler calculation from the following:

\[ \mu_n = \mu_L \lambda + \mu_g (1-\lambda) \]  
Eq. 3.8

where
\( \mu_n \) = mixture viscosity (cP),
\( \mu_L \) = liquid viscosity (cP),
\( \mu_g \) = gas viscosity (cP),
\( \lambda \) = flowing liquid volume fraction.

(3) Calculate the superficial liquid velocity from the following:

\[ V_{SL} = \frac{Q_L}{A} \]  
Eq. 3.9

where
\( V_{SL} \) = superficial liquid velocity (ft./s),
\( Q_L \) = liquid flow rate at flowing conditions (ft.\(^3\)/s),
\( A \) = cross-sectional area of pipe (ft.\(^2\)), \( = (\pi D^2)/4 \),
\( D \) = internal pipe diameter (ft.).

(4) Calculate the superficial gas velocity from the following:

\[ V_{sg} = \frac{Q_g}{A} \]  
Eq. 3.10

where
\( V_{sg} \) = superficial gas velocity (ft./s),
\( Q_g \) = gas flow rate at flowing conditions (ft.\(^3\)/s),
\( A \) = cross-sectional area of pipe (ft.\(^2\)), \( = (\pi D^2)/4 \), \( D \) = internal diameter, ft.)

(5) Calculate the mixture velocity from the following

\[ V_m = V_{sl} + V_{sg} \]  
Eq. 3.11

where
\( V_m \) = mixture velocity (ft./s),
\( V_{sl} \) = superficial liquid velocity (ft./s),
\( V_{sg} \) = superficial gas velocity (ft./s).

(6) Calculate the friction factor ratio from the following equation:

\[ f_{fr} = 1 / \left( 1.281 - 0.478 y + 0.444 y^2 - 0.094 y^3 + 0.00843 y^4 \right) \]  
Eq. 3.12

where
\( f_{fr} \) = Friction factor ratio
\( y = \frac{U_m}{(\lambda)} \)  
Eq. 3.13

(7) Calculate the two-phase mixture density from the following:

\[ \rho_k = \left( \rho L^2 / H_{LD} \right) + \left[ \rho_g (1-\lambda)^2 \right] / (1-H_{LD}) \]  
Eq. 3.14

where
ρk = two-phase mixture density (lb/ft.³),
ρ = density of liquid (lb/ft.³),
ρg = density of gas (lb/ft.³),
λ = flowing liquid fraction,
Hld = Dukler liquid holdup fraction, = (λ) for the first estimate (assumption).

(8) Calculate an estimate for the mixture Reynolds number from the following:

\[ \text{Re}_\gamma = (124) \frac{\rho_k V_m d}{\mu_n} \quad \text{Eq. 3.15} \]

where
\( \text{Re}_\gamma \) = mixture Reynolds number,
\( \rho_k \) = two-phase mixture density (lb/ft.³),
\( V_m \) = mixture velocity (ft./s),
\( d \) = pipe inside diameter (in.),
\( \mu_n \) = mixture viscosity (cP).

(9) Determine a better estimate for the Dukler liquid holdup fraction (Hld) from Figure 3.7, using λ and the Reynolds number calculated in step 8. (If Reynolds number value is higher than existed in Figure 3.7., go to step (13).

(10) Recalculate the mixture density (\( \rho_k \)) using the improved estimate of Hld in the previous equation in step 7:

\[ \rho_k = (\rho \lambda^2 / Hld) + [\rho_g (1-\lambda)^2] / (1-Hld) \quad \text{Eq. 3.14} \]

(11) Using this new value of \( \rho_k \), recalculate the mixture Reynolds number (\( \text{Re}_\gamma \)) in equation of step 8:

\[ \text{Re}_\gamma = (124) \frac{\rho_k V_m d}{\mu_n} \quad \text{Eq. 3.15} \]

(12) Go back to step 9 and use Figure 3.7 to determine a new value for Hld. Continue this iterative procedure until convergence.

(13) Calculate the single-phase friction factor from the following:

\[ f_n = 0.0056 + 0.5 (\text{Re}_\gamma)^{0.32} \quad \text{Eq. 3.16} \]

where
\( f_n \) = single-phase friction factor,
\( \text{Re}_\gamma \) = mixture Reynolds number.

(14) Calculate the frictional pressure drop from the (Eq. 3.5)

\[ \Delta P_f = f_n f_{pr} \rho_k V_m^2 L_m / (0.14623)d \quad \text{Eq. 3.5} \]

(15) Determine the Flanigan liquid holdup fraction from the following equation:

\[ H_{lf} = 1/[1+0.3264 (V_{sg})^{1.006}] \quad \text{Eq. 3.17} \]
where
\( H_{L1} \) = Flanigan liquid holdup fraction,
\( V_{sg} \) = superficial gas velocity (ft./s).

(16) Calculate the elevation component from Equation Eq. 3.6

\[
\Delta P_e = \left( \rho \frac{H_{L1}}{144} \right) \sum Z_e \quad \text{Eq. 3.6}
\]

(17) Calculate the overall two-phase pressure drop from Equation 3.4

\[
\Delta P_t = \Delta P_f + \Delta P_e \quad \text{Eq. 3.4}
\]

Example 3.1
Application of the AGA equation for two-phase flow
Given:
\( Q_g = 20 \text{ MMSCFD} \)
Condensate = 800 BPD
Water = 230 BPD
Condensate sp.gr = 0.87
Water sp.gr = 1.05
\( \rho \) (gas sp.gr) = 0.67
Inlet pressure = 900 psig = 914.7 psia
Temperature = 80 °F = 540 °R
\( Z \) = 0.85
Condensate viscosity = 3 cP
Gas viscosity = 0.013
Length = 7000 ft. ( = 1.326 mile)
Elevation = 100 ft.
Calculate the pressure drop with both liquid and gas flow in a single 4, 6, and 8 in. line using the AGA equation.
Solution:
(1) Calculate the liquid volume fraction:
\[ \lambda = \frac{Q_L}{Q_L + Q_g} \]
\[ Q_L = (1030 \text{ bbl/d}) \times 5.61458 = 5783 \text{ ft}^3/\text{d} \]
Since \( (PV = nzRT) \)
\[ (Q_g) \times V = 20 \times 10^6 \times 0.85 \times 10.73 \times 540 / 379.5 \times 914.7 = 283761 \text{ ft}^3 \]
\[ \lambda = 5783 / (5783 + 283761) = 0.02 \]
(2) Calculate the mixture viscosity:
\[ \mu_n = \mu_l \lambda + \mu_g (1 - \lambda) \]
\[ \mu_n = 3 \times 0.02 + 0.013(1-0.02) = 0.073 \text{ cP} \]
(3) Calculate the superficial liquid velocity:
\[ V_{sl} = \frac{Q_L}{A} \]
\[ Q_L = (1030 \times 5.61458)/(24 \times 60 \times 60) = \text{ft}^3/\text{sec} \]
\[ V_{sl} = 0.067 \text{ ft}^3/\text{s}/A \]
\[ A = \pi \times d^2/4 \text{ in}^2 \]
\[ A = 12.56 \text{ in}^2 = (0.0872 \text{ ft}^2) \]
\[ A \text{ for 6 in pipe} = 0.196 \text{ ft}^2 \quad \& \quad A \text{ for 8 in pipe} = 0.349 \text{ ft}^2 \]

<table>
<thead>
<tr>
<th>Term</th>
<th>Diameter</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>4 in.</td>
</tr>
<tr>
<td>V_{sl}</td>
<td>0.768 ft./s</td>
</tr>
</tbody>
</table>

(4) Calculate the superficial gas velocity:
\[ V_{sg} = \frac{Q_g}{A} \quad \text{Eq. 3.10} \]
\[ =2.97 \text{ ft}^3/\text{s}/A \]

<table>
<thead>
<tr>
<th>Term</th>
<th>Diameter</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>4 in.</td>
</tr>
<tr>
<td>V_{sg}</td>
<td>34 ft./s</td>
</tr>
</tbody>
</table>

(5) Calculate the mixture velocity:
\[ V_m = V_{sl} + V_{sg} \quad \text{Eq. 3.11} \]

<table>
<thead>
<tr>
<th>Term</th>
<th>Diameter</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>4 in.</td>
</tr>
<tr>
<td>V_{sl}</td>
<td>35.0 ft./s</td>
</tr>
</tbody>
</table>
(6) Calculate the friction factor ratio:
\[ f_{tpr} = \frac{1}{(1.281 - 0.478 y + 0.444 y^2 - 0.094 y^3 + 0.00843 y^4)} \quad \text{Eq. 3.12} \]

\[ y = 3.638 \text{, “refer to Eq. 3.13; } y = \frac{U_n}{(\lambda)} \text{”} \]
\[ f_{tpr} = 0.422 \]

(7) Calculate the two-phase mixture density:

Assume
\[ H_{LD} = \lambda \]
\[ \rho_k = (\rho \lambda^2 / H_{LD}) + [\rho_g (1-\lambda)^2] / (1-H_{LD}) \quad \text{Eq. 3.14} \]

where
\[ \rho = 62.4 \times \text{SG} = 62.4 \times 0.91 = 56.8 \text{lb/ft.}^3 \text{ and} \]
Since, \[ \rho_g = 2.70 \text{ GP/TZ} \text{ (refer to chapter 1, Eq. 1.21)} \]
=2.7X 0.67 x 914.7 / (540 x 0.85) = 3.6 lb/ft.\(^3\)

Since \( H_{LD} = \lambda \), then
\[ \rho_k = (\rho \lambda) + [\rho_g (1-\lambda)] \]
\[ \rho_k = 56.8 \times 0.02 + 3.6 \times (1-0.02) = 4.664 \text{ lb/ft.}^3 \]

(8) Calculate the mixture Reynolds number:
\[ \text{Re}_y = (124) \frac{\rho_k V_m d}{\mu_n} \quad \text{Eq. 3.15} \]
\[ \text{Re}_y = (124) \times 4.664 \times V_m d / 0.073 \]

\begin{center}
\begin{tabular}{|c|c|c|}
\hline
Term & Diameter & \\
\hline
 & 4 in. & 6 in. & 8 in. \\
\hline
\text{Re}_y & 1.2 \times 10^6 & 8 \times 10^5 & 6 \times 10^5 \\
\hline
\end{tabular}
\end{center}

(9) Determine a better estimate for the Dukler liquid holdup fraction (HLD) from Figure 3.7 using \( \lambda \) and the Reynolds number calculated in step 8. In this example, the Reynolds number is so large that the value of HLD is equal to \( \lambda \).

(10) If the HLD is not equal to \( \lambda \) (Value can be extracted from the graph “lower Reynolds number” Recalculate the two-phase mixture density using the improved estimate of HLD:
\[ \rho_k = (\rho \lambda^2 / H_{LD}) + [\rho_g (1-\lambda)^2] / (1-H_{LD}) \quad \text{Eq. 3.14} \]

(11) Recalculate the mixture Reynolds number using the new value of \( \rho_k \):
\[ \text{Re}_y = (124) \frac{\rho_k V_m d}{\mu_n} \quad \text{Eq. 3.15} \]

\begin{center}
\begin{tabular}{|c|c|c|}
\hline
Term & Diameter & \\
\hline
 & 4 in. & 6 in. & 8 in. \\
\hline
\text{Re}_y & 1.2 \times 10^6 & 8 \times 10^5 & 6 \times 10^5 \\
\hline
\end{tabular}
\end{center}

(12) Go back to step 9 and use Figure 3.7 to determine a new value for HLD; continue this iterative procedure until convergence.

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(13) Calculate the single-phase friction factor:
\[ f_n = 0.0056 + 0.5 (Re)^{-0.32} \] Eq. 3.16

<table>
<thead>
<tr>
<th>Term</th>
<th>Diameter</th>
</tr>
</thead>
<tbody>
<tr>
<td>4 in.</td>
<td>6 in.</td>
</tr>
<tr>
<td>( f_n )</td>
<td>0.011</td>
</tr>
</tbody>
</table>

Steps 14, 15, and 16 calculate the frictional and elevation pressure drop components. These are used to determine the overall two-phase pressure drop.

(14) Calculate the frictional pressure drop component:
\[ \Delta P_f = f_n \cdot \frac{f_{pr} \cdot \rho_k \cdot V_{m} \cdot L_m}{(0.14623)d} \] Eq. 3.5

\[ \Delta P_f = f_n \cdot 0.422 \cdot 4.667 \cdot V_{m}^2 \cdot (1.326) / (0.14623)d \]

<table>
<thead>
<tr>
<th>Term</th>
<th>Diameter</th>
</tr>
</thead>
<tbody>
<tr>
<td>4 in.</td>
<td>6 in.</td>
</tr>
<tr>
<td>( \Delta P_f )</td>
<td>74 psi</td>
</tr>
</tbody>
</table>

(15) Calculate the Flanigan liquid holdup fraction from Equation:
\[ H_{Lf} = \frac{1}{1+0.3264 (V_{sg})^{1.006}} \] Eq. 3.17

<table>
<thead>
<tr>
<th>Term</th>
<th>Diameter</th>
</tr>
</thead>
<tbody>
<tr>
<td>4 in.</td>
<td>6 in.</td>
</tr>
<tr>
<td>( H_{Lf} )</td>
<td>0.07</td>
</tr>
</tbody>
</table>

(16) Calculate the elevation pressure drop component.
\[ \Delta P_e = (\rho \cdot H_{Lf} / 144) \sum Z_e \] Eq. 2.6

<table>
<thead>
<tr>
<th>Term</th>
<th>Diameter</th>
</tr>
</thead>
<tbody>
<tr>
<td>4 in.</td>
<td>6 in.</td>
</tr>
<tr>
<td>( \Delta P_e )</td>
<td>2.9</td>
</tr>
</tbody>
</table>

17) Add the frictional component (\( \Delta P_f \)) and the elevation component (\( \Delta P_e \)) to obtain the total overall pressure drop (\( \Delta P_t \)).
\[ \Delta P_t = \Delta P_f + \Delta P_e \] Eq. 3.4

<table>
<thead>
<tr>
<th>Term</th>
<th>Diameter</th>
</tr>
</thead>
<tbody>
<tr>
<td>4 in.</td>
<td>6 in.</td>
</tr>
<tr>
<td>( \Delta P_t )</td>
<td>77</td>
</tr>
</tbody>
</table>
Example 3.2
Application of API RP 14E approximation equation
Given:
SG (Liquid) = 0.91
G = 0.85
Z = 0.67
Q_g = 23 MMSCFD
Q_c = 800 BPD
Q w = 230 BPD
Q_l = 1030 BPD
f (assume) = 0.204 (rough pipe)
L = 7000 ft
Determine:
Calculate the pressure drop with both liquid and gas flow in a single 4, 6, and 8 in. line using the API RP 14E equation.
Solution:
Calculate mixture flow rate:
W = 3180Q_gG + 14.6Q_l(SG)
=3180x23 (0.85) + 14.6x1030x(0.91)
=75,854 lb/h
Calculate gas-liquid ratio:
R=23,000,000/1030
=22,330 ft.3/bbl
Calculate the mixture density at flowing pressure of 900 psi:
ρ_m = [12409 SG P + 2.7 R_a G P] / (198.7 P + R_aTZ) Eq. 3.3
ρ_m=[12409 (0.91)(915) +2.7(22,330) (0.85) (915)] /(198.7 x 915 + 22,330 x 540 x 0.67)
=6.93 lb/ft.3
Calculate the approximate pressure drop:
ΔP = 3.4 x 10^-6 x fLW^2/ρ_m d^5
=3.4 x 10^-6 x 0.024 x 7000 x (75,854)^2 / 6.930 x d^5
=3.98 x 10^5/d^5

<table>
<thead>
<tr>
<th>Term</th>
<th>Diameter</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>4 in.</td>
</tr>
<tr>
<td>ΔP, (psi)</td>
<td>398</td>
</tr>
</tbody>
</table>

3.6 Sizing Criteria for Gas/Liquid Two-Phase Lines.
a. Erosional velocity.
Flow lines, production manifolds, process headers and other lines transporting gas and liquid in two-phase flow should be sized primarily on the basis of flow velocity. Experience has shown that loss of wall thickness occurs by a process of erosion/corrosion. This process is accelerated by
high fluid velocities, presence of sand, corrosive contaminants, such as CO$_2$ and H$_2$S and fittings, which disturb the flow path such as elbows.

The following procedure for establishing as “erosional velocity can be used where no specific information as the erosive/corrosive properties of the fluid is available.

1- The velocity above which erosion may occur can be determined by the following empirical equation:

\[ V_e = \frac{c}{(\rho_m)^{0.5}} \quad \text{Eq. 3.18} \]

where
\[ V_e = \text{fluid velocity, feet/second} \]
\[ c = \text{empirical constant} \]
\[ \rho_m = \text{gas / liquid mixture density at flowing pressure and temperature, lbs/ft}^3. \]

Industry experience indicates that for solid-free fluids values of \( c = 100 \) for continuous service and \( c = 125 \) for intermittent service are conservative. For solid-free fluid where corrosion is not anticipated or when corrosion is controlled by inhibition or by employing corrosion resistant alloys, values of \( c = 150 \) to \( 200 \) may be used for continuous service; values up to \( 250 \) have been used for intermittent service.

If solids production is anticipated, fluid velocities should be significantly reduced. Different values of “c” may be used where specific application studies have shown them to be appropriate.

Where solids and/or corrosive contaminants are present or where “c” values higher than 100 for continuous service are used, periodic surveys to assess pipe wall thickness should be considered.

The design of any piping system where solids are anticipated should consider the installation of sand probes, cushions flow tees, and a minimum of three feet of straight piping downstream of choke outlets.

2- The density of the gas/liquid mixture may be calculated using Eq. 3.3

Once \( V_e \) is known, the minimum cross sectional area required to avoid fluid erosion may be determined from the following equation

\[ A = \left[ 9.35 + \left(\frac{Z R_a T}{21.25 P} \right) \right] / V_e \quad \text{Eq. 3.20} \]

Where
\[ A = \text{minimum pipe cross-sectional flow area required, in}^2/1000 \text{ barrels of liquid per day.} \]

b. Minimum Velocity.
If possible, the minimum velocity in two-phase lines should be about 10 feet per second to minimize slugging of separation equipment. This is particularly important in long lines with elevation changes.

c. Pressure Drop.
The pressure drop in a two phase steel piping system may be estimated as described in section 3.5.
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