Two-Phase Flow

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Introduction to Two-Phase Flow - General

- More complex than single-phase flow
- Physical properties such as density and viscosity of the flowing fluid are considered as mixture (gas-liquid)
- Flow velocity in the conduit (pipe or any other geometrical shape) is conceptualized as “Superficial Velocity”
- The basis of flow correlations is single-phase but the further developed correlations have a lot of empirical & experimental data embedded in them
- The concept of “Holdup” is applied for either phase

Definitions and Equations -1

- Superficial Velocity

  Superficial velocity is a hypothetical flow velocity calculated as if the given phase (gas or liquid) were the only one flowing or present in a given cross sectional area. It is unambiguous and well defined compared to true velocity which is spatially dependent.

  For equations refer next slide

Superficial Liquid Velocity

$$\boxed{V_{sL} = \frac{Q_L}{A}}$$

Superficial Gas Velocity

$$\boxed{V_{sg} = \frac{Q_g}{A}}$$

where:

V_{sL} =	superficial liquid velocity, m/s
V_{sg} =	superficial gas velocity, m/s
Q_L =	liquid volumetric flow rate at flowing conditions, m^3/h
Q_g =	gas volumetric flow rate at flowing conditions, m^3/h
A =	pipe cross-sectional area, m^2

$$\boxed{A = \frac{\pi \times D^2}{4}}$$

where:

D = internal diameter of pipe, m

Superficial Mixture Velocity

$$\boxed{V_m = V_{sL} + V_{sg}}$$

where:

V_m = mixture velocity, m/s

Definitions and Equations -2

Equations for Superficial Velocity

Definitions and Equations - 3

- **Holdup**
  - ➢ Liquid Holdup, H_L

    It is defined as the fraction of the pipe's cross-sectional area occupied by liquid

    $$\boxed{H_L = \frac{A_L}{A}}$$

  - ➢ Gas Holdup, H_G

    The Gas holdup is given by

    $$\boxed{H_G = \frac{A_G}{A}}$$

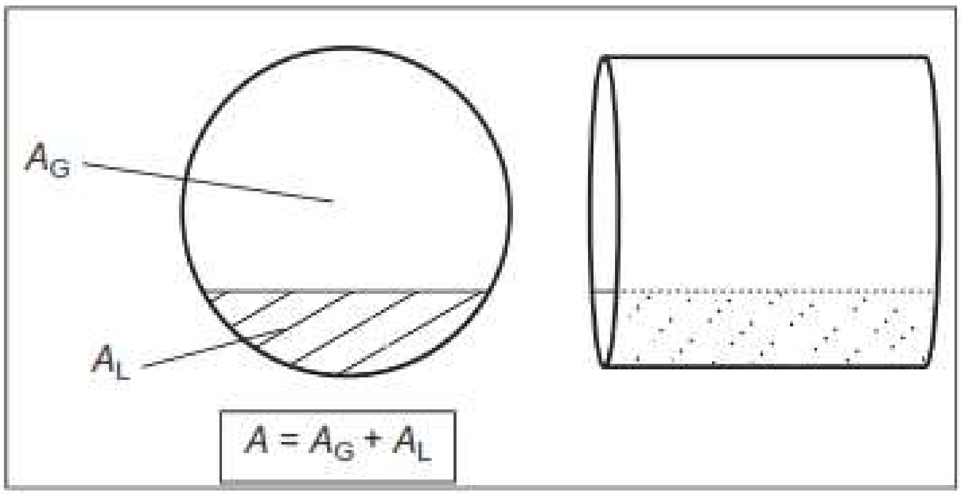
    $$\boxed{A = A_L + A_G}$$

where:

A_L = Cross-sectional area of the pipe occupied by liquid

A_G = Cross-sectional area of the pipe occupied by gas

A = Total cross-sectional area of the pipe



Definitions and Equations - 4

Definitions and Equations - 5

- Holdup (contd.)
  - Slip Velocity and Liquid Holdup
    - ✓ The gas and liquid are traveling at different velocities
    - ✓ Due to density difference the gas tends to flow faster than liquid
    - ✓ This kind of flow found in practical flow applications of oil/gas flow
  - No-Slip Velocity & Liquid Holdup
    - ✓ Gas and Liquid are traveling at the same velocity
    - ✓ Fluid properties are taken as the average of the gas and liquid phases
    - ✓ Friction factors are calculated using the single phase Moody correlation
    - ✓ For homogeneous flow slip velocity is zero

Definitions and Equations - 6

- Slip Ratio, S
  It is defined as the ratio of the velocity of the gas phase to the velocity of the liquid phase.

- Gas Void Fraction, ε_g
  It is defined as the ratio of the gas volume in a pipeline segment to the whole volume of the pipeline segment.

- Mass Quality
  It is the mass of the gas divided by the total mass in the pipe

For equations refer the next slide

Definitions and Equations - 7

$$V_{slip} = V_g - V_L = \frac{V_{sg}}{1-H_L} = \frac{V_{sL}}{H_L}$$

$$V_g = \frac{Q_g}{A_G}$$

$$V_L = \frac{Q_L}{A_L}$$

$$H_{L(no-slip)} = \lambda = \frac{V_{sL}}{V_{sL}+V_{sg}} = \frac{Q_L}{Q_L+Q_g}$$

$$S = \frac{V_g}{V_L}$$

For no-slip case

$$V_g = V_L$$

$$\varepsilon_g = \frac{V_{sg}}{(S \times V_{sL}) + V_{sg}} = \frac{\rho_L \times x}{(S \times \rho_g \times (1-x)) + \rho_L \times x}$$

For the case where S =1 (no-slip case)

$$\varepsilon_g = \lambda$$

where:

V_{slip} = slip velocity

V_g = Average Gas Velocity

V_L = Average Liquid Velocity

Q_L = Liquid Volumetric flow rate

Q_g = Gas Volumetric flow rate

λ = Flowing Liquid Volume fraction

Note: This is also equal to the no-slip liquid holdup $H_{L(no-slip)}$

S = Slip Ratio

ε_g = gas void fraction

ρ_L = liquid phase density

ρ_g = gas phase density

x = mass quality

$$x = \frac{Q_g \times \rho_g}{(Q_L \times \rho_L) + (Q_g \times \rho_g)}$$

Definitions and Equations - 8

- Mixture Viscosity

$$\mu_n = (\mu_L \times \lambda) + \mu_g \times (1-\lambda)$$

- Mixture Density

$$\rho_k = \frac{\rho_L \times \lambda^2}{H_{Ld}} + \frac{\rho_g \times (1-\lambda)^2}{1-H_{Ld}}$$

Where,

μ_n = mixture dynamic viscosity

ρ_k = mixture density

λ = flowing liquid volume fraction, $\lambda = \frac{Q_L}{Q_L + Q_g}$

$H_{\mathbf{Ld}}$ = Liquid Holdup calculated based on Dukler's correlation

Definitions and Equations - 9

- Mixture Reynolds Number, Re_y

$$Re_y = \frac{0.001 \times \rho_k \times V_m \times d}{\mu_n}$$

Metric Units

$$Re_y = \frac{1488 \times \rho_k \times V_m \times d}{\mu_n}$$

US Customary Units

Where,

ρ_k = Mixture Density, kg/m^3 (lb/ft^3)

V_m = Mixture Velocity, m/s (ft/sec)

d = Pipe inside diameter, mm (inches)

μ_n = Mixture Dynamic Viscosity, Pa.s (cP)

Definitions & Equations – 10

- Generalized Pressure Drop Equation (Two-Phase Flow)

$$\boxed{-\frac{dP}{dL} = \left(\frac{g}{g_c} \times \rho_m \times \operatorname{Sin}\theta\right) + \left(\frac{f_m \times \rho_m \times V_m^{\ 2}}{2 \times g_c \times d}\right) + \left(\frac{\rho_m \times V_m \times dV_m}{g_c \times dL}\right)}$$

Where,

f_m = 2-phase friction factor (specific to the flow correlation used)

ρ_m = mixture density (specific to the flow correlation)

Definitions and Equations -11

- Total Pressure drop in 2-phase flow is the sum of the frictional pressure drop, the acceleration pressure drop and the gravitational pressure drop

$$\boxed{\Delta P_{total} = \Delta P_f + \Delta P_a + \Delta P_{grav}}$$

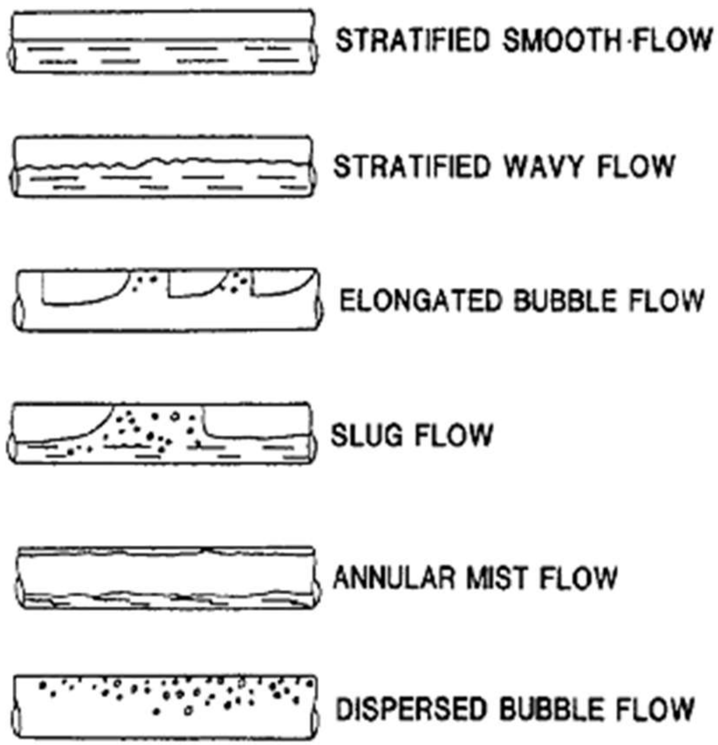
- For horizontal pipelines the gravitational pressure drop is zero
- Acceleration pressure drop can be neglected in adiabatic flow

  Note: Empirical two-phase flow correlations based on laboratory experiments on air-water two-phase flow are under adiabatic flow conditions which mimic the flow of oil-gas mixture flow in field pipelines such as flowlines

Definitions and Equations -12

- Non-Adiabatic (Diabatic) Two-Phase Flow
  - These involving heat transfer between the phases and examples are flow of boiling or condensing two-phase vapor-liquid flows
  - The mechanism of two-phase flow in boiling and condensing fluids is complex and beyond the scope of this presentation

Flow Regimes: Horizontal Pipe Flow-1



Two-phase Flow Patterns in Horizontal Flow (Source: P. Griffith, "Multiphase Flow in Pipes," JPT, March 1984, pp. 363-367)

Flow Regimes: Horizontal Pipe Flow-2

- Bubble Flow

  Very low gas-liquid ratios. Gas bubbles rise to the top
- Elongated Bubble

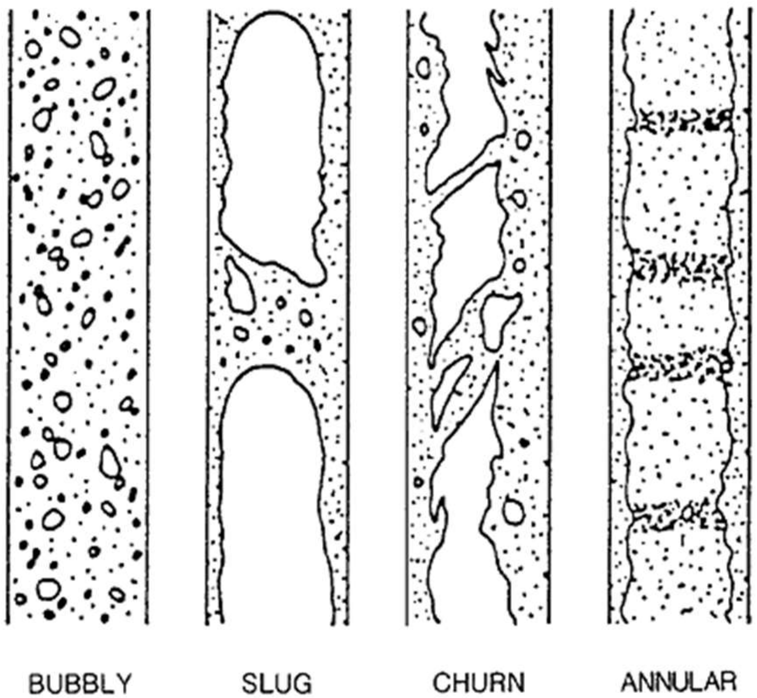
  With increasing gas-liquid ratios, bubbles become larger and form gas plugs.
- Stratified

  Further increase in gas-liquid ratios make the plugs become longer until the gas and liquid are in separate layers.
- Wavy

  As the gas rate increases, the flowing gas causes waves in the flowing liquid.
- Slug

  At even higher gas rates, the waves touch the top of the pipe, trapping gas slugs between wave crests. The length of these slugs can be several hundred feet long in some cases.
- Annular Mist

  At extremely high gas-liquid ratios, the liquid is dispersed into the flowing gas stream.



Two-phase Flow Patterns in Vertical Flow
(Source: J.P. Brill, "Multiphase Flow in Wells,"
JPT, January 1987, pp. 15-21)

Flow Regimes: Vertical Pipe Flow -1

Flow Regimes: Vertical Pipe Flow -2

- Bubble

  Small gas-liquid ratio with gas present in small, randomly distributed bubbles. The liquid moves up at a uniform velocity. Gas phase has little effect on pressure gradient.

- Slug Flow

  The gas phase is more pronounced. Although the liquid phase is still continuous, the gas bubbles coalesce into stable bubbles of the same size and shape, which are nearly the diameter of the pipe. These bubbles are separated by slugs of liquid. Both phases have a significant effect on the pressure gradient.

Flow Regimes: Vertical Pipe Flow -3

- Transition or Churn Flow

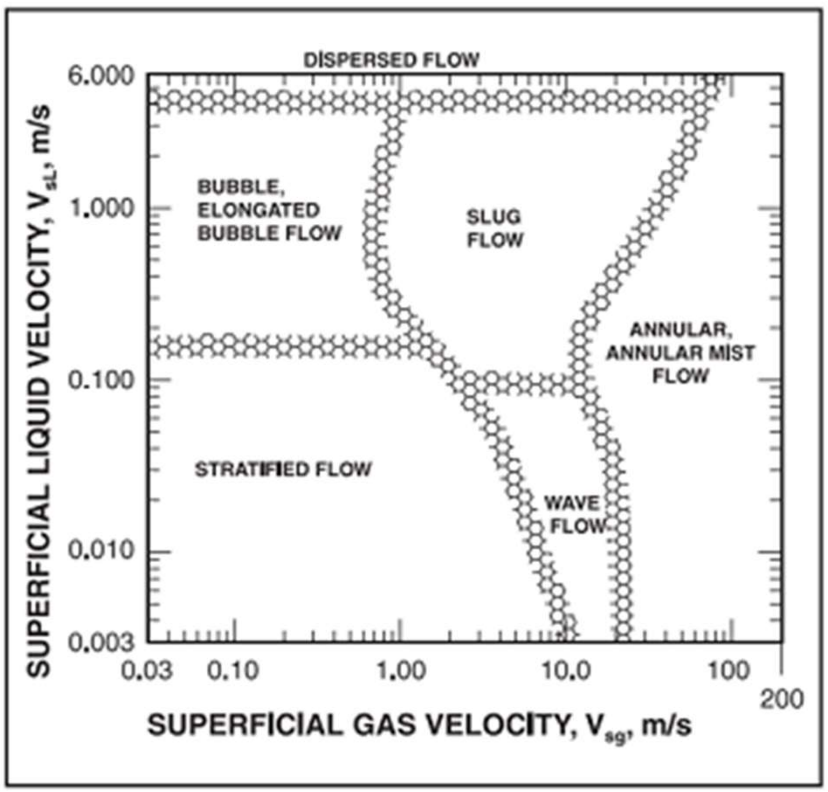
  The change from a continuous liquid phase to a continuous gas phase occurs in this region. The gas phase is pre-dominant and the liquid becomes entrained in the gas. The effects of the liquid are still significant.

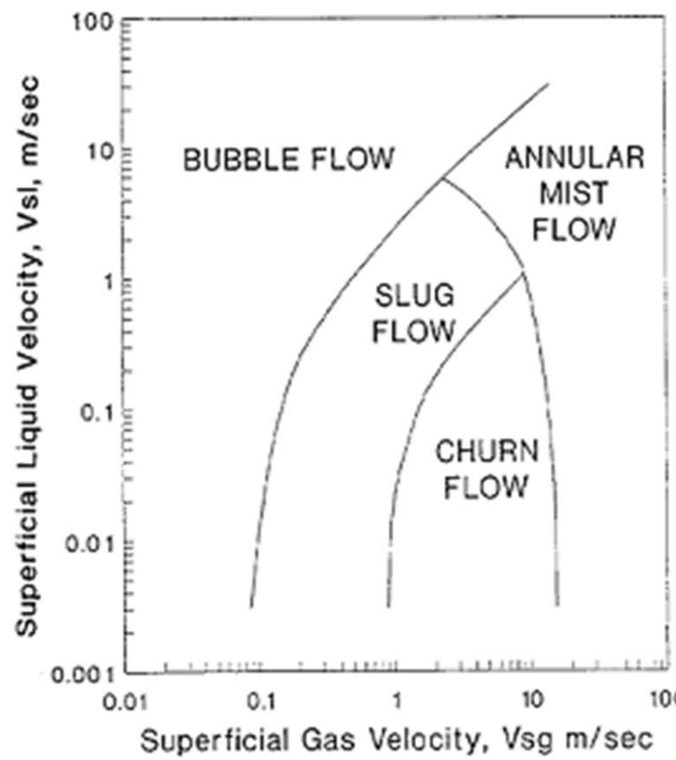
- Annular Mist Flow

  The gas phase is continuous and the bulk of the liquid is entrained in and carried by the gas. A film of liquid wets the pipe wall and its effects are secondary. The gas phase is the controlling factor.

Flow Regime Maps – Horizontal and Vertical

Horizontal Flow Regime Map

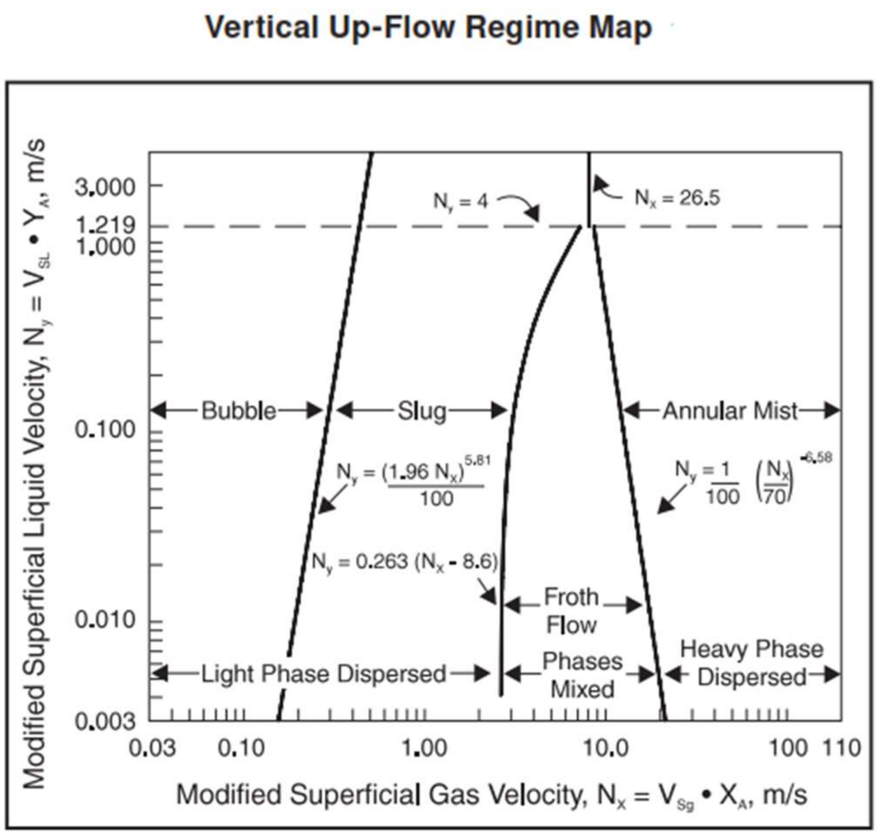




Vertical Multiphase Flow Map

Aziz Map for Vertical Up-flow Regime

- For pipe inclinations greater than 10-20 degrees, flow regime patterns resemble those of vertical flow more than those of horizontal flow, and the Aziz vertical map should be used.



$$N_x = V_{sg} X_A$$

$$N_y = V_{sL} Y_A$$

$$X_A = \left(\frac{\rho_g}{\rho_a}\right)^{0.333} Y_A$$

$$Y_A = \left(\frac{\rho_L \sigma_{wa}}{\rho_w \sigma}\right)^{0.25}$$

where:

N_x =	Horizontal coordinate of Aziz Map, m/s
N_y =	Vertical coordinate of Aziz Map, m/s
X_A =	Aziz fluid property correction factor (horizontal axis)
Y_A =	Aziz fluid property correction factor (vertical axis)
ρ_g =	gas density, kg/m^3
ρ_a =	air density at 15°C and 1.22 kg/m^3
ρ_L =	liquid density, kg/m^3
σ_{wa} =	interfacial tension of air and water at 15°C and 101.56 kPa (abs), 0.0724 N/m
ρ_w =	water density at 15°C and 101.56 kPa (abs), 999.5 kg/m^3
σ =	interfacial tension at flowing conditions, N/m

Calculating Flow Regime based on Aziz Map

Example Calculation: Flow Regime for Aziz Map

Example Calculation for Vertical Upward Flow Flow Regime

Inputs

Q_L =	17.3	m^3/h
Q_g =	51	m^3/h
ρ_L =	832.8	kg/m^3
ρ_g =	32	kg/m^3
ρ_w =	999.5	kg/m^3
ρ_a =	1.224	kg/m^3
σ =	0.02	N/m
d =	200	mm
σ_{wa} =	0.0724	N/m

Outputs

Y_A =	1.32	
X_A =	3.9	
D =	0.2	m
V_{sg} =	0.451	m/s
V_{sL} =	0.153	m/s
N_x =	1.76	m/s
N_y =	0.202	m/s

Figure shows flow is in slug flow regime

Common Flow Correlations for Two Phase Flow

Horizontal Flow	**Eaton-Flanigan**	This correlation is a hybrid correlation of the Eaton hold-up and friction loss correlations and the Flanigan inclined pipe correlation
	Eaton-Dukler - Flanigan	This correlation is another hybrid correlation of the Eaton hold-up correlation, the Dukler friction correlation, and the Flanigan inclined pipe correlation.
	Beggs and Brill (no slip)	The no slip assumption is only applicable in flow regimes where liquid and gas velocities are the same. One of the few multi-phase flow correlations capable of modeling vertical, inclined, or horizontal flow. Assumes smooth pipe.
	Beggs and Brill (with Darcy-Weisbach friction factor)	The no slip assumption is only applicable in flow regimes where liquid and gas velocities are the same. One of the few multi-phase flow correlations capable of modeling vertical, inclined, or horizontal flow. Pipe is allowed to include roughness.

- Horizontal Flow

Common Flow Correlations for Two Phase Flow

Vertical Flow	**Fancher and Brown** (no slip and no flow pattern map)	The no slip assumption and no pattern map imply that the correlation is not generally applicable. The no slip assumption is only applicable in flow regimes where liquid and gas velocities are the same.
	Hagedorn and Brown	Developed from experiments on 1,500 ft experimental well using 1 inch to 4 inch tubing. Experiments included thre-phase flow. One of the most commonly used multi-phase flow correlations for vertical or near vertical wells.
	Beggs and Brill (no slip)	The no slip assumption is only applicable in flow regimes where liquid and gas velocities are the same. One of the few multi-phase flow corrections capable of modeling vertical, inclined, or horizontal flow. Assumes smooth pipe.
	Beggs and Brill (with Darcy-Weisbach friction factor)	The no slip assumption is only applicable in flow regimes where liquid and gas velocities are the same. One of the few multi-phase flow corrections capable of modeling vertical, inclined, or horizontal flow. Pipe is allowed to include roughness.
	Orkiszewski	Took existing correlations and compared them to field results. Selected the best correlations for different regimes and developed a single correlation. This is apopular multi-phase flow correlation, but may exhibit discontinuities when crossing regime boundaries.
	Gray	Developed for gas condensate reservoirs (most accurate for these reservoirs). Uses non-compositional approach. It is based on the observation that hold-up is not as great in condensate wells as in oil wells. Roughness is ignored, but uses an efficiency instead.
	Gray (with Darcy-Weiabach friction factor)	Developed for gas condensate reservoirs (most accurate for these reservoirs). Similar to the standard Gray correlation, but roughness is incorporated through the Moody Diagram.
	Duns and Ros	Uses combined experimental and field measurements. The first multi-phase flow correlation to use flow pattern mapping. A popular multi-phase flow correlation.

- Vertical Flow

Aspen HYSYS Correlations for Two Phase Flow

Correlation	Pipe Geometry Applicability			Category
	Hoizontal	Inclined	Vertical	
Tulsa Unified Model	O	O	O	Mechanistic
OLGAS	O	O	O	Mechanistic
Beggs & Brill (1973)	O	O	O	Empirical
Beggs & Brill (1979)	O	O	O	Empirical
Gregory et al.	O			Empirical
HTFS	O	O	O	Mechanistic
Aziz et al.			O	Empirical
Duns & Ros			O	Empirical
Orkiszewski			O	Empirical
Hagedorn & Brown			O	Empirical
Poettmann & Carpenter			O	Empirical
Baxendall & Thomas			O	Empirical
Lockhart & Martinelli	O			Empirical
Dukler	O			Empirical

Flow correlations in Aspen HYSYS

Flow Regime, Liquid Holdup, Frictional Pressure Drop, Elevation Pressure Drop & Acceleration Drop Eqns for various Two-Phase flow correlations

- Reference:

  Table 3.5, Chapter 3, “Natural Gas Processing Technology and Engineering Design” – by Alireza Bahadori, Ph.D., Elseiver

Further Recommendations on Two Phase Flow Correlations Applicability -1

- Gas-dominated (high GLR) with subcritical Flow
  - Duckler-Eaton correlation
  - Low Liquid loadings (0.056 m^3/1000 Sm^3) require to be bracketed with Beggs, Brill & Moody
  - For $0.1<H_L<0.35$ Mukherjee-Brill provides good results
- Crude Oil (low GLR)

  Beggs, Brill & Moody
- Two-Phase Fluids / Steam with downward flow

  Beggs and Brill No-Slip correlation
- Wet Steam (except downward flow)

  Beggs, Brill and Moody
- High Velocity / Critical Flow Systems

  High-velocity modifications to the standard Beggs and Brill, and Beggs, Brill, and Moody correlations.

Further Recommendations on Two Phase Flow Correlations Applicability -2

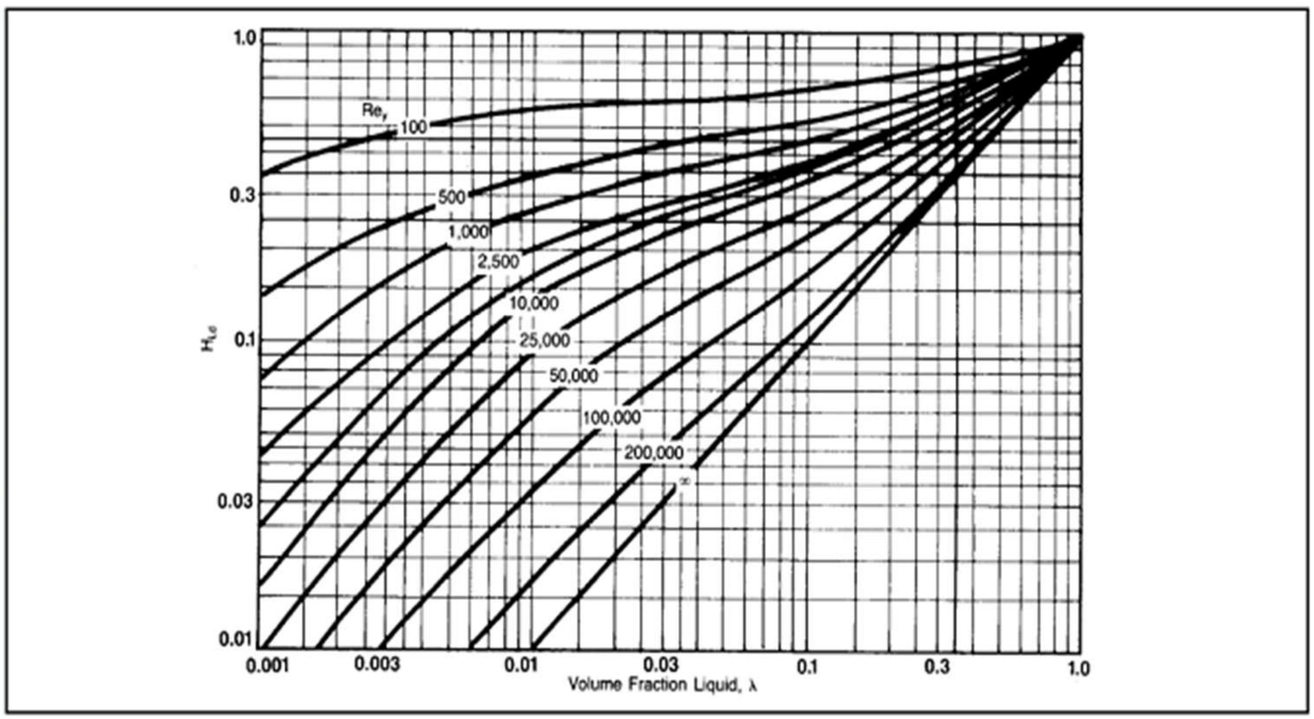
Recommendations for Two-phase Pressure Drop Correlations	
Correlation	**Correlation Recommendations**
Horizontal Flow	
Lockhart–Martinelli	Widely used in the chemical industry. Applicable for annular and annular mist flow regimes if flow pattern is known a priori. Do not use for large pipes. Generally overpredicts pressure drop
Eaton	Do not use for diameters <2 in. Do not use for very high or low liquid holdup. Underpredicts holdup for $H_L < 0.1$. Works well for $0.1 < H_L < 0.35$
Dukler	Good for horizontal flow. Tends to underpredict pressure drop and holdup. Recommended by API for wet gas lines
Beggs and Brill	Use the no-slip option for low holdup. Underpredicts holdup. Most consistent and well-behaved correlation
Inclined Flow	
Mukherjee–Brill	Recommended for hilly terrain pipelines. New correlation based heavily on in situ flow pattern. Only available model that calculates flow patterns for all flow configurations and uses this information to determine modeling technique

2-Phase Friction Pressure Drop AGA (Dukler)-1

- Liquid Holdup for Dukler Pressure Drop

FIG. 17-18 (GPSA Engineering Databook, 13th Edition)

Liquid Holdup Correlation



2-Phase Friction Pressure Drop AGA (Dukler)-2

- Single Phase Friction Factor, f_n

$$\boxed{f_n = 0.0056 + 0.5 \times \mathrm{Re}_y^{-0.32}}$$

where,

Re_y is calculated as per "Definitions and Equations-9"

2-Phase Friction Pressure Drop AGA (Dukler)-3

- Friction Factor Ratio, f_{tpr}

$$f_{tpr} = 1 + \left[\frac{y}{1.281 - 0.478y + 0.444y^2 - 0.094y^3 + 0.00843y^4}\right]$$

where:

$y = -\ln(\lambda)$

λ = flowing liquid volume fraction

$$\lambda = \frac{Q_L}{Q_L + Q_g}$$

2-Phase Friction Pressure Drop AGA (Dukler)-4

- Frictional Pressure Drop

$$\Delta P_f = \frac{f_n \times f_{tpr} \times \rho_k \times V_m^{\,2} \times L_m}{2 \times D}$$

where,

ΔP_f = frictional pressure drop, kPa

f_n, f_{tpr} = as defined in *“2-Phase Friction Pressure Drop AGA (Dukler)-2 and 3”*

ρ_k = as defined in slide *“Definitions and Equations-8”*, kg/m^3

$V_m = (V_{sL} + V_{sg})$, m/s

L_m = Pipe length, km

D = Pipe internal diameter, m

2-Phase Friction Pressure Drop AGA (Dukler)-5

- Problem Statement

  A pipeline segment with a 6-inch NPS (152.4 mm ID), 1.2 km long, transports a mixture of gas and oil. The pipeline has a gradual upward slope and rises 30 m over the 1.2 km length. The inlet pressure of the pipeline is 2800 kPa (abs), liquid viscosity is 0.02 Pa.s, the vapor viscosity is 0.000015 Pa.s, and the interfacial surface tension is 1.5 x 10^{-6} N/m. The liquid flow rate is 17 m^3/h and the vapor flow rate is 425 m^3/h. The density of the liquid phase is 880 kg/m^3, and the density of the gas phase is 20.8 kg/m^3 at operating conditions. What is the pressure at the downstream end of the line segment, and what is the liquid inventory of the line?

2-Phase Friction Pressure Drop AGA (Dukler)-6

- Calculations (Horizontal Line)

Inputs		
P_1 =	2800	kPa(abs)
Q_L =	17	m^3/h
Q_g =	425	m^3/h
ρ_L =	880	kg/m^3
ρ_g =	20.8	kg/m^3
μ_L =	0.02	Pa.s
μ_g =	0.000015	Pa.s
σ =	1.5E-06	N/m
L_m =	1.2	km
d =	152.4	mm

2-Phase Friction Pressure Drop AGA (Dukler)-7

Outputs			
D =	0.1524	m	(d/1000)
A =	0.01824	m^2	
V_{sL} =	0.259	m/s	
V_{sg} =	6.472	m/s	
λ =	0.0385		
μ_n =	0.000784	Pa.s	
First Iteration for Liquid Holdup (Assume $H_{Ld} = \lambda$)			
H_{Ld} =	0.0385		
ρ_k =	53.85	kg/m^3	
V_m =	6.73	m/s	
Re_y =	70482		

From Fig 17-18 get a better estimate of H_{Ld}		
for 2nd iteration		
H_{Ld} =	0.12	
ρ_k =	32.70	kg/m^3
Re_y =	42804	
From Fig 17-18 get a better estimate of H_{Ld}		
for 3rd iteration		
H_{Ld} =	0.16	
ρ_k =	31.03	kg/m^3
Re_y =	40616	

This is within 5% of the calculated Reynolds No from 3rd iteration, hence OK			
f_n =	0.0224		
$y = -\ln(\lambda)$=	3.2581		
f_{tpr} =	2.53		
ΔP_f =	312.5	kPa	(frictional Pressure Drop)

Pressure Drop due to Elevation (Flanigan Correlation)-1

Elevation component of pressure drop (Flanigan Correlation)

The horizontal line as given in the problem statement is having an upward elevation of 30 m from the horiontal. Essentially it means that if we assume the line elevation at the inlet as 0 meters, the outlet of the line is 30 m from inlet (datum being 0 m)

$$\boxed{\Delta P_e = \frac{\rho_L \times H_{Lf}}{100} \times \Sigma Z_e}$$

where:

ΔP_e = elevation component of pressure drop, kPa

ΣZ_e = pipeline vertical elevation rise, m

H_{Lf} = Liquid Holdup fraction (Flanigan), dimensionless

$$\boxed{H_{Lf} = \frac{1}{1 + \left(1.078 \times V_{sg}{}^{1.006}\right)}}$$

- Elevation Component of Pressure Drop

Pressure Drop due to Elevation (Flanigan Correlation)-2

Calculations		
Inputs		
ΣZ_e =	30	m
Outputs		
H_{Lf} =	0.124	
ΔP_e =	31.016	kPa

- Calculations for Elevation Pressure Drop

Total Pressure Drop (Friction + Elevation)

- Total Pressure Drop in the line is the sum of the friction pressure drop (ΔP_f) and the elevation pressure drop (ΔP_e)

$$\Delta P_{total} = \Delta P_f + \Delta P_e$$

ΔP_{total} =	343.5	kPa	
P_2 =	2456.5	kPa (abs)	(pressure at line outlet)

Liquid Holdup for Pipeline Liquid Inventory

Liquid Holdup correlation shown as Fig. 17-18 is meant to be used only for the AGA (Dukler) frictional pressure drop calculation

Correlation by Eaton et al. is better suited for liquid holdup determination in liquid inventory

The Eaton Liquid Holdup (H_{Le}) as a fraction is described in the form of a chart where H_{Le} is plotted directly as a function of a dimensionless group N_E

Eaton Liquid Holdup H_{Le} – Dimensionless group N_E

Eaton Correlation for Liquid Holdup for Low Holdup Flows

$$N_e = \frac{1.84 \times N_{Lv}^{0.575} \times \left(\frac{P_{avg}}{P_b}\right)^{0.05} \times N_L^{0.1}}{N_{gv} \times N_d^{0.0277}}$$

where:

P_{avg} = average pressure, kPa (abs)

$$P_{avg} = \frac{2}{3} \times \left[P_1 + P_2 - \left(\frac{P_1 \times P_2}{P_1 + P_2}\right)\right]$$

P_b = base absolute pressure, kPa (abs)

101.56 kPa(abs)

N_{Lv} = liquid velocity number

$$N_{Lv} = 0.0565 \times V_{sL} \times \left(\frac{\rho_L}{\sigma}\right)^{0.25}$$

N_{gv} = gas velocity number

$$N_{gv} = 0.0565 \times V_{sg} \times \left(\frac{\rho_L}{\sigma}\right)^{0.25}$$

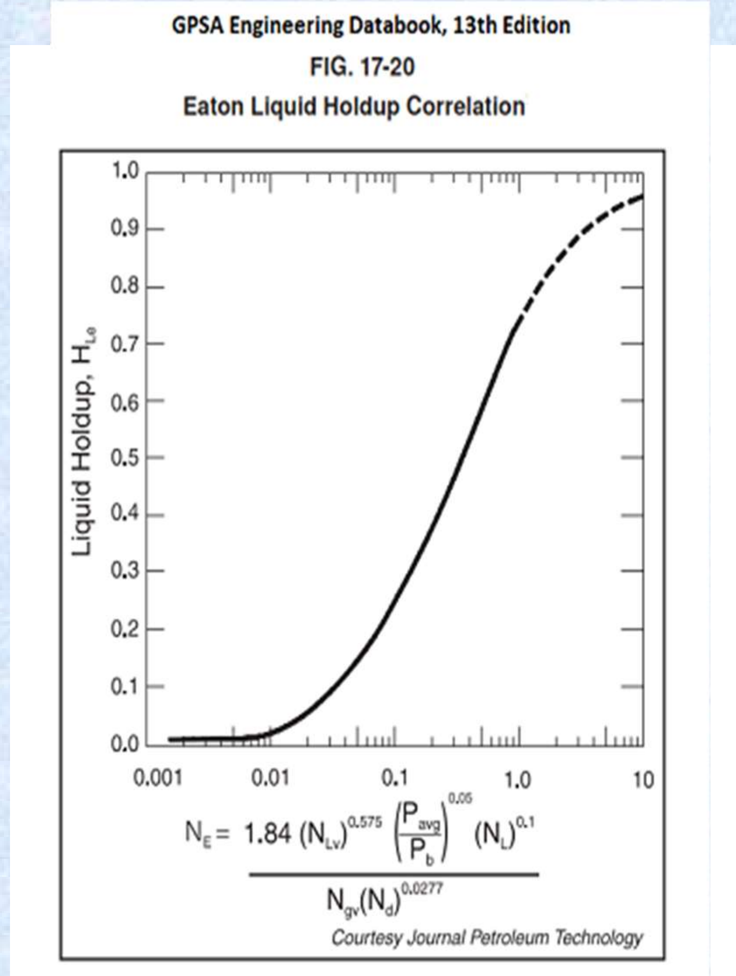
N_d = Pipe diameter number

$$N_d = 0.00003134 \times d \times \left(\frac{\rho_L}{\sigma}\right)^{0.50}$$

N_L = liquid viscosity number

$$N_L = 0.001769 \times \mu_L \times \left(\frac{1}{\rho_L \times \sigma^3}\right)^{0.25}$$

Eaton Liquid Holdup H_{Le} - Chart



Liquid Inventory in Pipe

- The liquid holdup fraction, H_{Le}, is the fraction of the flow area of the pipe occupied by liquid.
- To calculate the liquid inventory in the pipe, I_L, the pipe internal volume is multiplied by this holdup fraction.

$$\boxed{I_L = 7.853 \times 10^{-7} \times H_{Le} \times d^2 \times L}$$

where,

I_L = liquid inventory in pipeline, m^3

L = Length of pipeline, m

$$\boxed{L = L_m \times 1000}$$

H_{Le} = liquid holdup fraction (Eaton)

Eaton Liquid Holdup H_{Le} - Calculations

Inputs			
P_b =	101.56	kPa (abs)	
Outputs			
L_m =	1.2	km	(from Sheet "Presssure_Drop_Dukler_Flanigan)
P_1 =	2800	kPa (abs)	(from Sheet "Presssure_Drop_Dukler_Flanigan)
P_2 =	2456.50	kPa (abs)	(from Sheet "Presssure_Drop_Dukler_Flanigan)
V_{sL} =	0.259	m/s	(from Sheet "Presssure_Drop_Dukler_Flanigan)
V_{sg} =	6.47	m/s	(from Sheet "Presssure_Drop_Dukler_Flanigan)
ρ_L =	880	kg/m^3	(from Sheet "Presssure_Drop_Dukler_Flanigan)
μ_L =	0.02	Pa.s	(from Sheet "Presssure_Drop_Dukler_Flanigan)
σ =	1.5E-06	N.m	(from Sheet "Presssure_Drop_Dukler_Flanigan)
d =	152.4	mm	(from Sheet "Presssure_Drop_Dukler_Flanigan)
L =	1200	m	
P_{avg} =	2632.0	kPa (abs)	
N_{Lv} =	2.28		
N_{gv} =	56.91		
N_d =	115.69		
N_L =	0.152		
N_a =	0.0443		
From Fig 17-20, H_{Le} =	0.14		
l_L =	3.064	m^3	

Slugging in Two-Phase Flow-1

- What are slugs?

  A large (relative) mass of liquid traveling in two-phase flow pipelines at high velocities sometimes interspersed with gas bubbles and many times with definite boundary separation with the gas phase

- Why are slugs and slugging flow considered undesirable?

  - Causes damaging (erosion-corrosion) water hammer when the liquid slug impinges on pipe and equipment walls at every change of flow direction
  - If slug flow enters a distillation column, the alternating composition and density of the gas and liquid slugs cause cycling of composition and pressure gradients along the length of the column. The cycling causes problems with product quality and process control.

Slugging in Two-Phase Flow-2

- Slug Formation Mechanism (Four Mechanisms)
  - Caused by wave formation at the liquid-gas interface in "stratified flow". When liquid waves grow large enough to bridge the entire diameter of the pipe, the stratified flow pattern breaks down resulting in slug flow
  - Terrain change causes slug formation. Liquid collects at a sag in the pipeline and blocks the gas flow. The pressure in this blocked gas rises until it blows the accumulated liquid in the sag out as a slug.
  - Change in pipeline inlet flow. When the inlet flow rate increases, the liquid inventory in the pipeline decreases, and the excess liquid forms a slug or series of slugs.
  - Pigging can cause very large liquid slugs as the entire liquid inventory of the line is swept ahead of the pig.
- Slug Length

Of the four mechanisms described, wave growth normally produces the shortest slugs, followed in length by terrain generated slugs

- Slug Mechanism Studies
  - Wave Induced Slugs - Greskovich and Shrier, Brill et al
  - Terrain Induced Slugs – Schmidt
  - Inlet Flow Rate Induced Slugs – Cunliffe
  - Pigging Induced Slug Dynamics – McDonald & Baker

Slugging in Two-Phase Flow-3

- Remedial Measures to Avoid Slug flow and Slug formation
  - Use minimum pipe diameters to the extent possible utilizing maximum pressure differentials
  - Employ parallel pipeline networks to maintain capacity
  - By providing suitable drain collection system such as a drip vessel at low points in the pipeline system to prevent liquid accumulation
  - Arranging pipeline run in a manner to avoid multiple changes of direction and elevation

Slugging in Two-Phase Flow-4

- Slug Catchers

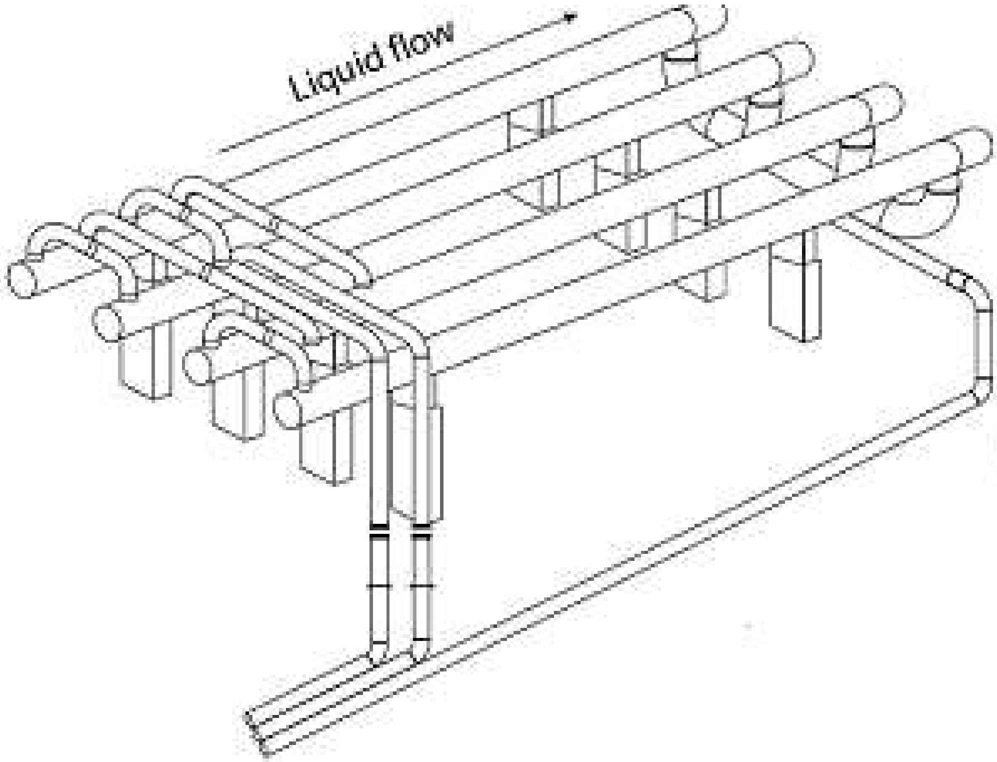
  Devices that are installed at downstream end, or other intermediate points of production or transmission lines used to absorb and hold the fluctuating liquid inlet flow rates caused by liquid slugging

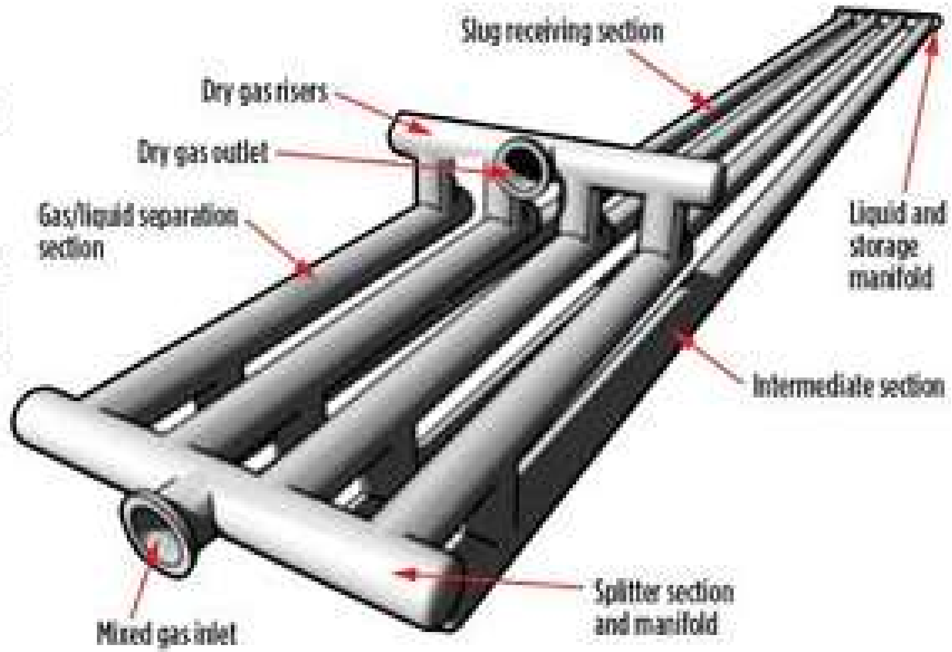
Slugging in Two-Phase Flow-5

- Types of Slug Catchers
  - Harp or Pipe Type
    - ✓ Constructed of multiple lengths of pipe
    - ✓ Treated as part of pipeline and generally designed as per pipeline codes instead of ASME Section VIII (Div. 1 or Div. 2)
    - ✓ Commonly use piping codes are ASME B31.3 and ASME B31.8
    - ✓ Require larger plot plan for installation
    - ✓ Utilized for high volume applications (thousands of barrels)
    - ✓ The upper section is short and consists of two or more pipe sections designed to reduce the gas velocity to provide the necessary separation
    - ✓ Gas flows from the upper section and liquid flows to a lower bank of piping
    - ✓ The lower liquid section consists of multiple downward sloped pipes with sufficient volume to provide storage for the required pipeline slug volume

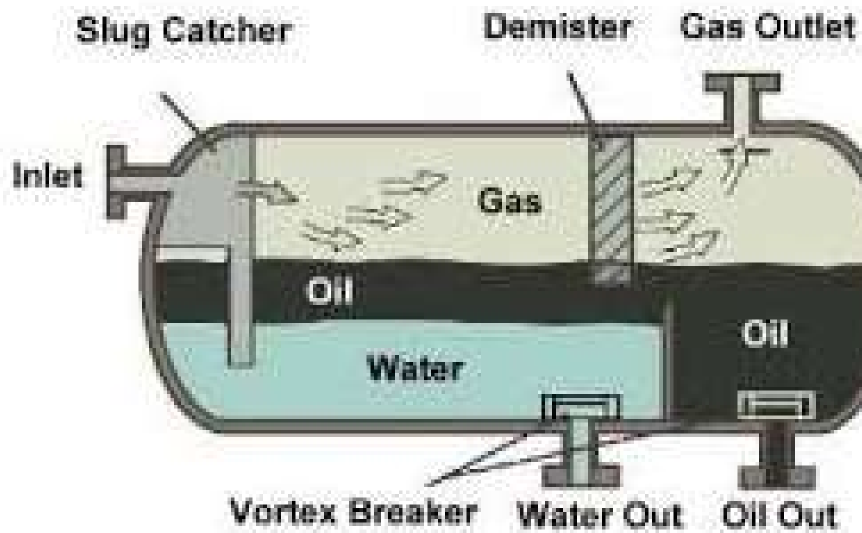
Slugging in Two-Phase Flow-6

- Types of Slug Catchers (contd.)
  - ➢ Vessel Type
    - ✓ Used in lower pressure services (below 3447 kPag)
    - ✓ Employed for smaller slug sizes (<159 m^3)
    - ✓ Designed as per pressure vessel code ASME Section VIII, Div. 1 or Div. 2
    - ✓ Requires smaller plot plan for installation
    - ✓ Has special internals, such as a unique inlet deflection baffle which reduces the momentum of the incoming liquid
    - ✓ A distinct advantage is the ability to incorporate a sand removal system, if required based on inlet fluid characteristics





Slugging in Two-Phase Flow-7: Harp Slug Catcher



Slugging in Two-Phase Flow-8: Vessel Slug Catcher

Slugging in Two-Phase Flow-9

- For a spreadsheet based approximate calculation for slug length and slug volume in a two-phase flow pipeline (US Customary units) refer the link below:

[Pipeline Two Phase Slug Length and Volume Calculator]

Only when I began studying chemical engineering at Oregon Agricultural College did I realize that I myself might discover something new about the nature of the world.

— Linus Pauling —