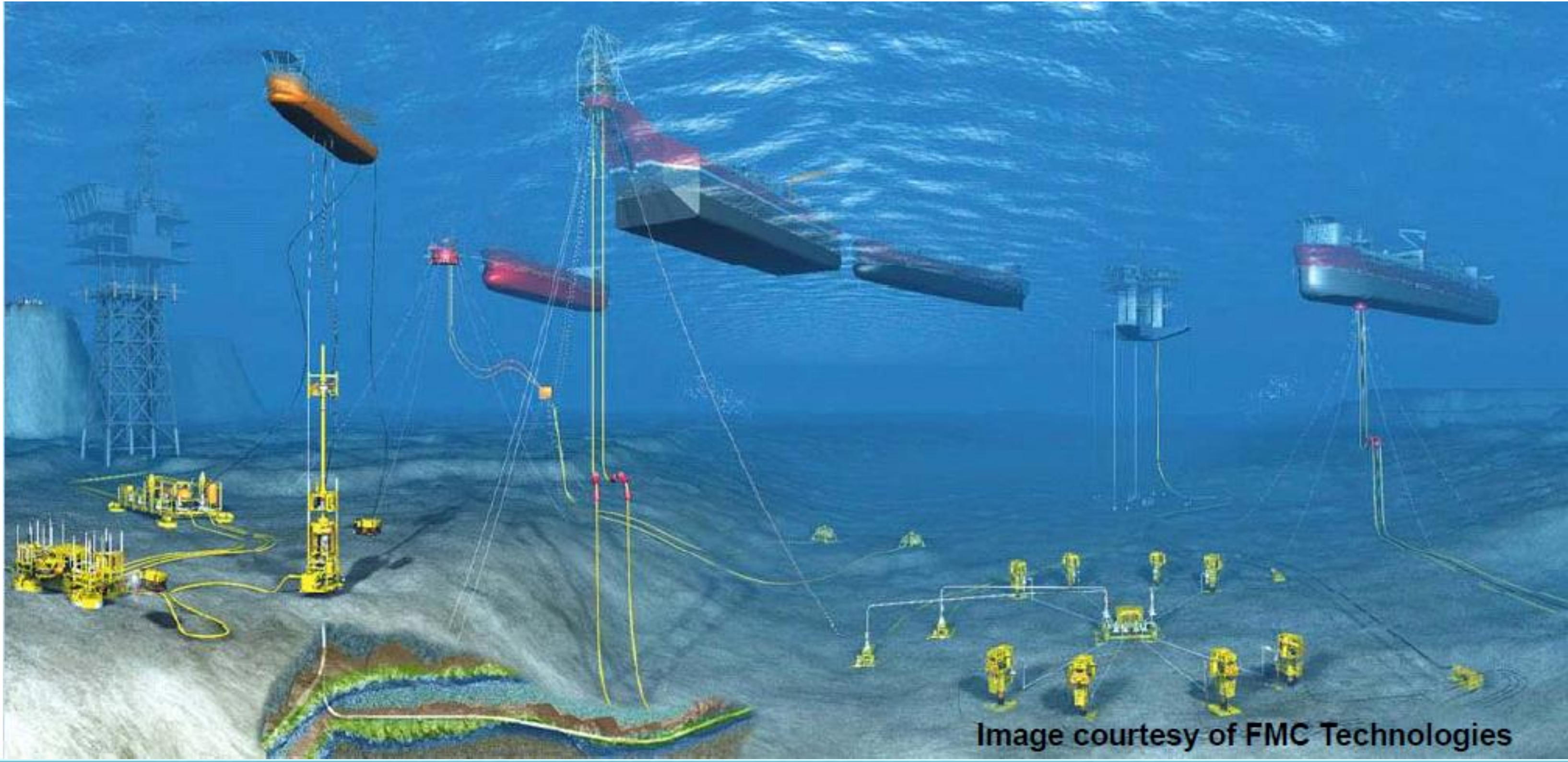
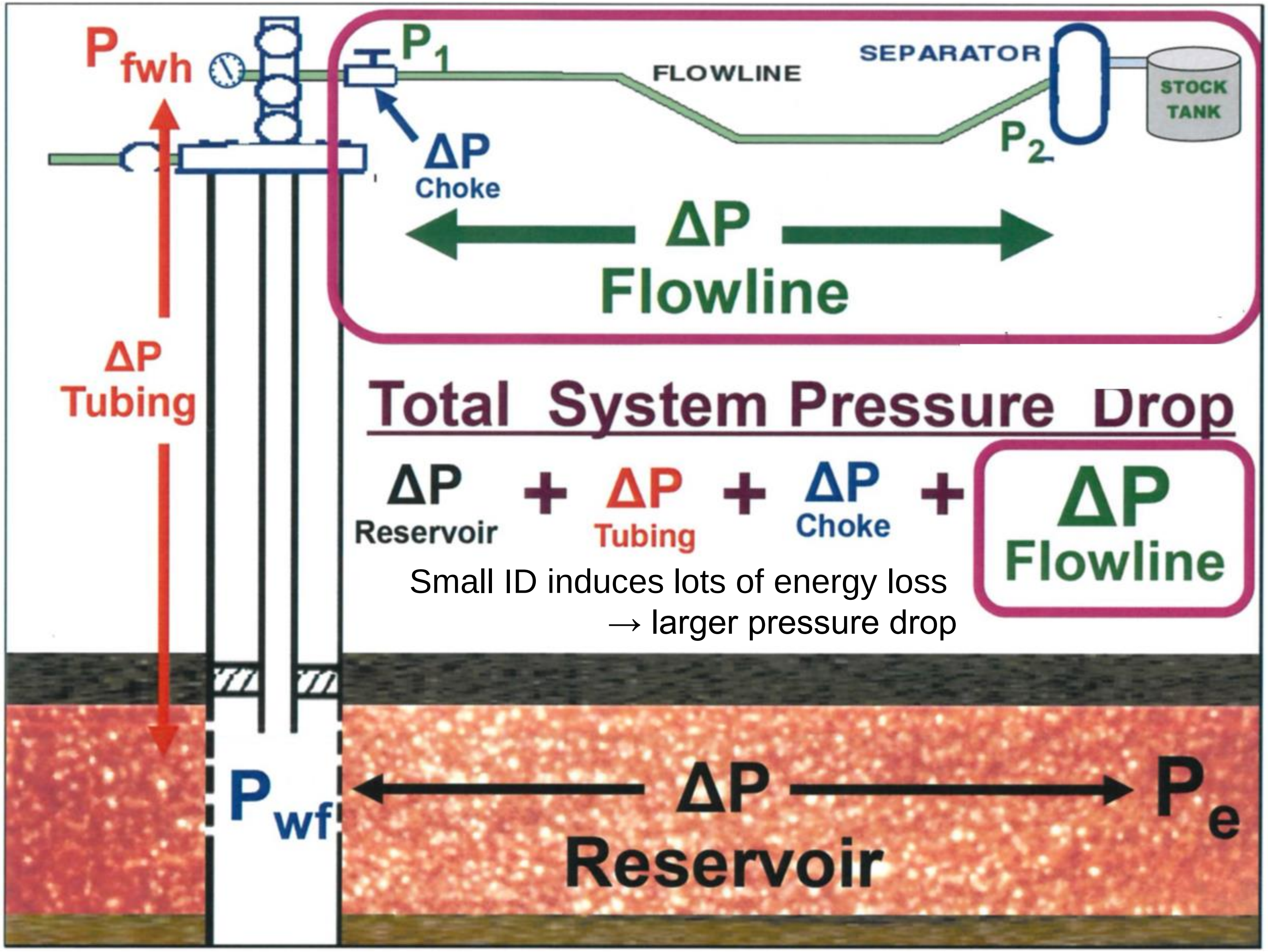


Flow Assurance

서유택

Gathering system

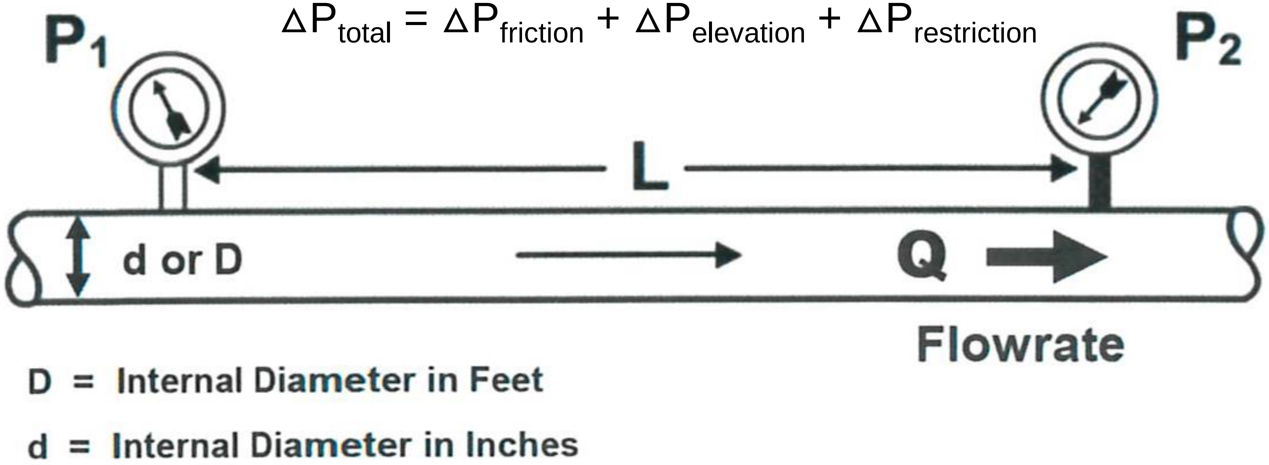


Small ID induces lots of energy loss
→ larger pressure drop

Pressure drop vs. Flowrate in oil field flowlines

$$\Delta P = P_1 - P_2$$

$$\Delta P_{total} = \Delta P_{friction} + \Delta P_{elevation} + \Delta P_{restriction}$$



Darcy – Weisbach Formula

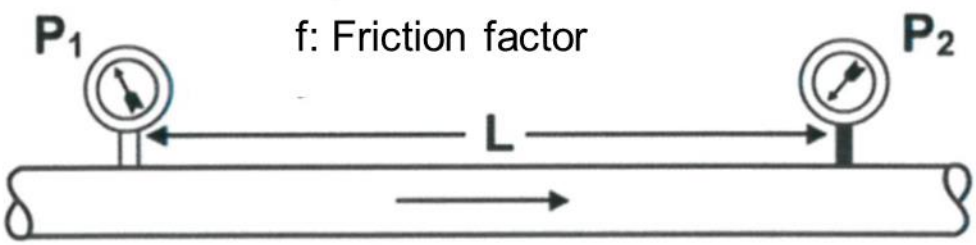
- Pressure drop expressed in feet of fluid head

$$h_{ft} = \frac{f\ L\ v^2}{D\ 2g}$$

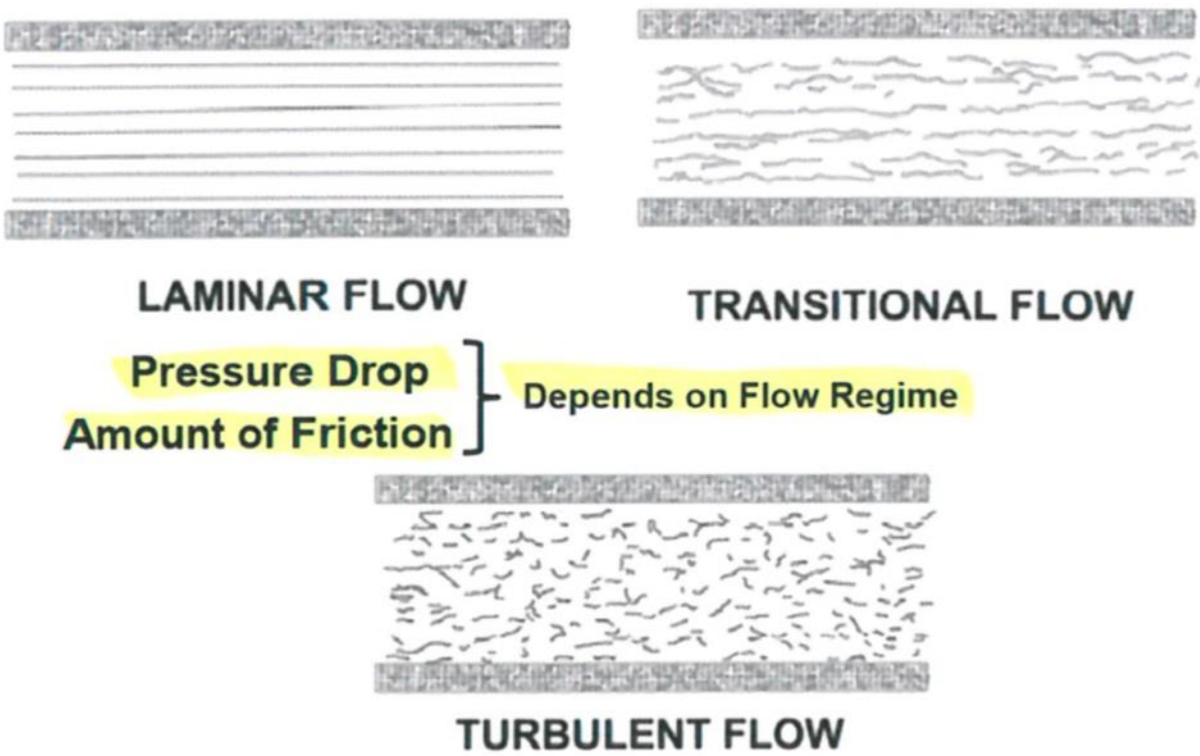
- Pressure drop expressed in psi

$$\Delta\ P = \frac{\rho\ f\ L\ v^2}{144\ D\ 2g}$$

g: correction factor
not gravity acceleration
(= 32.2 ft/s^2 = 9.81 m/s^2)



Flow regime in pipe



- Gas dominant stream is mostly turbulent
- Flow regime determined by Reynolds number

Reynolds number

- Dimensionless parameter

  : Ratio of Inertia forces to Viscous forces

$$R_e = \frac{\rho D v}{\mu_e}$$

ρ : lb/ft^3 D : ft v : ft/sec μ_e : lb/ft - sec

- Re < 2000 = Laminar flow

$$Liquid: Re = 92.1 \frac{SG_L Q_{BPD}}{d\ \mu}$$

$$Gas: Re = 20100 \frac{SG_G Q_{MMCFD}}{d\ \mu}$$

d: inches, μ: centipoise

Friction factor

- f = Dimensionless factor of proportionality

  f_m = Moddy friction factor

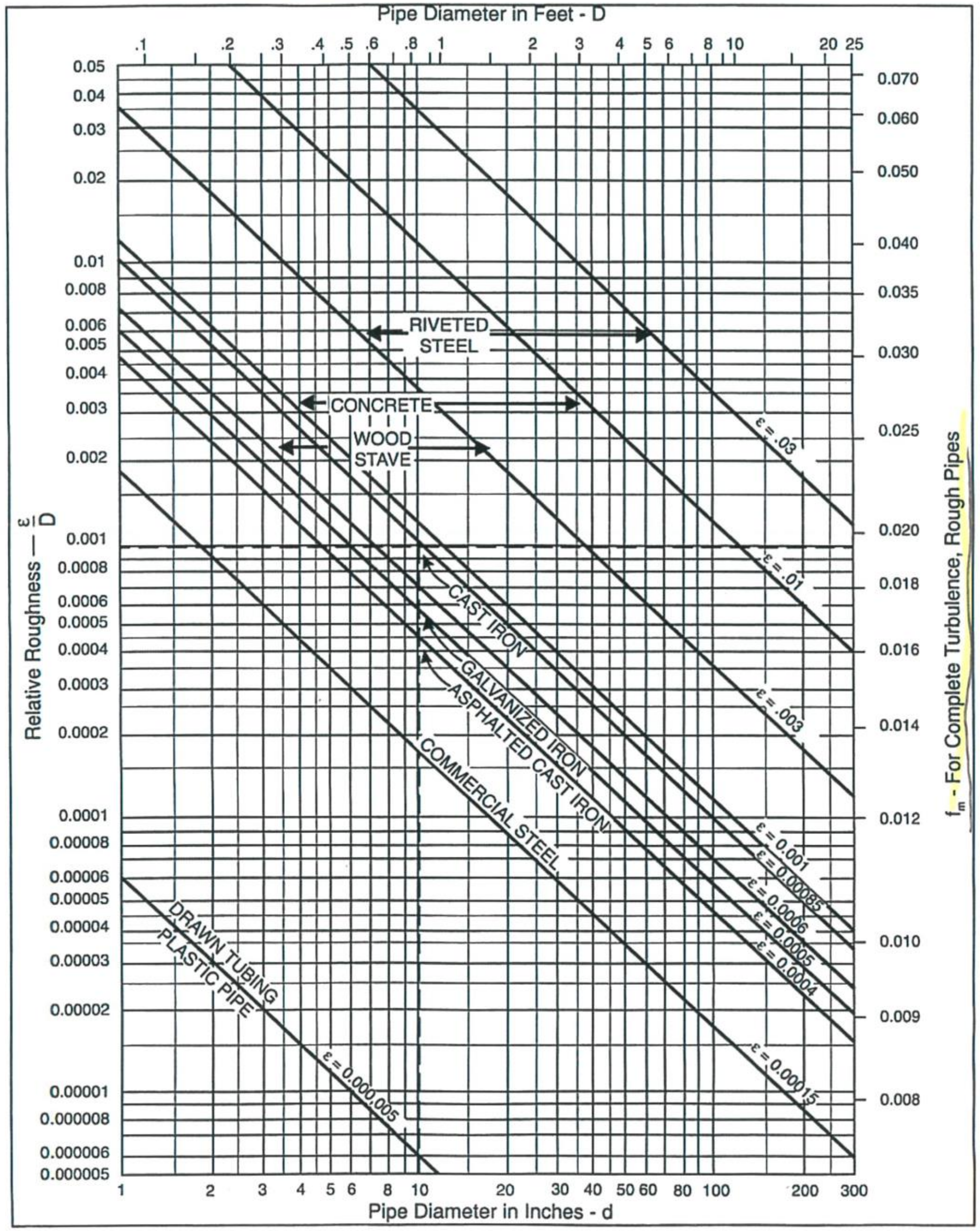
  f_f = Fanning fraction factor (f_f =1/4 f_m)

- Laminar flow: f_m = 64 / Re

- For transitional and turbulent flow
  - f_m a function of Re
  - Relative roughness: ε / D

- For complete turbulence
  - f_m a function of ε / D only

(Do NOT use right hand scale for f_m unless you KNOW flow is FULLY TURBULENT)



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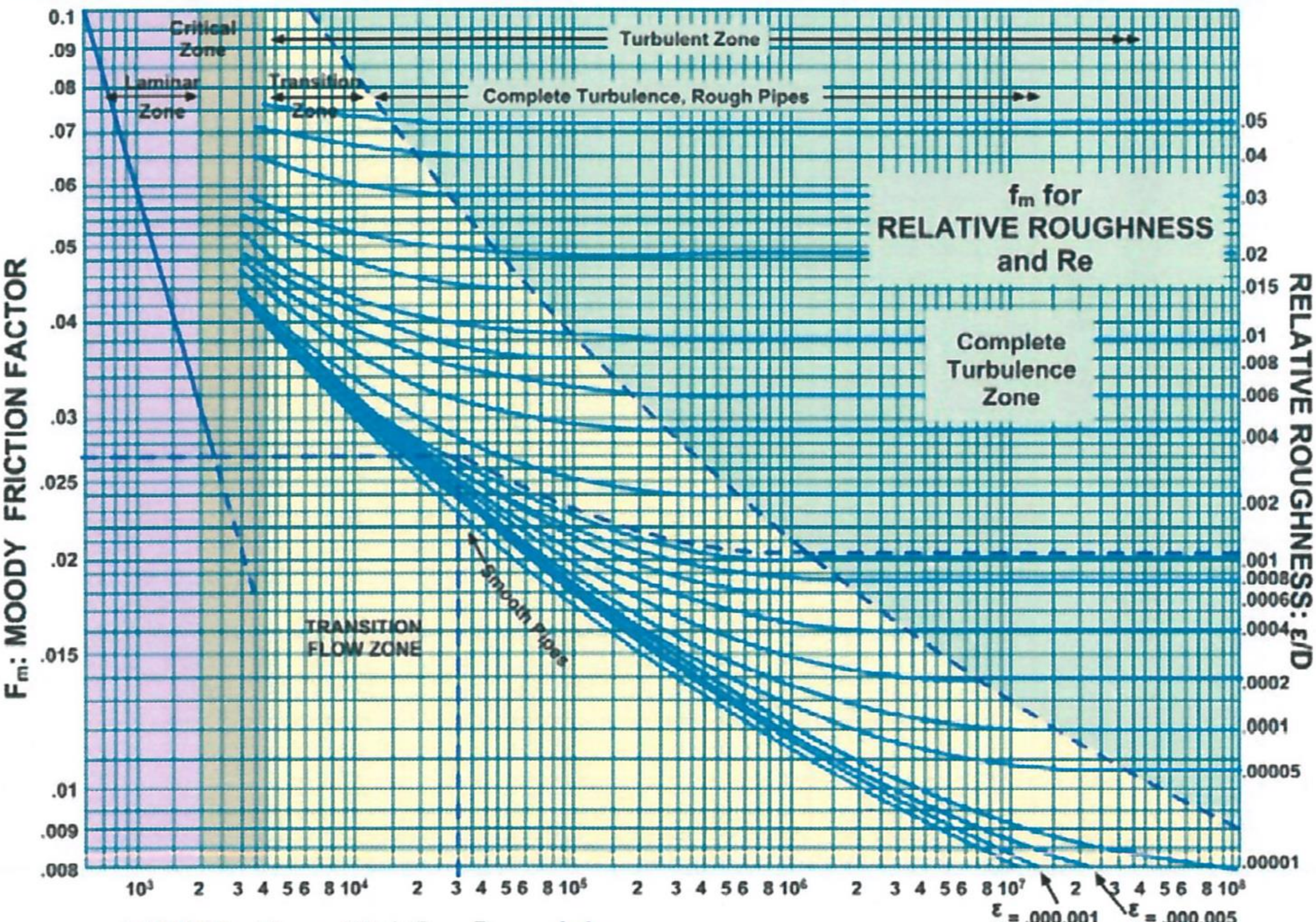
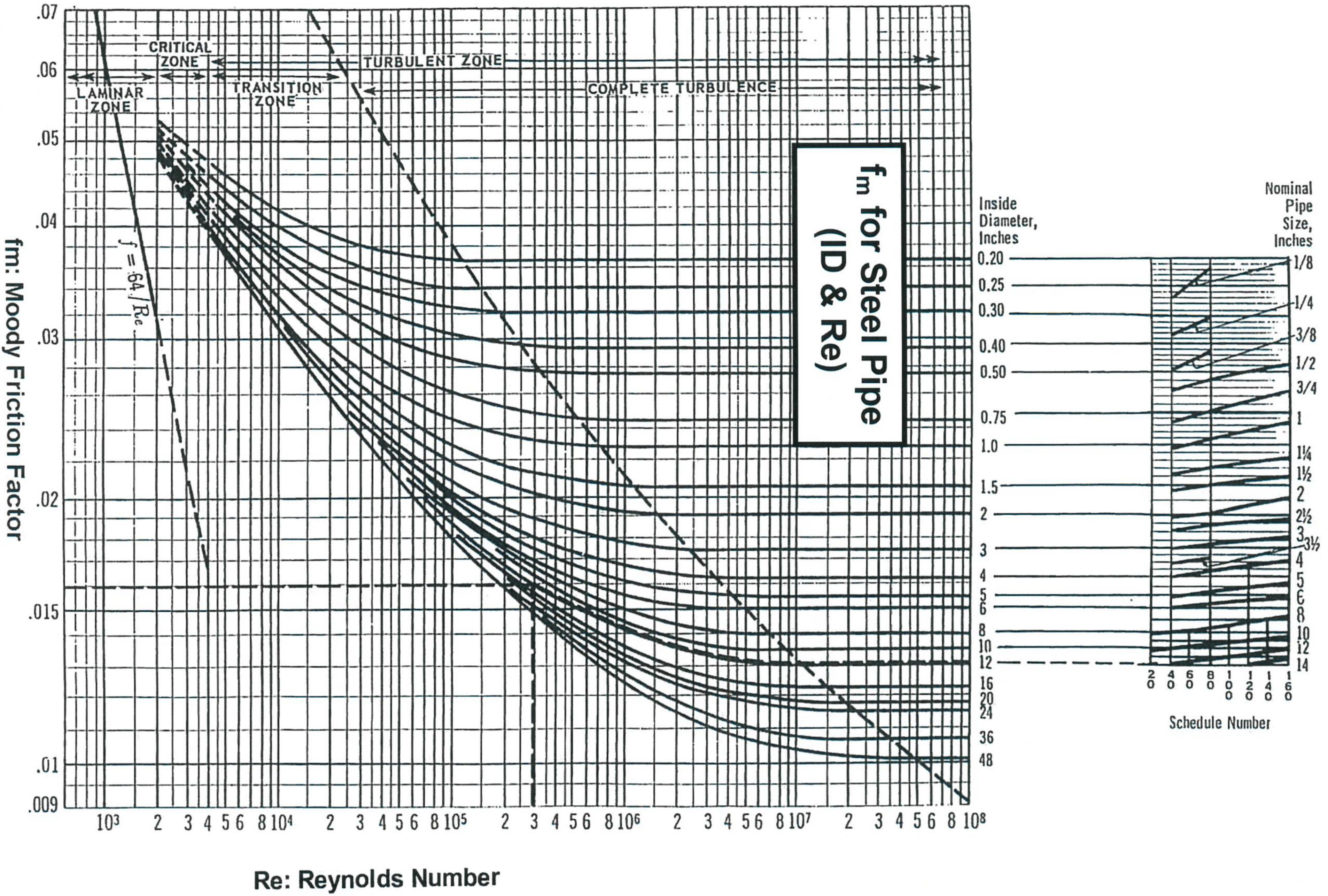
Note: Absolute Roughness units are in feet.
Relative Roughness is Dimensionless.

ε for Steel Pipe = 0. 00015 feet



LIQUID: $Re = 92.1\ S_{GL}\ Q_{BLPD} / d\mu$

GAS: $Re = 20{,}100\ SG_g\ Q_{MMCFD} / d\mu$ Re: REYNOLDS NUMBER


fm: Moody Friction Factor
CRITICAL ZONE
LAMINAR ZONE
TRANSITION ZONE
TURBULENT ZONE
COMPLETE TURBULENCE
f = 64/Re
fm for Steel Pipe (ID & Re)
Inside Diameter, Inches
Nominal Pipe Size, Inches
Schedule Number
Re: Reynolds Number

Pressure drop: Laminar flow (Re < 2000)

- Liquid

$$\Delta P_{psi} = 0.00068 \frac{\mu_{cp} L_{ft} V_{ft/sec}}{d_{in}^2}$$

$$\Delta P_{psi} = 7.95 \times 10^{-6} \frac{\mu_{cp} L_{ft} Q_{BPD}}{d_{in}^4}$$

- Gas

$$\Delta P_{psi} = \frac{0.040\ \mu_{cp} L_{ft} T_{^\circ R} Z Q_{MMCFD}}{P_{psi} d_{in}^4}$$

No "f_m" since f_m = 64/Re and Re = SG_L Q / d μ

Pressure drop: Transitional and Turbulent

- Liquid

$$\Delta P_{psi} = 11.5 \times 10^{-6} \frac{f_m L_{ft} Q_{BPD}^2 SG_L}{d_{in}^5}$$

- Gas

$$P_1^2 - P_2^2 = 25.1 \frac{f_m L_{ft} Q_{MMCFD}^2 SG_G Z T_R}{d_{in}^5}$$

Exercise ΔP: Liquid flow in Pipe

- What is the friction pressure drop in 10,000 ft of 2 inch ID pipe flowing 50 BPD of 35 ^{o}API crude oil (μ=1.2 cp and SG_L=0.85) ?

1. First calculate Reynold’s number to determine flow regime

   :

2. Use the equation for

   $\Delta P_{psi} =$

Exercise: Increasing flow rate 3000 BPD

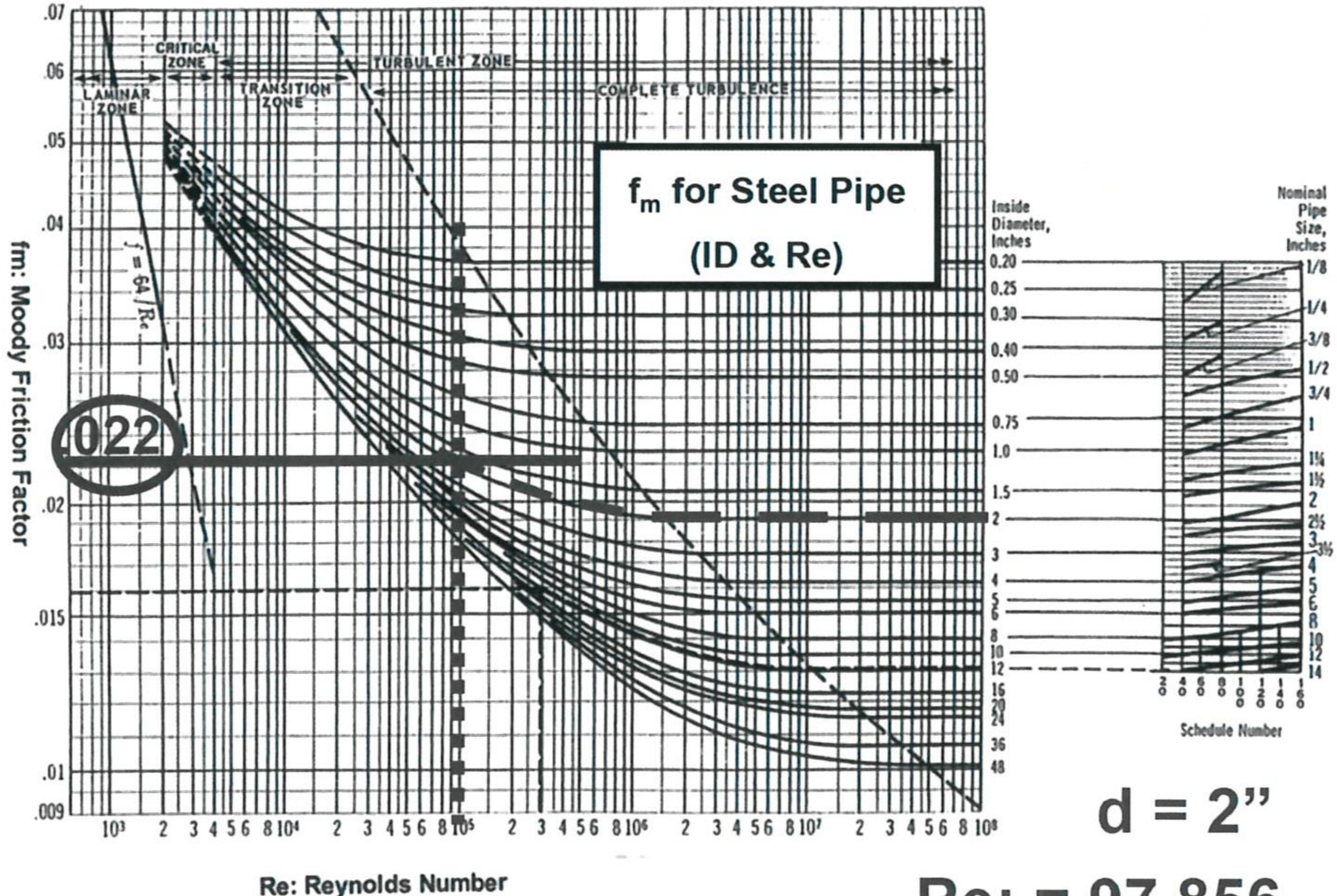
1. First calculate Reynold's number to determine flow regime

   :

2. Use the equation for

3. Determine f_m using chart

$\Delta P_{psi} =$



LIQUID: $Re = 92.1\ SG_L\ Q_{BLPD}\ /\ d\ \mu$

GAS: $Re = 20{,}100\ SG_g\ Q_{MMCFD}\ /\ d\ \mu$

d = 2"

Re: = 97,856

Pipeline sizing Summary

- **Consider Fluid Velocity**
  - Noise / Corrosion / Erosion
  - Liquid / Solids Build-Up
- **Contain Internal Pressure**

$$P = 2\,S\,t\,F\,E\,T\,/\,d$$

- **Pressure Drop: Horizontal Pipeline**

Laminar

$$\text{Liquid}: \Delta P_{psi} = 7.95 \times 10^{-6}\,\mu_{cp}\,L_{ft}\,Q_{BLPD}\,/\,d^4_{inch}$$

$$\text{Gas}: \Delta P_{psi} = 0.40\,\mu_g\,L_{ft}\,T\,Z\,Q_{MMCFD}\,/\,P_{psi}\,d^4_{inch}$$

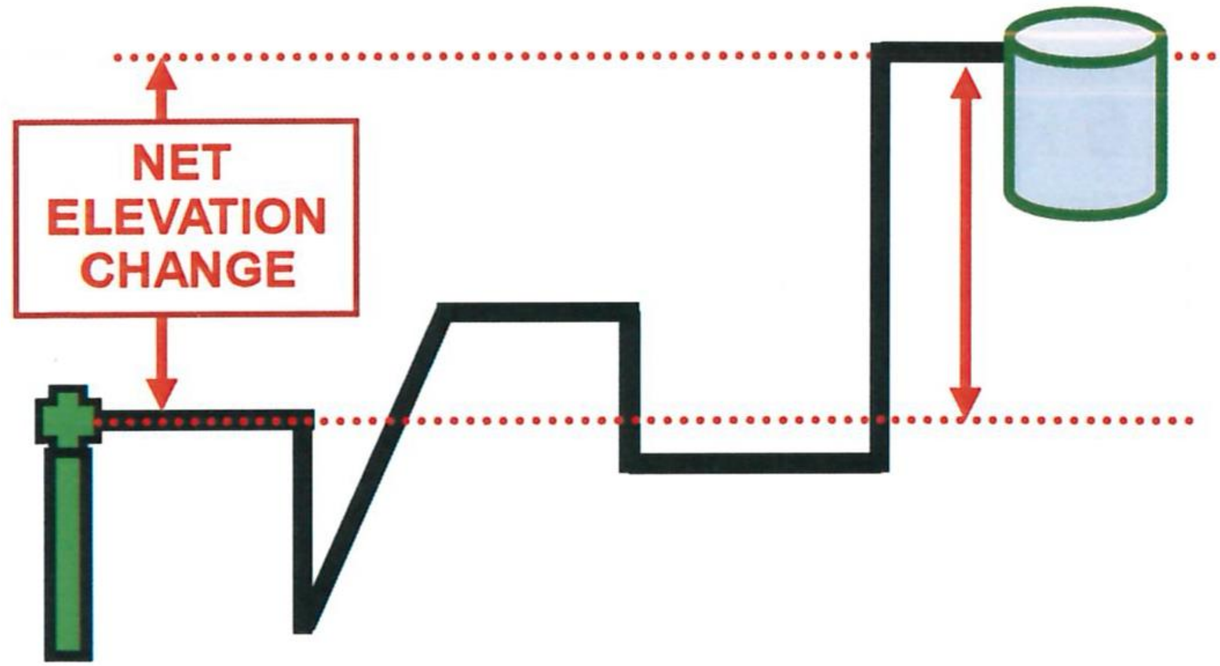
Non-Laminar

$$\text{Liquid}: \Delta P_{psi} = 11.5 \times 10^{-6}\,f_m\,L\,Q^2_{BLPD}\,SG_L\,/\,d^5$$

$$\text{Gas}: (P_1)^2 - (P_2)^2 = 25.1\,f_m\,L\,Q^2_{MMSCFD}\,SG_g\,Z\,T\,/\,d^5$$

Pipeline installation

What if pipeline is not horizontal?



Pressure drop due to Elevation

- Liquid: ΔP due to Elevation

$$\Delta P_{E(psi)} = \frac{\rho_{L(lb/ft^2)} H_{E(ft)}}{144} = 62.4\,\frac{SG_L H_{E(ft)}}{144}$$

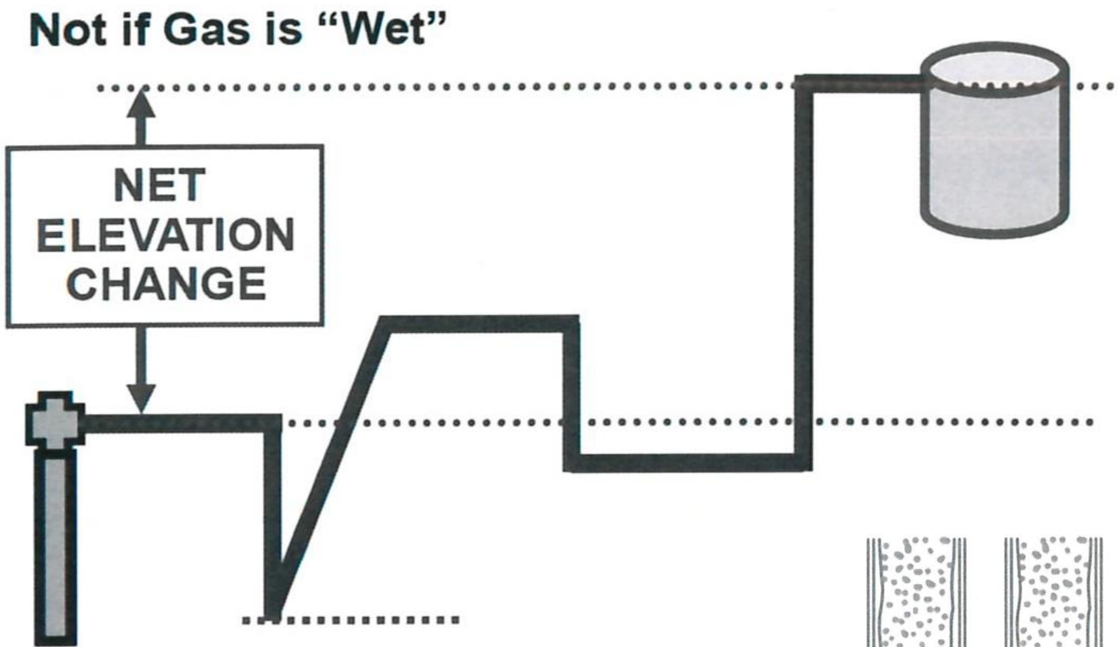
$$\Delta P_{E(psi)} = 0.433\ SG_L H_{E(ft)}$$

- Gas: ΔP due to Elevation

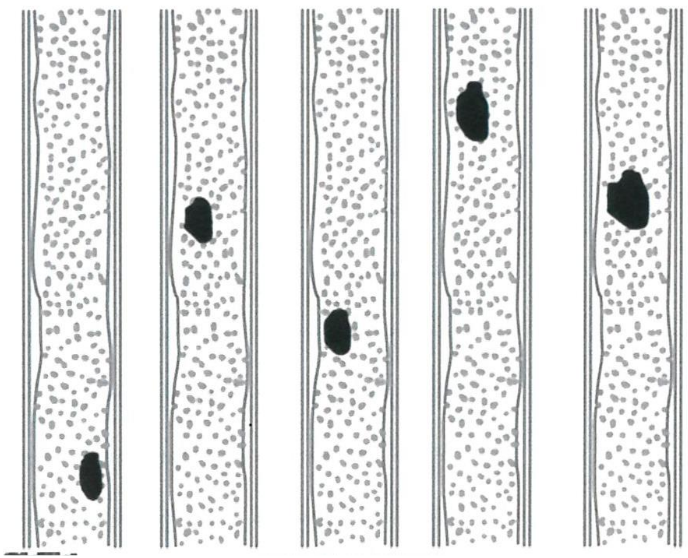
$$\Delta P_{E(psi)} = \frac{\rho_{G(lb/ft^2)} H_{E(ft)}}{144} = 2.70\ \frac{SG_G P_{psi}/T_{^oR} Z H_{E(ft)}}{144}$$

$$\Delta P_{E(psi)} = 0.188\ \frac{SG_G\ P_{psi}}{T_{^oR}\ Z\ H_{E(ft)}}$$

Not always true for gas flow

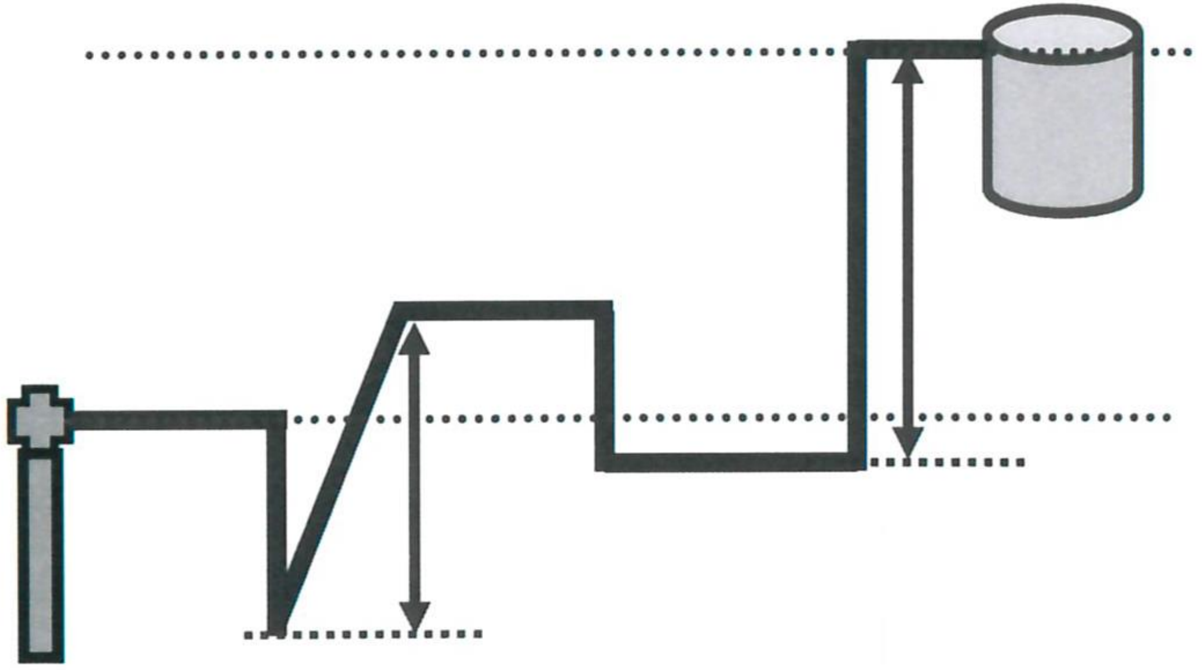
- 

- Big liquid droplets for annular flow



Pressure drop for Wet gas

- Sum the “Ups”



Estimating ΔP without using Friction Factor

- Empirical equations
  - Useful for quick calculation before use of PCs
  - Commonly accepted empirical equations

  : Hazen-Williams empirical equation (Liquid flow)

$$\Delta P \quad = 0.7 \times 10^{-6} \frac{Q^{1.85} L \; SG_L}{d^{4.87}}$$

$$(\Delta P \; in \; psi, Q \; in \; BLPD, L \; in \; feet, d = ID \; in \; inches)$$

  : Weymouth formula (gas flow)

$$P_2^2 = P_1^2 \quad - \left[\frac{0.8 \; L_{ft} T_R Z \; SG_G Q_{MMCFD}^2}{d_{in}^{5.334}} \right]$$

  - most common for oil field use
  - good for IDs between 0.75 inch & 16 inch
  - at Laminar rates, calculated ΔP is too low

  : Panhandle empirical equation (gas flow)

Panhandle: A & B Empirical equation

- For estimating ΔP without friction factor

A: $$Q_{MMCFD} = \left[\frac{0.020\,E\,(P_1^2 - P_2^2)^{0.51}\,d^{2.62}}{(SG_G^{0.853}\,zT^o{}_R L_{mi})^{0.539}}\right]$$

- For IDs between 6 inch and 24 inch
- Re between $5*10^6$ and $15*10^6$

B: $$Q_{MMCFD} = \left[\frac{0.028\,E\,(P_1^2 - P_2^2)^{0.51}\,d^{2.53}}{(SG_G^{0.961}\,zT^o{}_R L_{mi})^{0.51}}\right]$$

- For IDs between 6 inch and 24 inch
- Re > $15*10^6$

ΔP in psi, Q in MMCFD, L_{mi} in miles, d = ID in inches

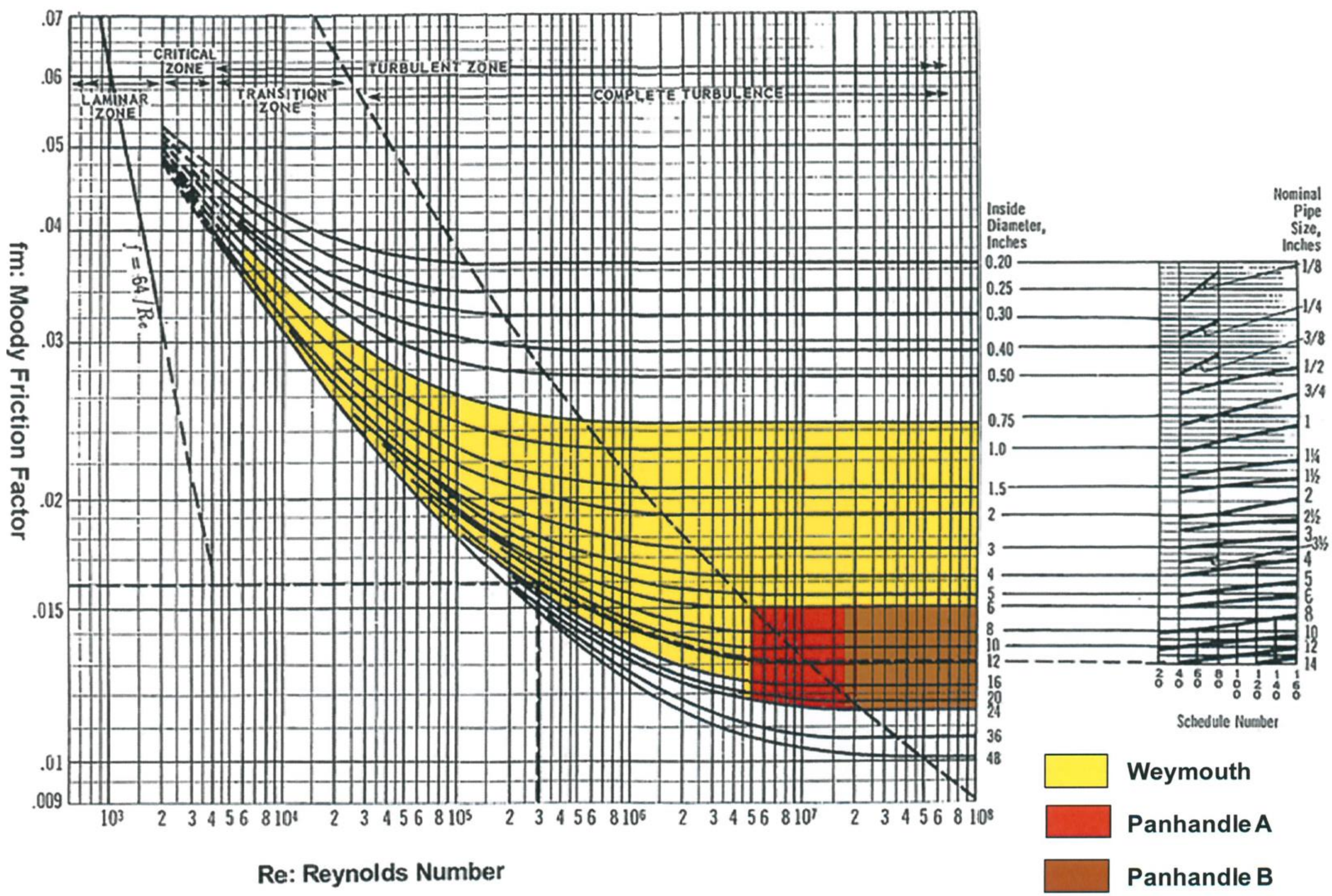
E factor : E = 1.00 for new pipe

= 0.95 for good condition

= 0.92 for average condition

= 0.85 for old pipe

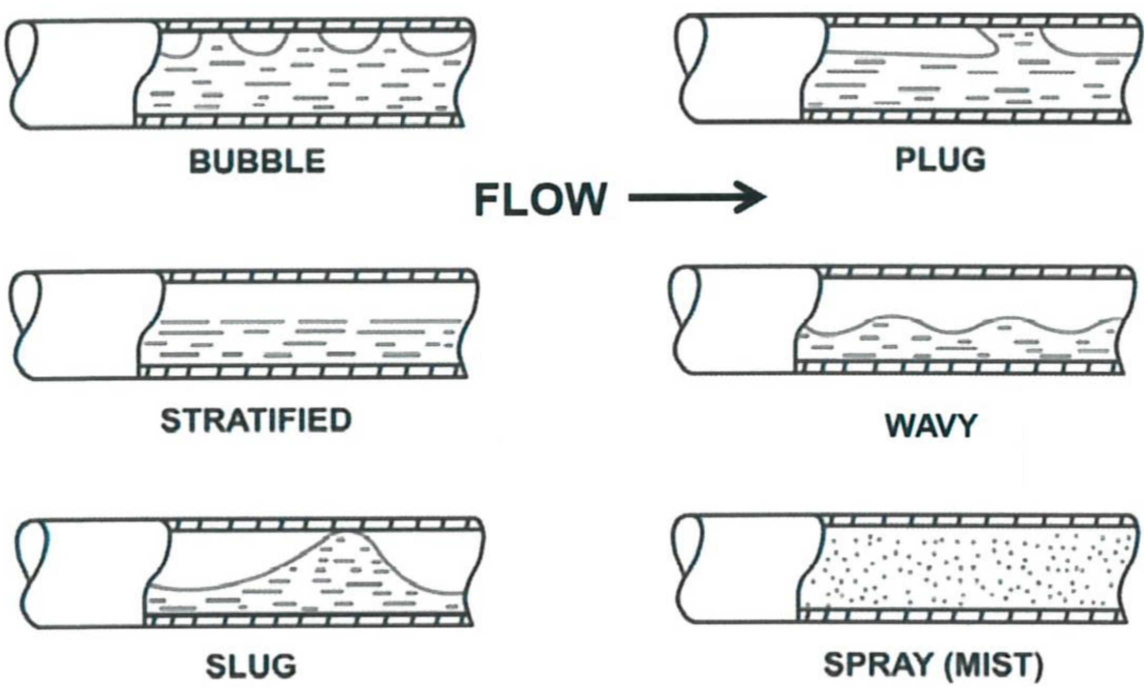
= 0.75 for corroded pipe



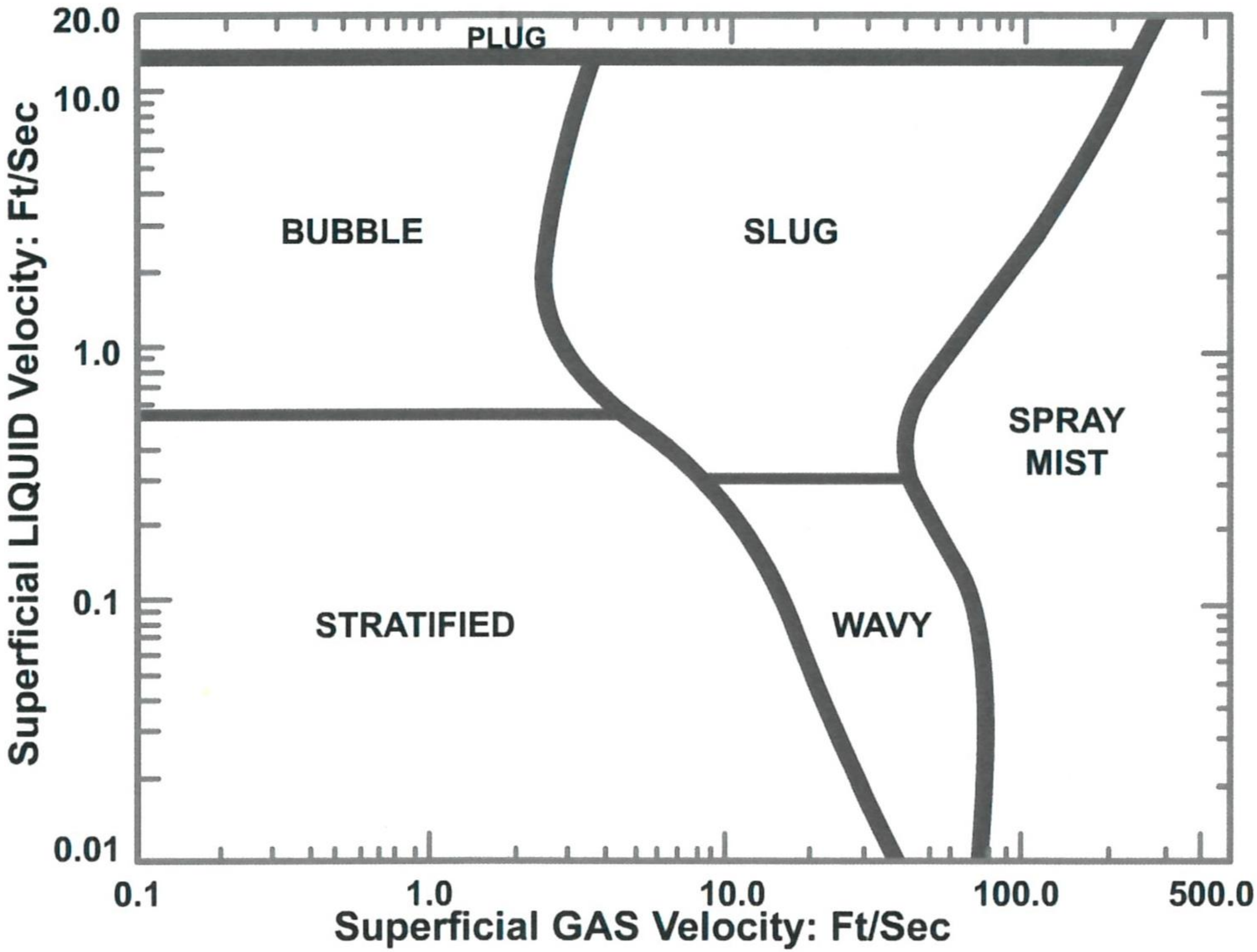
CRITICAL ZONE
TURBULENT ZONE
LAMINAR ZONE
TRANSITION ZONE
COMPLETE TURBULENCE
fm: Moody Friction Factor
Re: Reynolds Number
Inside Diameter, Inches
Nominal Pipe Size, Inches
Schedule Number
Weymouth
Panhandle A
Panhandle B

Pressure drop in pipe: Two phase flow

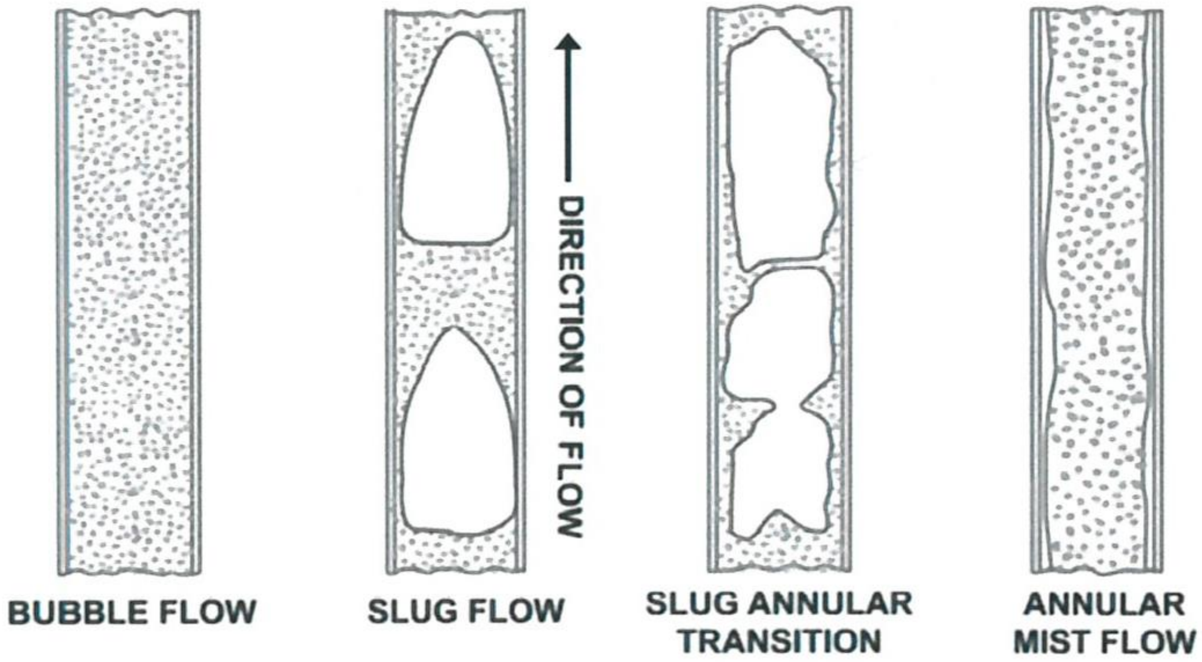
- With liquid and gas both flowing
  - Two phase flow
  - Three phase flow
- Horizontal flow patterns
  - Noise produced with bubbles
  - Using superficial velocities for gas and liquid



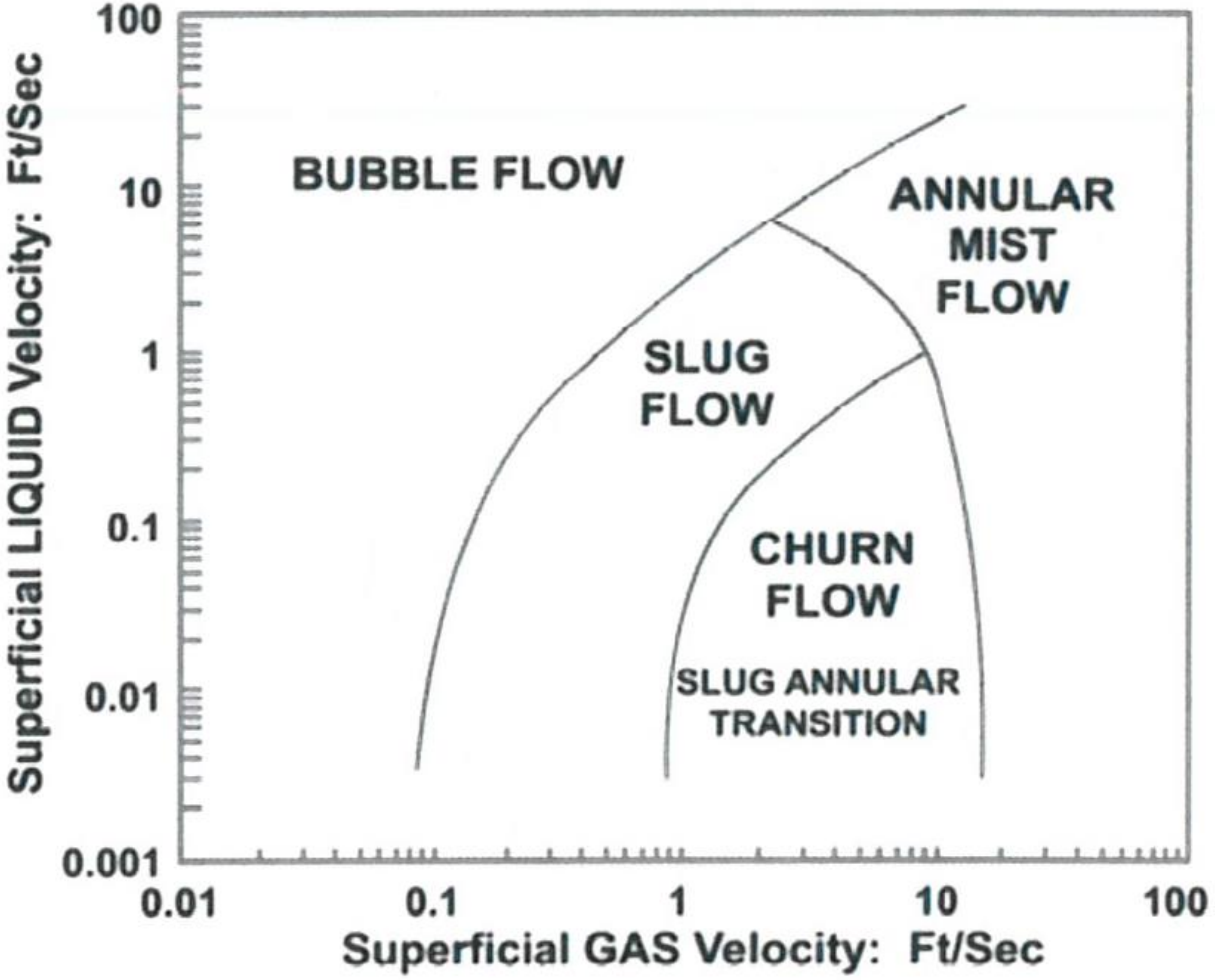
Two-phase horizontal flow regime



Vertical two-phase flow regimes



Two-phase vertical flow regimes



Pressure drop for two-phase flow

- Very complex: errors ≈ 20% common
  - Use simulation software and experience
- API RP 14 E gives following simplified method
  - Assumes: ΔP < 10%, bubble / mist flow, f=0.015

$$\Delta P = \frac{5 \times 10^{-8}\, L\, W^2}{d^5\ \ \rho_{mix}}$$

$$where,\ \text{W}= 3180\ Q_{MMCFD} SG_G + 14.6\ Q_{BPD} SG_L$$

$$and\ \rho_{mix} = \frac{12409\ SG_L P + 2.7\ R_{scf/bbl} SG_G P}{198.7\ P + R_{scf/bbl}\ T\ z}$$

Two phase flow: High GOR > 10,000 ft^3/bbl

- Use gas equations but change SG_G to :

$$SG_{mix} = \frac{SG_G + \frac{4591\, SG_L}{R_{scf/bbl}}}{1 + \frac{1123}{R_{scf/bbl}}}$$

If GOR < 10,000 scf/bbl, use two-phase correlations

AGA: Recommended multiphase ΔP calculations

HAND CALCULATION METHODS

- Frictional ΔP: Dukler, A.E., Moye Wicks, III, and R.G. Cleveland. "Frictional Pressure Drop in Two-Phase Flow: B. An Approach through Similarity Analysis" AIChE Journal, Vol 10, No. 1, January 1964, pp. 44-51.
- Elevation ΔP: Flanigan, Orin. "Effect of Uphill Flow on Pressure Drop in Design of Two-Phase Gathering Systems" Oil and Gas Journal, March 10, 1958, pp. 132-141.
- Liquid Hold-up: Eaton, Ben A., et al. "The Prediction of Flow Patterns, Liquid Holdup and Pressure Losses Occurring During Continues Two-Phase Flow in Horizontal Pipelines" J. Pet. Tech. AIME, JUNE 1967, pp.815-828.

For examples using these Methods : see AGA ENGINEERING DATA BOOK

Gas Processors Association. 1998

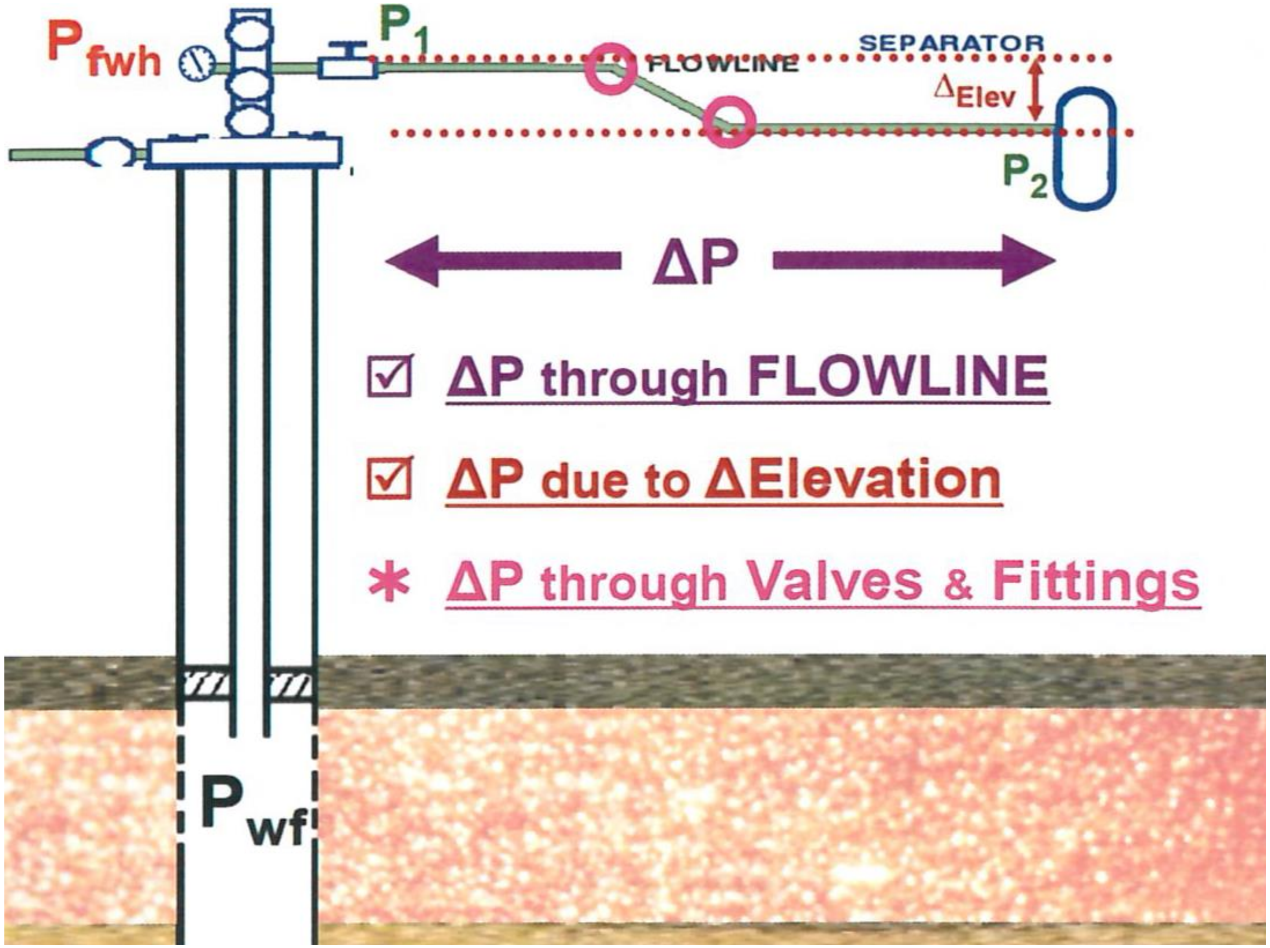
gpsa@gasprocessors.com

COMPUTER CALCULATION METHODS

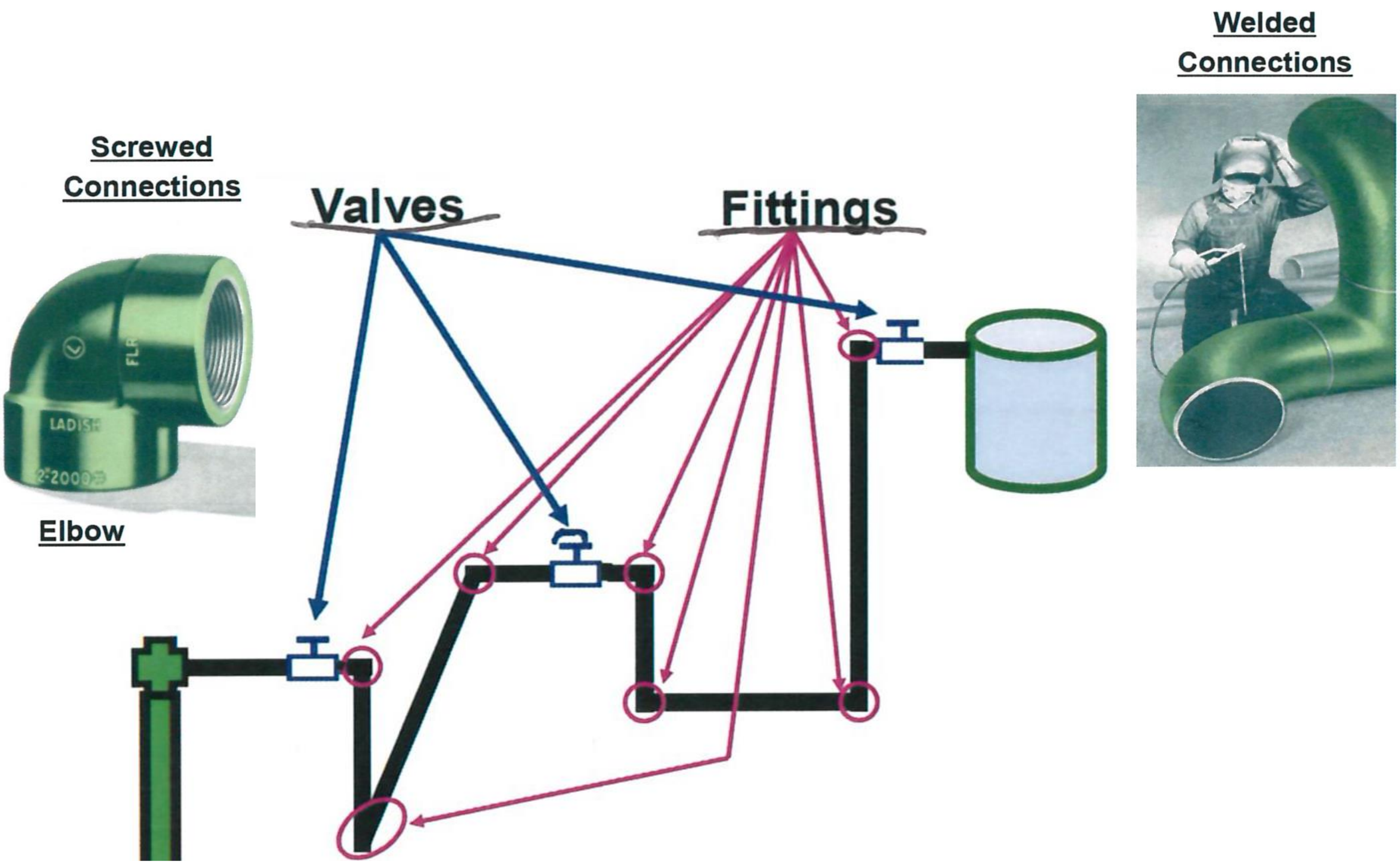
- Beggs, H. Dale, and James P. Brill. "A Study of Two-Phase Flow in Inclined Pipes" Trans. AIME, May 1973, pp. 606 – 617.
- Orkiszewski, J. "Predicting Two-Phase Pressure Drops in Vertical Pipe" Pet. Tech, AIME, 6/67 pp 829 – 838.
- Baker, O. , et al. "Gas-Liquid Flow in Pipelines, II. Design Manual" AGA - API Project NX-28, 10/ 70
- Brill & Mukherjee "Multiphase Flow in Wells" Monograph Vol 17 SPE Henry L Doherty Series

Ansari and Olga-S Transient Multiphase Simulator: OLGA

Pressure drop through valves and fittings



Pressure drop through valves and fittings



Pressure drop through valves and fittings

- Resistance coefficients: K_r
- Flow coefficients: liquid – C_v, Gas – C_g
- Equivalent length: L_E

Darcy's Law for valves and fittings

- Resistance coefficient: K_r

$$\Delta H = K_r \frac{v^2}{2\,g}, where\ K_r = \frac{f\,L}{D}$$

Larger K_r ⇒ Larger ΔP

- Liquid

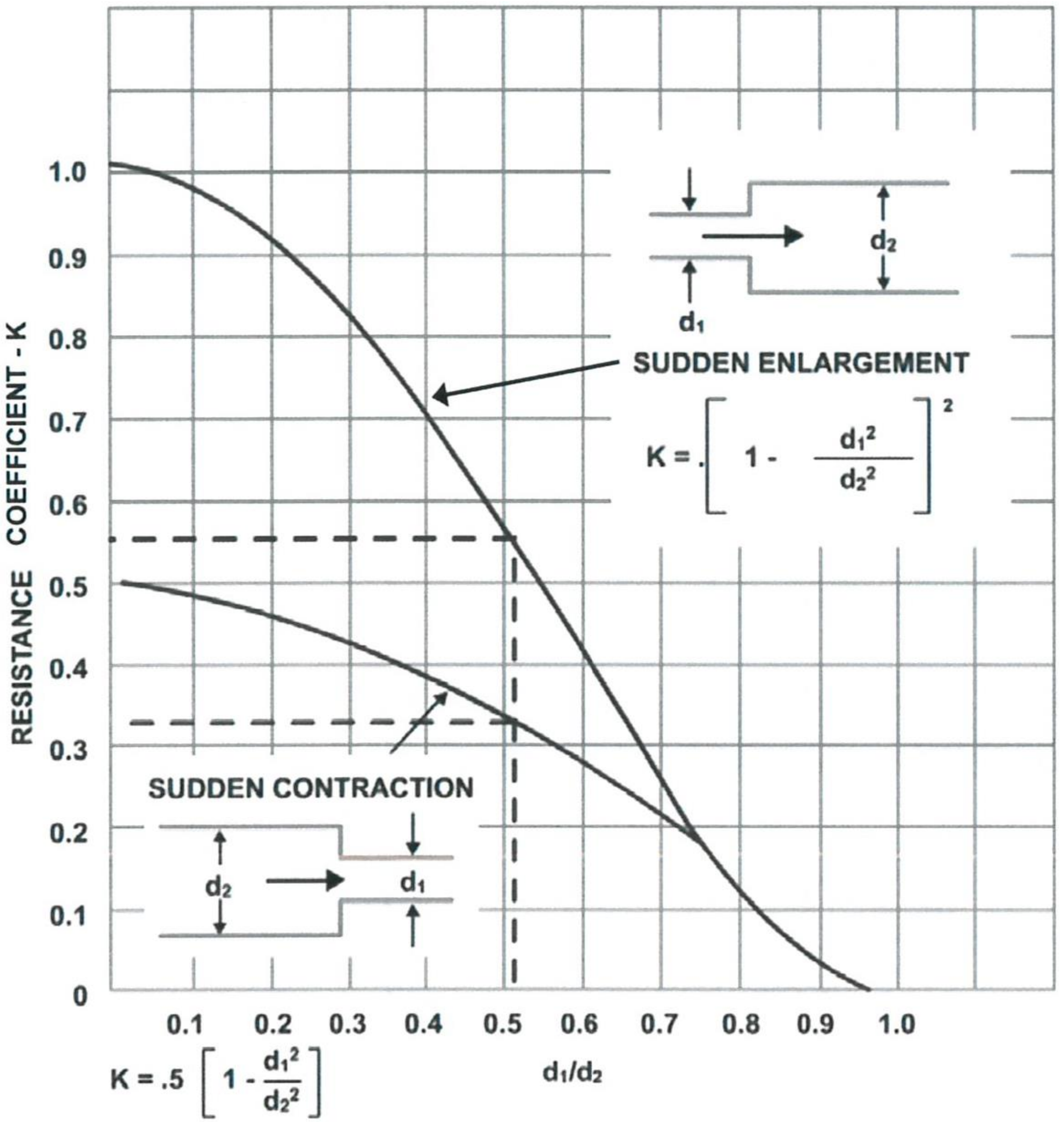
$$\Delta P_{psi} = 0.958 \times 10^{-6} \frac{K_r Q_{BPD}^2 SG_L}{d^4}$$

- Gas

$$P_1^2 - P_2^2 = 2.09 \frac{K_r Q_{MMCFD}^2 SG_G z\, T_{^oR}}{d^4}$$

Resistance coefficients

Global Valve, wide open	10.0
Angle Valve, wide open	5.0
Gate Valve, wide open	0.2
Tee	1.8
90° Elbow	0.9
45° Elbow	0.4



Darcy's law for valves and fittings

- Flow coefficient: C_v and C_g

Larger C_v (liquids) or C_g (gases) $\Rightarrow$ Smaller ΔP

- Liquid

$$\Delta P_{psi} = 8.5 \times 10^{-4} \frac{Q_{BPD}^2 SG_L}{C_v^2}$$

- Gas

$$P_1^2 - P_2^2 = 1.869 \frac{Q_{MMCFD}^2 SG_G z\, T_{°R}}{C_g^2}$$

Relationship between K_r and C_v

$$C_v = 29.9 \frac{d^2}{\sqrt{K_r}}$$

$$K_r = 894 \frac{d^4}{C_v^2}$$

Equivalent lengths

$$L_E = \frac{K_r \, d}{12 \, f_m}$$

$$L_E = 74.5 \, \frac{d^5}{f_m C_v^2}$$

L_E: Equivalent Length of Valves and Fittings, in Feet

Nominal Pipe size in.	Globe valve or ball check valve	Angle valve	Swing check valve	Plug cock	Gate or ball valve	45° ell Welded	45° ell Threaded	Short rad. ell Welded	Short rad. ell Threaded	Long rad. ell Welded	Long rad. ell Threaded	Hard T Welded	Hard T Threaded	Soft T Welded	Soft T Threaded	90° miter bends 2 miter	90° miter bends 3 miter	90° miter bends 4 miter	Enlargement Sudden d/D = 1/4	Enlargement Sudden d/D = 1/2	Enlargement Sudden d/D = 3/4	Enlargement Std. red. d/D = 1/2	Enlargement Std. red. d/D = 3/4	Contraction Sudden d/D = 1/4	Contraction Sudden d/D = 1/2	Contraction Sudden d/D = 3/4	Contraction Std. red. d/D = 1/2	Contraction Std. red. d/D = 3/4
1½	55	26	13	7	1	1	2	3	5	2	3	8	9	2	3				5	3	1	4	1	3	2	1	1	-
2	70	33	17	14	2	2	3	4	5	3	4	10	11	3	4				7	4	1	5	1	3	3	1	1	-
2½	80	40	20	11	2	2	-	5	-	3	-	12	-	3	-				8	5	2	6	2	4	3	2	2	-
3	100	50	25	17	2	2		6		4		14		4					10	6	2	8	2	5	4	2	2	-
4	130	65	32	30	3	3		7		5		19		5					12	8	3	10	3	6	5	3	3	-
6	200	100	48	70	4	4		11		8		28		8					18	12	4	14	4	9	7	4	4	1
8	260	125	64	120	6	6		15		9		37		9					25	16	5	19	5	12	9	5	5	2
10	330	160	80	170	7	7		18		12		47		12					31	20	7	24	7	15	12	6	6	2
12	400	190	95	170	9	9		22		14		55		14		28	21	20	37	24	8	28	8	18	14	7	7	2
14	450	210	105	80	10	10		26		16		62		16		32	24	22	42	25	9	-	-	20	16	8	-	-
16	500	240	120	145	11	11		29		18		72		18		38	27	24	47	30	10	-	-	24	18	9	-	-
18	550	280	140	160	12	12		33		20		82		20		42	30	28	53	35	11	-	-	26	20	10	-	-
20	650	300	155	210	14	14		35		23		90		23		46	33	32	60	38	13	-	-	30	23	11	-	-
22	688	335	170	225	15	15		40		25		100		25		52	36	34	65	42	14	-	-	32	25	12	-	-
24	750	370	185	254	16	16		44		27		110		27		56	39	36	70	46	15	-	-	35	27	13	-	-
30	-	-	-	312	21	21		55		40		140		40		70	51	44										
36	-	-	-		25	25		66		47		[illegible]		47		84	60	52										
42	-	-	-		30	30		77		55		200		55		98	69	64										
48	-	-	-		35	35		88		55		220		65		112	81	72										
54	-	[illegible]	[illegible]		40	40		99		70		250		70		126	90	80										
60	-	[illegible]	[illegible]		45	[illegible]		110		80		[illegible]		80		190	99	92										

Equiv. L in terms of small d

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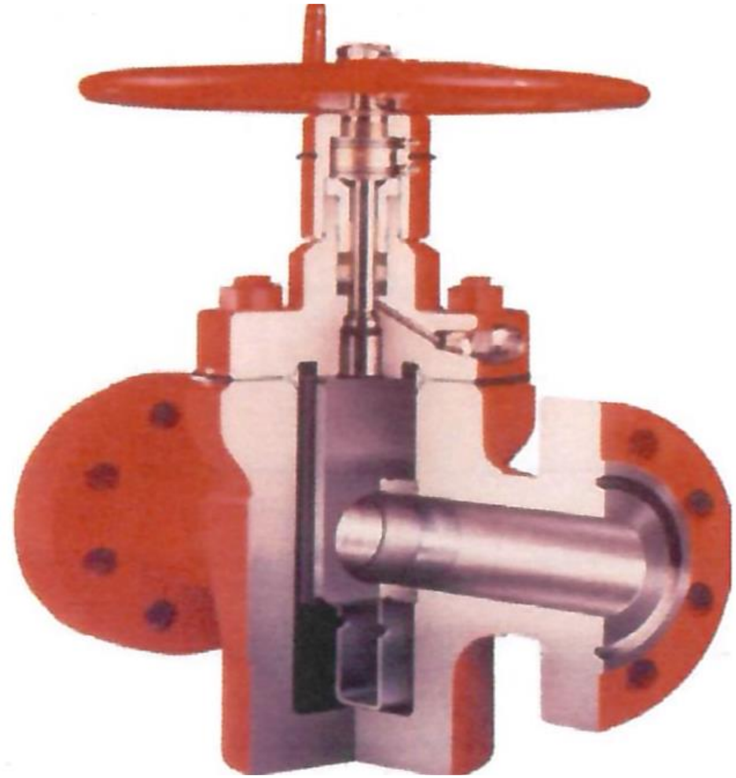
Full Opening - Small L_E

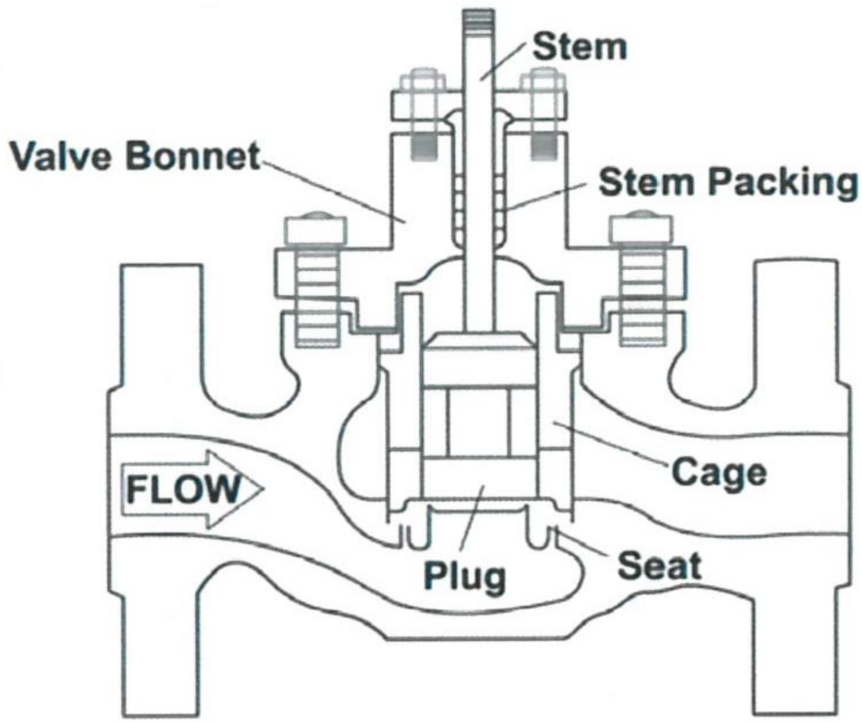
Reduced Opening - Large L_E

Gate valve



- **Full Opening Throughput**
- **Not Much ΔP**
- **Small L_E**

Globe valve



- Reduced Opening Throughput
- Larger ΔP
- Larger L_E

Thank you