THIRD EDITION

RULES OF THUMB FOR CHEMICAL ENGINEERS

A manual of quick, accurate solutions to everyday process engineering problems

CARL BRANAN

化学工程师用的经验法则

第3版

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process engineering problems

Third Edition

Carl R. Branan, Editor

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Equipment Design

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Velocity Head

Two of the most useful and basic equations are

$$\Delta h = \frac{u^2}{2g} \qquad (1)$$

$$\Delta P(V) + \frac{\Delta u^2}{2g} + \Delta Z + E = 0 \qquad (2)$$

where

Δh = Head loss in feet of flowing fluid
u = Velocity in ft/sec
g = 32.2 ft/sec^2
P = Pressure in lb/ft^2
V = Specific volume in ft^3/lb
Z = Elevation in feet
E = Head loss due to friction in feet of flowing fluid

In Equation 1 Δh is called the "velocity head." This expression has a wide range of utility not appreciated by many. It is used "as is" for

1. Sizing the holes in a sparger.
2. Calculating leakage through a small hole
3. Sizing a restriction orifice
4. Calculating the flow with a pitot tube

With a coefficient it is used for

1. Orifice calculations
2. Relating fitting losses, etc.

For a sparger consisting of a large pipe having small holes drilled along its length Equation 1 applies directly. This is because the hole diameter and the length of fluid travel passing through the hole are similar dimensions. An orifice on the other hand needs a coefficient in Equation 1 because hole diameter is a much larger dimension than length of travel (say ⅛ in. for many orifices). Orifices will be discussed under "Metering" in this chapter.

For compressible fluids one must be careful that when sonic or "choking" velocity is reached, further decreases in downstream pressure do not produce additional flow. This occurs at an upstream to downstream absolute pressure ratio of about 2 : 1. Critical flow due to sonic velocity has practically no application to liquids. The speed of sound in liquids is very high. See "Sonic Velocity" later in this chapter.

Still more mileage can be gotten out of $\Delta h = u^2/2g$ when using it with Equation 2, which is the famous Bernoulli equation. The terms are

1. The PV change
2. The kinetic energy change or "velocity head"
3. The elevation change
4. The friction loss

These contribute to the flowing head loss in a pipe. However, there are many situations where by chance, or on purpose, $u^2/2g$ head is converted to PV or vice versa.

We purposely change $u^2/2g$ to PV gradually in the following situations:

1. Entering phase separator drums to cut down turbulence and promote separation
2. Entering vacuum condensers to cut down pressure drop

We build up PV and convert it in a controlled manner to $u^2/2g$ in a form of tank blender. These examples are discussed under appropriate sections.

Source

Branan, C. R. *The Process Engineer's Pocket Handbook*, Vol. 1, Gulf Publishing Co., Houston, Texas, p. 1, 1976.

Piping Pressure Drop

A handy relationship for turbulent flow in commercial steel pipes is:

$$\Delta P_F = W^{1.8}\mu^{0.2}/20{,}000d^{4.8}\rho$$

where:

ΔP_F = Frictional pressure loss, psi/100 equivalent ft of pipe
W = Flow rate, lb/hr
μ = Viscosity, cp
ρ = Density, lb/ft^3
d = Internal pipe diameter, in.

This relationship holds for a Reynolds number range of 2,100 to 10^6. For smooth tubes (assumed for heat exchanger tubeside pressure drop calculations), a constant of 23,000 should be used instead of 20,000.

Source

Branan, Carl R. "Estimating Pressure Drop," *Chemical Engineering*, August 28, 1978.

Equivalent Length

The following table gives equivalent lengths of pipe for various fittings.

Table 1
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	Short rad. ell	Long rad. ell	Hard T.	Soft T.	90° miter bends			Enlargement					Contraction				
														Sudden			Std. red.		Sudden			Std. red.	
														Equiv. L in terms of small d									
						Weld thrd	Weld thrd	Weld thrd	Weld thrd	Weld thrd	2 miter	3 miter	4 miter	d/D = 1/4	d/D = 1/2	d/D = 3/4	d/D = 1/2	d/D = 3/4	d/D = 1/4	d/D = 1/2	d/D = 3/4	d/D = 1/2	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	26	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	36	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	170	47	84	60	52										
42	—	—	—		30	30	77	55	200	55	98	69	64										
48	—	—	—		35	35	88	65	220	65	112	81	72										
54	—	—	—		40	40	99	70	250	70	126	90	80										
60	—	—	—		45	45	110	80	260	80	190	99	92										

Sources

1. *GPSA Engineering Data Book*, Gas Processors Suppliers Association, 10th Ed. 1987.
2. Branan, C. R., *The Process Engineer's Pocket Handbook*, Vol. 1, Gulf Publishing Co., p. 6, 1976.

Recommended Velocities

Here are various recommended flows, velocities, and pressure drops for various piping services.

Sizing Steam Piping in New Plants Maximum Allowable Flow and Pressure Drop

	Laterals			Mains		
Pressure, PSIG	600	175	30	600	175	30
Density, #/CF	0.91	0.41	0.106	0.91	1.41	0.106
ΔP, PSI/100′	1.0	0.70	0.50	0.70	0.40	0.30
Nominal Pipe Size, In.	**Maximum Lb/Hr × 10 ³**					
3	7.5	3.6	1.2	6.2	2.7	0.9
4	15	7.5	3.2	12	5.7	2.5
6	40	21	8.5	33	16	6.6
8	76	42	18	63	32	14
10	130	76	32	108	58	25
12	190	115	50	158	87	39
14	260	155	70	217	117	54
16	360	220	100	300	166	78
18	...	300	130	...	227	101
20	...	...	170	...	...	132

Note:

(1) 600 PSIG steam is at 750°F., 175 PSIG and 30 PSIG are saturated.

(2) On 600 PSIG flow ratings, internal pipe sizes for larger nominal diameters were taken as follows: 18/16.5″, 14/12.8″, 12/11.6″, 10/9.75″.

(3) If other actual I.D. pipe sizes are used, or if local superheat exists on 175 PSIG or 30 PSIG systems, the allowable pressure drop shall be the governing design criterion.

Sizing Cooling Water Piping in New Plants Maximum Allowable Flow, Velocity and Pressure Drop

	LATERALS			MAINS		
Pipe Size in.	Flow GPM	Vel. ft/sec.	ΔP ft/100′	Flow GPM	Vel. ft/sec.	ΔP ft/100′
3	100	4.34	4.47	70	3.04	2.31
4	200	5.05	4.29	140	3.53	2.22
6	500	5.56	3.19	380	4.22	1.92
8	900	5.77	2.48	650	4.17	1.36
10	1,500	6.10	2.11	1,100	4.48	1.19
12	2,400	6.81	2.10	1,800	5.11	1.23
14	3,100	7.20	2.10	2,200	5.13	1.14
16	4,500	7.91	2.09	3,300	5.90	1.16
18	6,000	8.31	1.99	4,500	6.23	1.17
20	....	....	....	6,000	6.67	1.17
24	....	....	....	11,000	7.82	1.19
30	....	....	....	19,000	8.67	1.11

Sizing Piping for Miscellaneous Fluids

Dry Gas	100 ft/sec
Wet Gas	60 ft/sec
High Pressure Steam	150 ft/sec
Low Pressure Steam	100 ft/sec
Air	100 ft/sec
Vapor Lines General	Max. velocity 0.3 mach 0.5 psi/100 ft
Light Volatile Liquid Near Bubble Pt. Pump Suction	0.5 ft head total suction line
Pump Discharge, Tower Reflux	3–5 psi/100 ft
Hot Oil Headers	1.5 psi/100 ft
Vacuum Vapor Lines below 50 MM Absolute Pressure	Allow max. of 5% absolute pressure for friction loss

Suggested Fluid Velocities in Pipe and Tubing
(Liquids, Gases, and Vapors at Low Pressures to 50 psig and 50°F–100°F)

The velocities are suggestive only and are to be used to approximate line size as a starting point for pressure drop calculations. The final line size should be such as to give an economical balance between pressure drop and reasonable velocity.

Fluid	Suggested Trial Velocity	Pipe Material
Acetylene (Observe pressure limitations)	4000 fpm	Steel
Air, 0 to 30 psig	4000 fpm	Steel
Ammonia		
Liquid	6 fps	Steel
Gas	6000 fpm	Steel
Benzene	6 fps	Steel
Bromine		
Liquid	4 fps	Glass
Gas	2000 fpm	Glass
Calcium Chloride	4 fps	Steel
Carbon Tetrachloride	6 fps	Steel
Chlorine (Dry)		
Liquid	5 fps	Steel, Sch. 80
Gas	2000–5000 fpm	Steel, Sch. 80
Chloroform		
Liquid	6 fps	Copper & Steel
Gas	2000 fpm	Copper & Steel
Ethylene Gas	6000 fpm	Steel
Ethylene Dibromide	4 fps	Glass
Ethylene Dichloride	6 fps	Steel
Ethylene Glycol	6 fps	Steel
Hydrogen	4000 fpm	Steel
Hydrochloric Acid		
Liquid	5 fps	Rubber Lined
Gas	4000 fpm	R. L., Saran, Haveg
Methyl Chloride		
Liquid	6 fps	Steel
Gas	4000 fpm	Steel
Natural Gas	6000 fpm	Steel
Oils, lubricating	6 fps	Steel
Oxygen (ambient temp.)	1800 fpm Max.	Steel (300 psig Max.)
(Low temp.)	4000 fpm	Type 304 SS
Propylene Glycol	5 fps	Steel

Fluid	Suggested Trial Velocity	Pipe Material
Sodium Hydroxide		
0–30 Percent	6 fps	Steel
30–50 Percent	5 fps	and
50–73 Percent	4	Nickel
Sodium Chloride Sol'n.		
No Solids	5 fps	Steel
With Solids	(6 Min.– 15 Max.) 7.5 fps	Monel or nickel
Perchlorethylene	6 fps	Steel
Steam		
0–30 psi Saturated*	4000–6000 fpm	Steel
30–150 psi Saturated or superheated*	6000–10000 fpm	
150 psi up superheated	6500–15000 fpm	
*Short lines	15,000 fpm (max.)	
Sulfuric Acid		
88–93 Percent	4 fps	S. S.–316, Lead
93–100 Percent	4 fps	Cast Iron & Steel, Sch. 80
Sulfur Dioxide	4000 fpm	Steel
Styrene	6 fps	Steel
Trichlorethylene	6 fps	Steel
Vinyl Chloride	6 fps	Steel
Vinylidene Chloride	6 fps	Steel
Water		
Average service	3–8 (avg. 6) fps	Steel
Boiler feed	4–12 fps	Steel
Pump suction lines	1–5 fps	Steel
Maximum economical (usual)	7–10 fps	Steel
Sea and brackish water, lined pipe	5–8 fps } 3	R. L., concrete, asphalt-line, saran-lined, transite
Concrete	5–12 fps } (Min.)	

Note: R. L. = Rubber-lined steel.

Typical Design Vapor Velocities* (ft./sec.)

Fluid	≤6″	Line Sizes 8″–12″	≥14″
Saturated Vapor			
0 to 50 psig	30–115	50–125	60–145
Gas or Superheated Vapor			
0 to 10 psig	50–140	90–190	110–250
11 to 100 psig	40–115	75–165	95–225
101 to 900 psig	30–85	60–150	85–165

**Values listed are guides, and final line sizes and flow velocities must be determined by appropriate calculations to suit circumstances. Vacuum lines are not included in the table, but usually tolerate higher velocities. High vacuum conditions require careful pressure drop evaluation.*

Usual Allowable Velocities for Duct and Piping Systems*

Service/Application	Velocity, ft./min.
Forced draft ducts	2,500–3,500
Induced-draft flues and breeching	2,000–3,000
Chimneys and stacks	2,000
Water lines (max.)	600
High pressure steam lines	10,000
Low pressure steam lines	12,000–15,000
Vacuum steam lines	25,000
Compressed air lines	2,000
Refrigerant vapor lines	
High pressure	1,000–3,000
Low pressure	2,000–5,000
Refrigerant liquid	200
Brine lines	400
Ventilating ducts	1,200–3,000
Register grilles	500

**By permission,* Chemical Engineer's Handbook, *3rd Ed., p. 1642, McGraw-Hill Book Co., New York, N.Y.*

Typical Design* Velocities for Process System Applications

Service	Velocity, ft./sec.
Average liquid process	4–6.5
Pump suction (except boiling)	1–5
Pump suction (boiling)	0.5–3
Boiler feed water (disch., pressure)	4–8
Drain lines	1.5–4
Liquid to reboiler (no pump)	2–7
Vapor-liquid mixture out reboiler	15–30
Vapor to condenser	15–80
Gravity separator flows	0.5–1.5

**To be used as guide, pressure drop and system environment govern final selection of pipe size.*
For heavy and viscous fluids, velocities should be reduced to about ½ values shown.
Fluids not to contain suspended solid particles.

Suggested Steam Pipe Velocities in Pipe Connecting to Steam Turbines

Service—Steam	Typical range, ft./sec.
Inlet to turbine	100–150
Exhaust, non-condensing	175–200
Exhaust, condensing	400–500

Sources

1. Branan, C. R., *The Process Engineer's Pocket Handbook*, Vol. 1, Gulf Publishing Co., 1976.
2. Ludwig, E. E., *Applied Process Design for Chemical and Petrochemical Plants*, 2nd Ed., Gulf Publishing Co.
3. Perry, R. H., *Chemical Engineer's Handbook*, 3rd Ed., p. 1642, McGraw-Hill Book Co.

Two-phase Flow

Two-phase (liquid/vapor) flow is quite complicated and even the long-winded methods do not have high accuracy. You cannot even have complete certainty as to which flow regime exists for a given situation. Volume 2 of Ludwig's design books[1] and the GPSA Data Book[2] give methods for analyzing two-phase behavior.

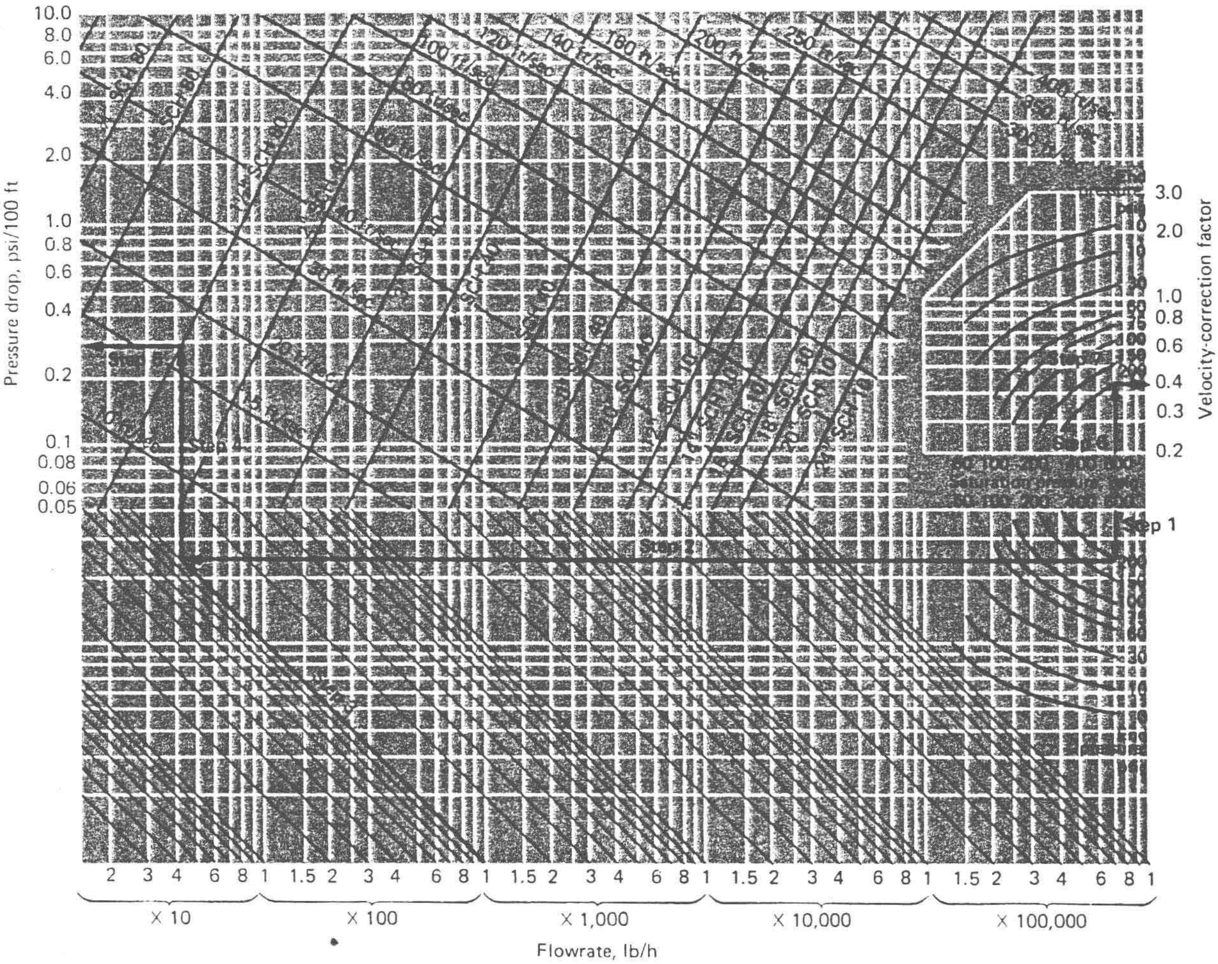
For our purposes, a rough estimate for general two-phase situations can be achieved with the Lockhart and Martinelli[3] correlation. Perry's[4] has a writeup on this correlation. To apply the method, each phase's pressure drop is calculated as though it alone was in the line. Then the following parameter is calculated:

$$X = [\Delta P_L / \Delta P_G]^{1/2}$$

where: ΔP_L and ΔP_G are the phase pressure drops

The X factor is then related to either Y_L or Y_G. Whichever one is chosen is multiplied by its companion pressure drop to obtain the total pressure drop. The following equation[5] is based on points taken from the Y_L and Y_G curves in Perry's[4] for both phases in turbulent flow (the most common case):

$$Y_L = 4.6X^{-1.78} + 12.5X^{-0.68} + 0.65$$
$$Y_G = X^2 Y_L$$



Sizing Lines for Flashing Steam-Condensate

The X range for Lockhart and Martinelli curves is 0.01 to 100.

For fog or spray type flow, Ludwig[1] cites Baker's[6] suggestion of multiplying Lockhart and Martinelli by two.

For the frequent case of flashing steam-condensate lines, Ruskan[7] supplies the handy graph shown above.

This chart provides a rapid estimate of the pressure drop of flashing condensate, along with the fluid velocities. *Example:* If 1,000 lb/hr of saturated 600-psig condensate is flashed to 200 psig, what size line will give a pressure drop of 1.0 psi/100 ft or less? Enter at 600 psig below insert on the right, and read down to a 200 psig end pressure. Read left to intersection with 1,000 lb/hr flowrate, then up vertically to select a 1½ in for a 0.28 psi/100 ft pressure drop. Note that the velocity given by this lines up if 16.5 ft/s are used; on the insert at the right read up from 600 psig to 200 psig to find the velocity correction factor 0.41, so that the corrected velocity is 6.8 ft/s.

Sources

1. Ludwig, E. E., *Applied Process Design For Chemical and Petrochemical Plants*, Vol. 1, Gulf Publishing Co. 2nd Edition., 1977.
2. *GPSA Data Book*, Vol. II, Gas Processors Suppliers Association, 10th Ed., 1987.
3. Lockhart, R. W., and Martinelli, R. C., "Proposed Correlation of Data for Isothermal Two-Phase, Two-Component Flow in Pipes," *Chemical Engineering Progress*, 45:39–48, 1949.

4. Perry, R. H., and Green, D., *Perry's Chemical Engineering Handbook*, 6th Ed., McGraw-Hill Book Co., 1984.
5. Branan, C. R., *The Process Engineer's Pocket Handbook*, Vol. 2, Gulf Publishing Co., 1983.
6. Baker, O., "Multiphase Flow in Pipe Lines," *Oil and Gas Journal*, November 10, 1958, p. 156.
7. Ruskan, R. P., "Sizing Lines For Flashing Steam-Condensate," *Chemical Engineering*, November 24, 1975, p. 88.

Compressible Flow

For "short" lines, such as in a plant, where $\Delta P > 10\%$ P_1, either break into sections where $\Delta P < 10\%$ P_1 or use

$$\Delta P = P_1 - P_2 = \frac{2P_1}{P_1 + P_2}\left[0.323\left(\frac{fL}{d} + \frac{\ln(P_1/P_2)}{24}\right)S_1 U_1^2\right]$$

from Maxwell[1] which assumes isothermal flow of ideal gas.

where:

ΔP = Line pressure drop, psi
P_1, P_2 = Upstream and downstream pressures in psi ABS
S_1 = Specific gravity of vapor relative to water = $0.00150 MP_1/T$
d = Pipe diameter in inches
U_1 = Upstream velocity, ft/sec
f = Friction factor (assume .005 for approximate work)
L = Length of pipe, feet
AP = Pressure drop in psi (rather than psi per standard length as before)
M = Mol. wt.

For "long" pipelines, use the following from McAllister[2]:

Equations Commonly Used for Calculating Hydraulic Data for Gas Pipe Lines

Panhandle A.

$$Q_b = 435.87 \times (T_b/P_b)^{1.0778} \times D^{2.6182} \times E \times \left[\frac{P_1^2 - P_2^2 - \dfrac{0.0375 \times G \times (h_2 - h_1) \times P_{avg}^2}{T_{avg} \times Z_{avg}}}{G^{0.8539} \times L \times T_{avg} \times Z_{avg}}\right]^{0.5394}$$

Panhandle B.

$$Q_b = 737 \times (T_b/P_b)^{1.020} \times D^{2.53} \times E \times \left[\frac{P_1^2 - P_2^2 - \dfrac{0.0375 \times G \times (h_2 - h_1) \times P_{avg}^2}{T_{avg} \times Z_{avg}}}{G^{0.961} \times L \times T_{avg} \times Z_{avg}}\right]^{0.51}$$

Weymouth.

$$Q = 433.5 \times (T_b/P_b) \times \left[\frac{P_1^2 - P_2^2}{GLTZ}\right]^{0.5} \times D^{2.667} \times E$$

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

P_{avg} is used to calculate gas compressibility factor Z

Nomenclature for Panhandle Equations

Q_b = flow rate, SCFD
P_b = base pressure, psia
T_b = base temperature, °R
T_{avg} = average gas temperature, °R
P_1 = inlet pressure, psia
P_2 = outlet pressure, psia
G = gas specific gravity (air = 1.0)
L = line length, miles
Z = average gas compressibility
D = pipe inside diameter, in.
h_2 = elevation at terminus of line, ft
h_1 = elevation at origin of line, ft
P_{avg} = average line pressure, psia
E = efficiency factor
E = 1 for new pipe with no bends, fittings, or pipe diameter changes

E = 0.95 for very good operating conditions, typically through first 12–18 months
E = 0.92 for average operating conditions
E = 0.85 for unfavorable operating conditions

Nomenclature for Weymouth Equation

Q = flow rate, MCFD
T_b = base temperature, °R
P_b = base pressure, psia
G = gas specific gravity (air = 1)
L = line length, miles
T = gas temperature, °R
Z = gas compressibility factor
D = pipe inside diameter, in.
E = efficiency factor. (See Panhandle nomenclature for suggested efficiency factors)

Sample Calculations

Q = ?
G = 0.6
T = 100°F
L = 20 miles
P_1 = 2,000 psia
P_2 = 1,500 psia
Elev diff. = 100 ft
D = 4.026-in.
T_b = 60°F
P_b = 14.7 psia
E = 1.0
P_{avg} = 2/3(2,000 + 1,500 − (2,000 × 1,500/2,000 + 1,500))
= 1,762 psia

Z at 1,762 psia and 100°F = 0.835.

Panhandle A.

$$Q_b = 435.87 \times (520/14.7)^{1.0788} \times (4.026)^{2.6182} \times 1 \times \left[\frac{(2,000)^2 - (1,500)^2 - \dfrac{0.0375 \times 0.6 \times 100 \times (1,762)^2}{560 \times 0.835}}{(0.6)^{.8539} \times 20 \times 560 \times .835}\right]^{0.5394}$$

Q_b = 16,577 MCFD

Panhandle B.

$$Q_b = 737 \times (520/14.7)^{1.020} \times (4.026)^{2.53} \times 1 \times \left[\frac{(2,000)^2 - (1,500)^2 - \dfrac{0.0375 \times 0.6 \times 100 \times (1,762)^2}{560 \times 0.835}}{(0.6)^{.961} \times 20 \times 560 \times .835}\right]^{0.51}$$

Q_b = 17,498 MCFD

Weymouth.

$$Q = 0.433 \times (520/14.7) \times \left[(2,000)^2 - (1,500)^2 / (0.6 \times 20 \times 560 \times 0.835)\right]^{1/2} \times (4.026)^{2.667}$$

Q = 11,101 MCFD

Source

Pipecalc 2.0, Gulf Publishing Company, Houston, Texas. Note: Pipecalc 2.0 will calculate the compressibility factor, minimum pipe ID, upstream pressure, downstream pressure, and flow rate for Panhandle A, Panhandle B, Weymouth, AGA, and Colebrook-White equations. The flow rates calculated in the above sample calculations will differ slightly from those calculated with Pipecalc 2.0 since the viscosity used in the examples was extracted from Figure 5, p. 147. Pipecalc uses the Dranchuk et al. method for calculating gas compressibility.

Equivalent Lengths for Multiple Lines Based on Panhandle A

Condition I.

A single pipe line which consists of two or more different diameter lines.

Let L_E = equivalent length
$L_1, L_2, \ldots L_n$ = length of each diameter
$D_1, D_2, \ldots D_n$ = internal diameter of each separate line corresponding to $L_1, L_2, \ldots L_n$
D_E = equivalent internal diameter