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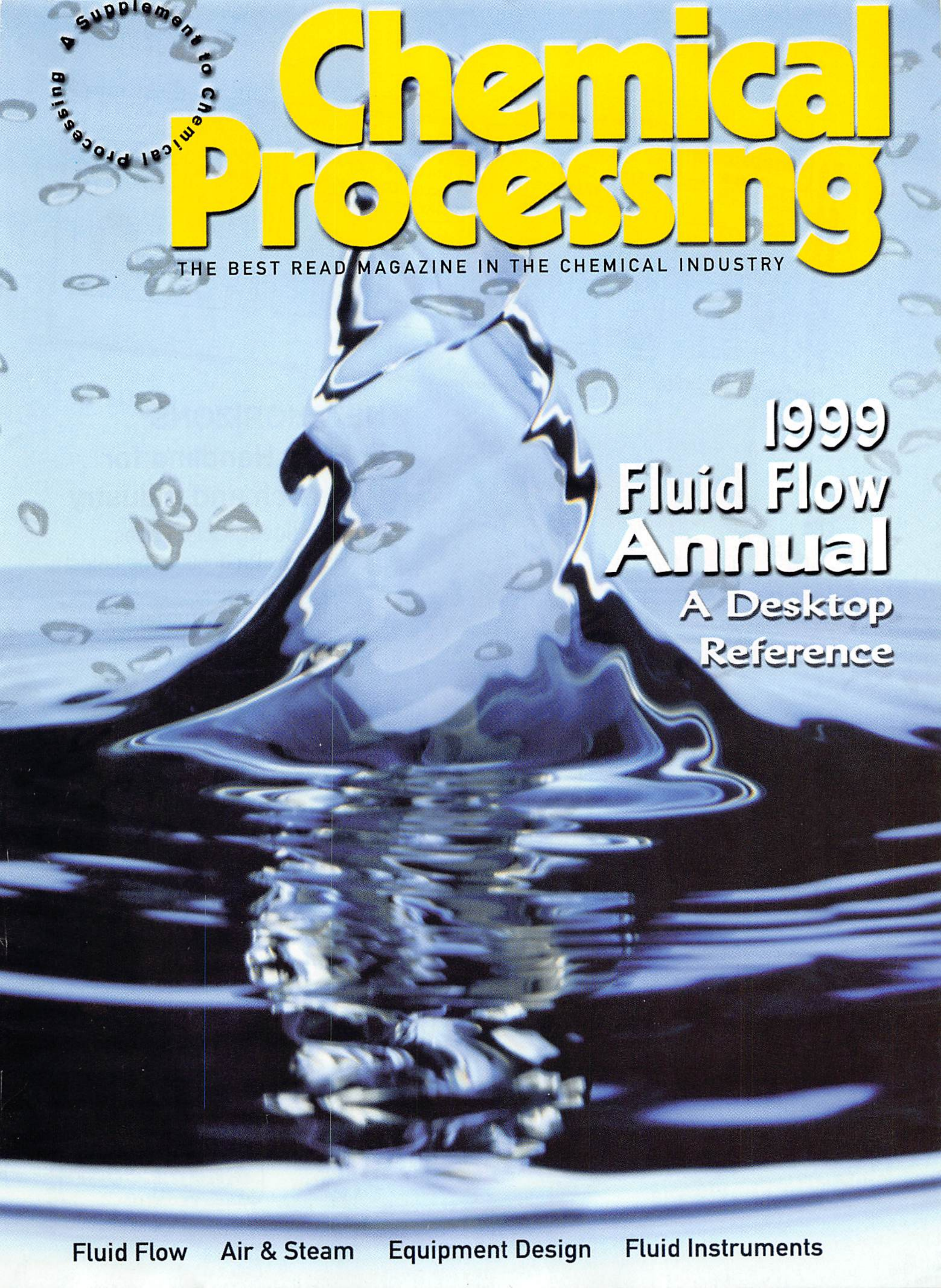

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Control valve sizing

This guide briefly discusses several subjects relevant to sizing and selecting the right control valve for a particular application, followed by simplified step-by-step procedures for performing sizing calculations by hand along with worked examples.

Selection of control valve style

The choice of control valve style (globe, ball, butterfly, etc.) is often based on tradition or plant preference. For example, a majority of the control valves in a pulp and paper mills are usually ball or segmented ball valves. Refineries traditionally use a high percentage of globe valves for control.

Globe valves offer the widest range of options for flow characteristics, pressure, temperature and noise and cavitation reduction. Globe valves also tend to be the most expensive.

Segment ball valves tend to have a higher rangeability, and size for size, nearly twice the flow capacity of globe valves and in addition are less expensive than globe valves. On the other hand, segment ball valves are limited in availability for extremes of temperature and pressure and are more prone to noise and cavitation than are globe valves.

High-performance butterfly valves are even less expensive than ball valves, especially in larger sizes (say 8-in and larger). They also have less rangeability than the ball valves and are more prone to cavitation. The eccentric rotary plug valves combine features of rotary valves, such as high cycle life stem seals and compact construction with the rugged construction of glove valves. Unlike the other rotary valves whose flow capacity is approximately double that of globe valves, the flow capacity of eccentric rotary plug valves is on a par with globe valves. Table 1 is generalized comparison of the various styles of control valves.

Although an extensive discussion of the selection of the proper valve flow characteristics for a particular application is beyond the scope of this article, as a general rule, systems with a significant amount of pipe and fittings, and/or with centrifugal pumps (the most common case) are best suited to equal percentage valves. Systems with very little pipe tend to be better suited to linear characteristics valves. For the final design of more critical systems the valve manufacturer should calculate the installed characteristic of the selected valve in a particular system to ensure a good match between system and valve.

Process data

A valve sizing calculation will only be reliable if the process data used in the calculation accurately represents the true process. Perhaps the most misunderstood area of control valve sizing is the selection of the pressure drop, Δp, to use in the sizing calculation. The Δp cannot be arbitrarily specified without regard for the actual system into which the valve will be installed. The correct procedure for determining the pressure drop across a control valve is to start upstream of the valve at a point where the pressure is known (for example, a pump where the pressure can be determined from the head curve) and subtract the pressure losses due to the upstream pipe and fittings. When the inlet to the valve p_1 is

NOMENCLATURE

C_v	Valve flow coefficient;
d	Valve inlet diameter, in;
Δp	Pressure differential across a control valve, $p_1 - p_2$, psi, bar;
Δp_D	Pressure differential beyond which cavitation damage is likely, psi, bar;
Δp_T	Pressure differential at which liquid flow becomes choked, psi, bar;
F_F	Liquid critical pressure ratio factor, dimensionless;
F_k	Ratio of specific heats factor, dimensionless;
F_L	Liquid pressure recovery factor, dimensionless;
γ_1	Specific weight of gas or vapor at upstream conditions, lb/cu ft, kg/cu m;
G_f	Liquid-specific gravity at upstream conditions (ratio of density of liquid at flowing temperature to density of water at 60°F), dimensionless;
G_g	Gas-specific gravity (ratio of density of flowing gas to density of air with both at standard conditions, which is equal to the ratio of the molecular weight of gas to the molecular weight of air), dimensionless;
k	Ratio of specific heats, dimensionless;
M	Molecular weight, atomic mass units;
p_1	Upstream absolute static pressure, psia, bara;
p_2	Downstream absolute static pressure, $p_1 - \Delta p$, psia, bara;
p_c	Absolute thermodynamic critical pressure, psia, bara;
PDC	Pipe diameter correction for aerodynamic noise calculation, dBA;
PSC	Pipe schedule correction for aerodynamic noise calculation, dBA;
p_v	Absolute vapor pressure of liquid at inlet temperature, psia, bara;
q	Volumetric flowrate of gas, scfh, N cu m/hr; or volumetric flowrate of liquid, gpm, cu m/hr
SPL	Sound pressure level, dBA;
T_1	Absolute upstream temperature, °R, °K;
VSC	Valve style correction for aerodynamic noise calculation, dBA;
w	Weight or mass flowrate, lb/hr, kg/hr;
x	Ratio of pressure drop to absolute inlet pressure, dimensionless;
x_T	Terminal or limiting pressure drop ratio for gas, dimensionless;
Y	Expansion factor, dimensionless;
Z	Compressibility factor, dimensionless.

known, the next step is to go to a point downstream of the control valve where the pressure is known (for example, a tank where the head is known) and then work upstream toward the control valve, adding the pressure losses of the pipe and fittings. (The pressure losses are added by working in the direction opposite to the flow.) At the valve outlet p_2 is known. The actual pressure drop across the control valve is the different between the upstream and downstream pressure, i.e., $\Delta p = p_1 - p_2$. To perform sizing calculations at more than one flowrate (e.g., at both maximum and minimum design flows), the calculation of p_1 and p_2 at each flowrate must be repeated, as the system pressure losses (and pump head) are dependent on the flow.

Liquid choked flow, cavitation, flashing

When liquid flow in a control valve passes through the *vena contracta* (the point at which the cross-sectional area of the flow stream is at a minimum), the flow velocity reaches a maximum and the pressure decreases to a minimum.

The static pressure at the *vena contracta* is a function of three things—the pressure immediately upstream of the valve (p_1); the pressure drop (Δp) across the valve; and the valve geometry expressed in the manufacturer's literature as the liquid pressure recovery factor, F_L (see Fig. 1 for typical values of F_L).

For a fixed value of p_1, as the pressure drop across a control valve increases, the pressure at the *vena contracta* decreases. If the pressure drop across the control valve increases to a point where the *vena contracta* pressure decreases to slightly below the vapor pressure, p_v, of the liquid, vapor bubbles form in the *vena contracta*. Once this happens, additional increases in pressure drop across the valve do not result in additional flow, and flow is choked. This limiting or choking pressure drop is called the terminal pressure drop, Δp_T. (The same thing is also occasionally referred to as Δp_{MAX} or $\Delta p_{allowable}$.) The calculation of Δp_T (see step 2 of the sizing method for liquids) is important, because when the actual pressure drop, Δp, is greater than Δp_T, then Δp_T and not Δp must be used in the sizing equations to prevent undersizing the valve.

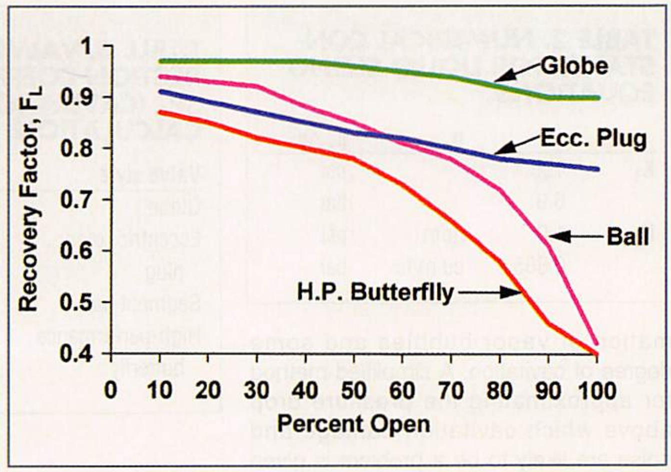

Fig. 1. Liquid pressure recovery factor (FL) of control valves.

Choked flow produces either flashing or cavitation. Flashing results if the pressure downstream of the valve, p_2, is less than the vapor pressure of the liquid. In this case, the vapor bubbles that formed at the *vena contracta* continue downstream. Flashing conditions have the potential for erosive damage to the valve by drops of liquid entrained in high velocity vapor and selection of erosion-resistant materials (such as stainless-steel bodies and hardened trim) is advisable.

Cavitation results from choked flow when p_2 is greater than p_v. In this case, as the vapor bubbles travel downstream from the *vena contracta* they collapse violently, resulting in vibration, noise and damage. The use of hard or erosion-resistant materials is not very effective in preventing cavitation damage and, as a rule, cavitation should be avoided.

Cavitation can usually be eliminated by selecting a valve style with a higher value of F_L. In general, as the F_L increases, so does the price of the valve. In addition to the standard valve styles shown in Fig. 1, there are special cavitation-resistant valves available.

In practice, at pressure drops approaching, but below the calculated value of Δp_T, there is usually some for-

TABLE 1. COMPARISON OF CONTROL VALVE TYPES.

	Top-guided globe	Cage-guided globe	Segment ball	Eccentric rotary plug	High-performance butterfly
Cost	High	High	Medium	Medium	Low
Weight	High	High	Medium	Medium	Low
Flow capacity (compared to globe)	1X	1X	2X	1X	2X
Cavitation potential	Low	Low	Medium	Medium	High
In-line repairable	Yes	Yes	No	No	No
Inherent flow characteristic	= %, linear, quick opening	= %, linear, quick opening	= %	Modified linear	Modified = %
Cavitation/noise reduction options	No	Yes	Some	Some	No
Suitable for high pressure differential	Limited	Yes	Limited	Yes	Limited
Suitable for dirty service	Yes	No	Yes	Yes	Yes
Suitable for slurries	Limited	No	Yes	Yes	Limited
Suitable for pulp stock	No	No	Yes	No	Limited

TABLE 2. NUMERICAL CONSTANTS FOR LIQUID SIZING EQUATIONS.

	q	p, Δp
K_1	100	psi
	6.9	bar
K_2	1.0	gpm	psi
	0.865	cu m/hr	bar

TABLE 3. VALVE STYLE CORRECTION COEFFICIENTS FOR ΔP_D (CAVITATION DAMAGE) CALCULATION.

Valve style	R	S
Globe	1.0	0.5
Eccentric rotary plug	1.0	0.35
Segment ball	0.7	0.2
High-performance butterfly	0.6	0.16

mation of vapor bubbles and some degree of cavitation. A simplified method for approximating the pressure drop above which cavitation damage and noise are likely to be a problem is given in Equation 3.

$$\Delta p_D = RF_L^2\left(\frac{K_1}{p_1}\right)^S (p_1 - p_v) \quad (3)$$

Sizing method for liquids

The following seven steps are used for control valve sizing for liquids (see nomenclature sidebar). Constants K_1 and K_2 (Table 2) are included to make the formulas readily adaptable to customary U.S. or S.I. units. The values of p_1, p_2 and p_v must always be expressed in absolute units.

1. Calculate the critical pressure ratio factor, F_F:

$$F_F = 0.96 - 0.28\sqrt{\frac{p_v}{p_c}} \quad (1)$$

2. Calculate the terminal pressure drop, Δp_T:

$$\Delta p_T = F_L^2(p_1 - F_F p_v) \quad (2)$$

F_L is a function of valve opening that is yet to be determined, so it will be necessary to make an initial estimate of valve opening. A good initial estimate is the value of F_L at 80% open. For manufacturers using factors other than F_L, the following is used to convert from one factor to another: $C_f = K_m^{1/2} = F_L$.

3. If $\Delta p > \Delta p_T$, flow will be choked. Use Δp_T in place of Δp in Step 7. Go to Step 4. If $\Delta p < \Delta p_T$, skip Step 4 and go to Step 5.

4. If $p_2 < p_v$, flow will be flashing, erosion-resistant valve materials are required; go to Step 7. Otherwise continue with Step 5.

5. Calculate the pressure drop at which cavitation damage is likely to begin (Δp_D): The values of R and S for the various valve styles are found in Table 3.

6. If $\Delta p > \Delta p_D$, there is the potential for cavitation damage. A valve with a higher F_L should be used. If $\Delta p < \Delta p_D$, there is limited danger of cavitation and excessive noise.

7. Calculate the required C_v:

$$C_v = K_2 q \sqrt{\frac{G_f}{\Delta p}} \quad (4)$$

Using the calculated C_v, an appropriate valve size is chosen from the manufacturers' tables of C_v vs. valve opening as represented in Tables 4 and 5. The goal is to select a valve that will be as far open as possible without exceeding 80% open at the maximum design flow.

For valves installed with reducers, selecting a valve with an opening not exceeding 75% according to the tables will ensure that, with the reducer effect, it will not be in excess of 80% open in actual operation.

TABLE 4. TYPICAL VALVE FLOW COEFFICIENTS, C_v, FOR SEGMENT BALL CONTROL VALVES.

	Relative opening (%)				
Size	20	40	60	75	100
1	1	4	9	14	45
1.5	3	11	25	40	110
2	5	17	38	62	180
3	12	39	88	143	420
4	19	63	141	228	620
6	34	115	258	418	1,260
8	55	187	418	677	2,030
10	87	294	658	1,068	3,210
12	123	418	937	1,516	4,490

TABLE 5. TYPICAL VALVE FLOW COEFFICIENTS, C_v, FOR EQUAL PERCENTAGE GLOBE CONTROL VALVES.

	Relative opening (%)				
Size	20	40	60	80	100
1	0.1	0.6	2	6	13
1.5	1.1	3.6	9.2	23	37
2	2	5	17	45	60
3	5	12	38	91	128
4	6	13	40	106	170
6	9	23	89	253	362
8	14	47	142	420	650
10	29	62	170	566	950

TABLE 6. RATIO OF SPECIFIC HEATS FACTORS, F_K.

Gas	F_K
Air	1.0
Ammonia	0.92
Butane	0.96
Carbon dioxide	0.93
Ethane	0.85
Freon	0.96
Hydrogen	1.0
Methane	0.94
Natural gas	0.94
Nitrogen	1.0
Oxygen	1.0
Steam	0.93

After selecting a valve size and determining the percent of opening corresponding to the calculated C_v, the F_L at that opening is determined.

If the actual F_L is less than the value used in Steps 1 and 5, all the steps are repeated because the calculated values of Δp_T and Δp_D will be larger than their real values and there is the possibility of overlooking choked flow and potential cavitation problems.

If the actual F_L is greater than the value used in the calculation, the calculation is conservative.

To check for potential cavitation problems, the C_v at the minimum design flow may be calculated. Making the initial estimate of valve opening for the purpose of obtaining the initial value of F_L may require some trial and error. Estimating an opening of 40% is a good starting point, but p_1 and Δp will most likely be higher at the minimum design rate than they are at the maximum design flowrate.

Liquid sizing example

The stated problem is to select a properly sized segment ball control valve for the following process conditions:

Fluid: water;

Pipe size: 6 in;

Maximum design flow, q: 630 gpm;

Pressure upstream of the valve, p_1: 42 psig;

Local atmospheric pressure: 14.7 psia;

Critical pressure, p_C: 3208 psia;

Vapor pressure, p_V: 1.1 psia;

Pressure drop, Δp: 20 psid;

Specific gravity, G_f: 1.0.

1. Calculate the critical pressure ratio

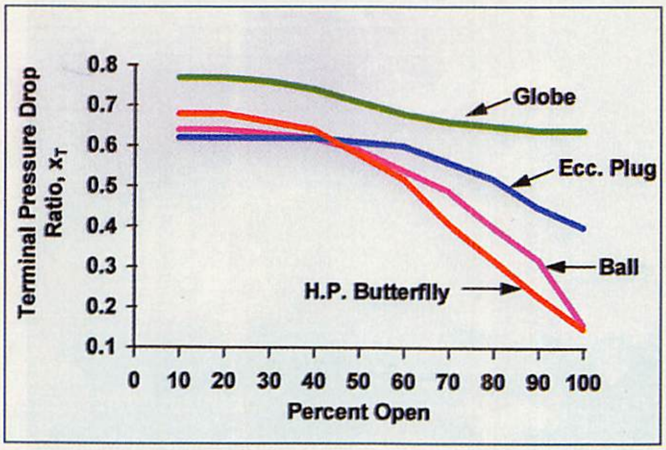

Fig. 2. Terminal pressure drop ratio (X_T) of control valves.

factor, F_F, from Equation 1:

$$F_F = 0.96 - 0.28\sqrt{\frac{1.1}{3{,}208}}$$

$$F_F = 0.955$$

2. Calculate the terminal pressure drop, Δp_T, from Equation 2, using an initial estimate for F_L of 0.72 (from Fig. 1) for a segment ball valve operating at 80% open:

$$\Delta p_T = 0.72^2[(42+14.7)-(0.955)(1.1)]$$

$$\Delta p_T = 28.85\text{psid}$$

3. Because Δp (20 psid) is less than Δp_T (28.85 psid), flow will not be choked. Proceed to Step 5.

4. Skipped per Step 3.

5. Calculate the pressure drop at which cavitation damage is likely to begin (Δp_D) in Equation 3, using the values of R and S for segment ball valves from Table 3 and the same value of F_L as used in Step 2.

$$\Delta p_D = (0.7)(0.72)^2\left[\frac{100}{42+14.7}\right]^{0.2}[(42+14.7)-1.1]$$

$$\Delta p_D = 22.6\text{psid}$$

6. Because Δp (20 psid) is less than Δp_D (22.6 psid), there is no potential for cavitation damage.

7. Calculate the required C_v using Equation 4:

$$C_v = (1.0)(630)\sqrt{\frac{1.0}{20}}$$

$$C_v = 140.9$$

A 3-in segment ball valve is selected from Table 4 that will be slightly less than 75% open, ignoring the effect of pipe reducers, which means that in actual operation with reducers attached to the valve will be less than 80% open.

It can be seen in Fig. 1 that when a segment ball valve's opening decreases, the value of F_L increases, meaning that the initial estimate of F_L (80% open) was on the conservative side and there is no need to repeat the Δp_T and Δp_D calculations.

Gas flow, aerodynamic noise

As with liquids, when a gas flow stream in a control valve reaches the *vena contracta,* the flow velocity increases to a maximum. Assuming a constant pressure upstream of the valve, p_1, increasing the pressure drop by decreasing the downstream pressure, p_2, results in increased flow through the valve until a point is reached where the velocity at the *vena contracta* becomes sonic. Any further increase in pressure drop has no effect on increasing flow. Hence, the flow has become choked.

Unlike liquids where the formulas use pressure drop, Δp, for gases it is more convenient to use the pressure drop ratio, x.

$$x = \frac{\Delta p}{p_1} \qquad (5)$$

The pressure drop ratio at which flow becomes choked when the medium is air is defined as the terminal pressure drop ratio, x_T, and is dependent on valve geometry and the degree of opening. Control valve manufacturers test their valves to determine the values of x_T and publish the results in their valve sizing literature. Some valve manufacturers use other factors (C_f, C_1) for the same thing and are related by the following formula:

$$0.84C_f^2 \approx C_1^2/1{,}600 \approx x_T \qquad (6)$$

Because x_T is determined by tests with air, a correction called the ratio of specific heats factor, F_k (Table 6), is added to account for gases with a sonic velocity that differs from that of air. For any gas, flow will choke when:

$$x = F_k x_T \qquad (7)$$

Sizing method for gases

Constants K_3 through K_6 (Table 7) are included to make the formulas readily adaptable to either customary U.S. or S.I. units. The values of p_1, p_2 and T_1 must be expressed in absolute units.

1. Calculate the pressure drop ratio, x, using Equation 5.

$$x = \frac{\Delta p}{p_1}$$

2. Determine the choked pressure drop ratio from Equation 7.

$$F_k x_T$$

Because x_T is a function of valve opening that is yet to be determined, it will be necessary to make an initial estimate of valve opening. A good initial estimate is the value of x_T at 80% open. Equation 6 is used when a manufacturer publishes a factor other than x_T.

3. If $x \geq F_k x_T$, flow will be choked and $F_k x_T$ is used in place of x in the following calculations:

4. Calculate the expansion factor, Y:

$$Y = 1 - \frac{x}{3F_k x_T} \qquad (8)$$

Under choked flow conditions when $F_k x_T$ is substituted for x, Equation 8 reduces to Y = 2/3.

5. Calculate C_v from one of the following equations.

For gas using volumetric flow units:

$$C_v = \frac{q}{K_3 p_1 Y}\sqrt{\frac{G_g T_1 Z}{x}} \qquad (9)$$

For gas using mass flow units:

$$C_v = \frac{w}{K_4 p_1 Y}\sqrt{\frac{T_1 Z}{xM}} \qquad (10)$$

For vapor or steam using mass flow units:

$$C_v = \frac{w}{K_5 Y \sqrt{x p_1 \gamma_1}} \qquad (11)$$

The compressibility factor, Z, appears in the above equations where density is implied from specific gravity or molecular weight along with pressure and temperature and compensates for the degree to which a particular gas deviates from perfect gas behavior. Using the assumption that Z = 1.0 is usually satisfactory for valve sizing purposes for most industrial gases at the pressures and temperatures at which they are normally encountered in chemical plants.

An appropriate valve size is selected from valve manufacturers tables of C_v versus opening (Tables 4 and 5 are examples of this). The goal is to select a valve that will be as far open as possible without exceeding 80% at the maximum design flow.

For valves installed with reducers, selecting a valve with an opening not exceeding 75% according to the tables will ensure

TABLE 7. NUMERICAL CONSTANTS FOR GAS SIZING EQUATIONS.

	w	q	$p_1, \Delta p$	γ_1	T_1
K_3 1,360		scfh	psi		°R
417		Ncu	bar		°K
	m/hr				
K_4 19.3	lb/hr		psi		°R
94.8	kg/hr		bar		°K
K_5 63.3	lb/hr		psi	lb/cu ft	
27.3	kg/hr		bar	kg/cu m	
K_6 40.4			psi		
61.3			bar		

TABLE 8. PIPE DIAMETER CORRECTION, PDC, FOR AERODYNAMIC NOISE CALCULATION.

Downstream pipe diameter, in	PDC, dBA
1–10	0
12	-1.0
16	-4.0
20	-7.0

TABLE 9. PIPE SCHEDULE CORRECTION, PSC, FOR AERODYNAMIC NOISE CALCULATION.

Downstream pipe schedule	PSC, dBA
40	0
80	-4.0
160	-10.0

TABLE 10. VALVE STYLE CORRECTION, VSC, FOR AERODYNAMIC NOISE CALCULATION.

Valve style	VSC, dBA
Globe	0
Eccentric rotary plug	-1.0
Ball	+3.0
Butterfly	+2.0

that it will not be in excess of 80% open in actual operation with the reducer effect.

After selecting a valve size and determining the percent of opening corresponding to the calculated C_v, the x_T at that opening is determined. If the actual x_T is less than the value used in the preceding steps, repeat the process from Step 2, because the value used for $F_k\,x_T$ is larger than its real value and there is the possibility of overlooking choked flow.

Even if the smaller and correct value of x_T does not predict choked flow, the new value of x_T should be used to repeat the calculations, because calculated C_V will increase with decreased x_T and the valve opening in actual operation will be greater than that just predicted. If the actual x_T is greater than the value used in the calculation, you have made a conservative calculation.

6. Calculate the noise level:

$$SPL = 14\log(C_v) + 18\log(p_1) + 20\log\left[\log\left(\frac{p_1}{p_2}\right)\right] + K_6 + PDC + PSC + VSC \qquad (12)$$

Tables 8 through 10 contain values for PDC, PSC and VSC. Aerodynamic noise levels of 110 dBA or greater when calculated for uninsulated Schedule 40 pipe will most likely result in severe vibration damage. Also, most plant standards limit noise to 90 dBA or less.

C_v and the noise level should also be calculated at the minimum design flow. Making the initial estimate of valve opening for the purpose of obtaining a starting value for x_T may require some trial and error. Estimating an opening of 40% is a good starting point. It is worth noting that p_1 and Δp will most likely be higher at the minimum design rate than they are at the maximum design flowrate.

Gas sizing example

The stated problem is to select a properly sized equal-percentage globe control valve for the following process conditions:

Fluid: nitrogen;
Pipe size: 3-in Schedule 40;
Maximum design flow, q: 130,000 scfh;
Upstream pressure, p_1: 105 psig;
Local atmospheric pressure: 14.7 psia;
Upstream temperature, T_1: 100ºF;
Compressibility factor, Z: 1.0;
Pressure drop, Δp: 40 psid;
Specific gravity, G_g: 0.97;
Ratio of specifics heats factor, F_k: 1.0.

1. Calculate the pressure drop ratio, x, from Equation 5.

$$x = \frac{40}{105+14.7}$$
$$x = 0.33$$

2. Calculate the choked pressure drop ratio, $F_k\ x_T$, using Equation 7. An initial estimate for x_T of 0.65 is used for a globe valve operating at 80% open (Fig. 2).

$$F_k x_T = (1.0)(0.65)$$
$$F_k x_T = 0.65$$

3. Because $x < F_k\ x_T$ (0.33 < 0.65), flow will not be choked. Therefore the actual value of x = 0.33 is used in the following calculations.

4. Calculate the expansion factor, Y, using Equation 8:

$$Y = 1 - \frac{0.33}{(3)(0.65)}$$
$$Y = 0.83$$

5. Calculate the required C_V using Equation 9 for volumetric flow units because it corresponds to the given process data:

$$C_v = \frac{130{,}000}{1{,}360(105+14.7)(0.83)}\sqrt{\frac{(0.97)(100+460)(1.0)}{0.33}}$$
$$C_v = 39$$

From Table 5, it can be seen that a 3-in globe valve will be about 60% open and a 2-in globe valve will be about 75% open. The 2-in valve is selected because it is the closest to 80% open. Fig. 2 shows that x_T at the actual opening of about 75% is for all practical purposes the same as the 80% open value of 0.65 that was used in the calculations.

6. Calculate the noise level using Equation 12:

$$SPL = 14\log(39) + 18\log(105+14.7) +$$
$$20\log\left[\log\left(\frac{105+14.7}{65+14.7}\right)\right] + 40.4 + 0.0 + 0.0$$
$$SPL = 85\,\text{dBA}$$

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