MARY KAY O'CONNOR
PROCESS SAFETY CENTER
TEXAS A&M ENGINEERING EXPERIMENT STATION

19th Annual International Symposium
October 25-27, 2016 • College Station, Texas

Make Inlet Piping and PSV One-System

Boyd George
B11846@aol.com

Abstract

For pressure safety valves and associated inlet piping, the API recommends the non-recoverable pressure loss should not exceed 3 % of the set pressure at rated capacity flow, with some exceptions, e.g., for remote sensing pilot operated pressure safety valves. The API further notes pressure losses above 3 % are allowable if an engineering analysis shows valve performance is not impacted during relief. The API provides little guidance on the recommended engineering analysis. Calculations show the inlet piping and pressure safety valve (PSV) should be considered one system. Analysis of the system improves the basis for judging pressure safety valve performance, especially when compared with treating inlet piping separately from the PSV and then somewhat arbitrarily judging performance adequacy. With increasing inlet piping pressure loss, the energy in the velocity head at the inlet to the PSV grows in significance. Analyzing the inlet piping pressure loss separate from the PSV neglects this energy and makes the experience and knowledge of the judging engineer paramount. Given current computational capability and today's litigious regulatory environment, an analytical and consistent basis for judging PSV performance may be of interest.

Introduction

Much has been written about Pressure Relief Systems. ASME Code, API Recommended Practices and Standards, Institute of Petroleum and Energy Institute analyses, and other industry related groups generating innumerable literature articles, theses, and dissertations. In its entirety, the publications form the basis of the *"Recognized and Generally Accepted Good Engineering Practices,"* or RAGAGEP, which operating organizations use as their benchmark for pressure relief systems. A pressure relief system is typically inlet piping, a pressure relief or safety valve, and outlet piping.

The API[i] notes: *"… any pressure drop in the inlet pipe will reduce the relieving capacity."* API further notes that when the pressure drop in the inlet piping to a pressure safety valve (PSV) exceeds 3 % of the PSV setpoint, an engineering analysis affirming the necessary pressure protection *should* be done. An engineering analysis can show assured over-pressure protection with pressure drops in excess of 3 %.

The API's expressed concern about the 3 % threshold is the possibility of *chattering* or instability of the PSV, a condition that could not be assessed without also knowing the blowdown pressure[ii] of the PSV. Excessive pressure drop in the inlet piping and/or an oversized PSV could chatter and self-destruct resulting in a loss of containment. For the best discussion of the *"3 % rule,"* and valve instability, refer to the Occupational Safety and Health Review Commission Docket No. 10-0637. *"After conducting a $30,000,000.00 study to determine how best to ensure valve stability, the EPRI concluded there was no correlation that would predict whether or not a valve would become unstable.*[iii] No correlation exists that predicts PSV instability meaningfully and the 3 % threshold has no basis beyond industry acceptance.

This work argues that the estimated capacity determination requires *joint* analysis of the inlet piping and the PSV, regardless of the percentage pressure drop in the inlet piping. Additionally, the designer or performance rater should be aware of the PSV blowdown pressure.

Common Practice for Capacity Determination

The most common pressure relief flowing situation is critical gas or vapor flow behavior. Alternative conditions are sub-critical gas or vapor flow behavior and liquid flow behavior. Only super-critical gas will be discussed herein, though the argument for treating the inlet piping and PSV as one system extrapolates to sub-critical gas and liquid relief as well.

Critical flow behavior for an ideal gas or vapor can be estimated using the equation[iv]:

$$\begin{aligned}
\frac{P_{cf}}{P_1} &= \left(\frac{2}{k+1}\right)^{k/(k-1)} \\
P_{cf} &\equiv Critical\ flow\ nozzle\ pressure,\ psia \\
P_1 &\equiv Upstream\ relieving\ pressure,\ psia \\
k &\equiv Ideal\ gas\ heat\ capacity\ ratio,\ {C_p}/{C_v}
\end{aligned} \qquad (1)$$

Critical flow occurs when the downstream pressure of the PSV is equal to or less than the critical flow nozzle pressure. For the PSV preliminary sizing for ideal gas the API[v] offers:

$$
\begin{aligned}
&A = \frac{W}{CK_d P_1 K_b K_c}\sqrt{\frac{TZ}{M}} \\
&A \equiv Required\ discharge\ area,\ in^2 \\
&W \equiv Required\ \ flow\ rate,\ {}^{lb_m}\!/\!_{hr} \\
&C \equiv 520\sqrt{k\left(\frac{2}{k+1}\right)^{\frac{k+1}{k-1}}} \\
&K_d \equiv Effective\ discharge\ coefficient \\
&P_1 \equiv Upstream\ relieving\ \ pressure,\ psia \\
&K_b \equiv Backpressure\ capacity\ correction\ \ factor \\
&K_c \equiv Rupture\ disk\ combination\ correction\ \ factor \\
&T \equiv Relieving\ temperature,\ {}^{\circ}R \\
&Z \equiv Compressibility\ \ factor\ at\ T\ \ and\ \ P_1 \\
&M \equiv Molecular\ \ weight,\ {}^{lb}\!/\!_{lb-mole}
\end{aligned}
\tag{2}
$$

API offers an alternative method of sizing when the gas is not ideal. This equation is also considered valid for two-phases –Homogenous Equilibrium Model (HEM) – using volumetric averaged densities. It is more rigorous though kinetic energy may not be adequately addressed. It is a numerical integration of the isentropic nozzle flow[vi]:

$$
\begin{aligned}
&G^2 = \left[\rho_t^2\right]\cdot\left(-2\int_{P_0}^{P}\frac{4633dP}{\rho}\right)_{\max} \\
&G \equiv Mass\ \ flux,\ {}^{lb}\!/\!_{sec-ft^2} \\
&\rho \equiv Density,\ {}^{lb}\!/\!_{ft^3} \\
&P \equiv Stagnation\ \ pressure,\ psia \\
&0 \equiv Inlet\ to\ nozzle \\
&t \equiv Throat
\end{aligned}
\tag{3}
$$

And with G determined, the required A can be found using:

$$
\begin{aligned}
&A = \frac{0.04W}{K_d K_b K_c K_v G} \\
&K_v \equiv Viscosity\ correction\ \ factor \\
&\quad for\ liquids \\
&\quad = \left(0.9935 + \frac{2.878}{Re^{0.5}} + \frac{342.75}{Re^{1.5}}\right)^{-1.0} \\
&Re \equiv Reynolds\ number
\end{aligned}
\tag{4}
$$

The actual area would be determined using the next larger area from API 526[vii], and the actual capacity of the system is determined by ratio of the actual area to the required area. All piping hydraulic losses would be determined using the actual capacity, which is what the valve would pass when opened. Note: in a modulating PRV, the actual capacity is the same as the required capacity.

Alternatively, choked flow can be determined rigorously by iterating on the pressure at the throat, P_t, isentropically determining the temperature where the stagnation enthalpy, H_1, determined at the upstream stagnation pressure, P_1, and the enthalpy at the throat, H_t, with the velocity head at sonic velocity (ft/sec) or kinetic energy are equal. In equation form[viii],

$$
\begin{aligned}
& H_1 - H_t - \frac{a_t^2 - a_1^2}{2} = 0 \\
& H_1 \equiv {}^{Btu}\!/_{lb-mole} \ \ at\ stagnation \\
& H_t \equiv {}^{Btu}\!/_{lb-mole} \ at\ throat \\
& a_t \equiv sonic\ velocity\ at\ throat \\
& a_1 \equiv upstream\ velocity, usually \approx 0
\end{aligned}
\qquad (5)
$$

The above equation is a rigorous determination of the capacity, along with K_d and other adjustment coefficients of a pressure relief system. The stagnation entropy should be adjusted with the adiabatic inlet pipe pressure drop using the equation:

$$
\begin{aligned}
& \Delta S = c_p \ln \frac{T_2}{T_1} - R \ln \frac{P_2}{P_1} \\
& T_1, P_1 \equiv Inlet\ \ pipe\ inlet\ conditions \\
& T_2, P_2 \equiv Inlet\ \ pipe\ exit\ conditions
\end{aligned}
\qquad (6)
$$

This entropy adjustment is usually small. It adds precision to the calculation.

Common Practice for Inlet Pipe Pressure Drop

A number of pressure drop correlations exist. This work allows the choice of one of three commonly available correlations, Crane[ix], Beggs and Brill[x], and the Fanno equation. For the friction factor, Colebrook[xi] (aka Moody) is used.

$$\frac{1}{\sqrt{f}} = -2\log\left(\frac{\varepsilon}{3.7D} + \frac{2.51}{Re\sqrt{f}}\right)$$
$$f \equiv Friction\ \ factor$$
$$\varepsilon \equiv Pipe\ roughness, inches$$
$$D \equiv Pipe\ inside\ diameter, inches$$
$$Re \equiv Reynolds\ number$$

The correlations give differing inlet pressure drop results suggesting a comparison with plant data to discern the correlation best suited for any given operation.

Common Scenario for Capacity Estimation

One common practice scenario is the use of Equation (2) to estimate the required area. Use API 526 to obtain the actual area. Determine the *"actual"* capacity of the PRV. Use this *actual* capacity to determine the inlet pressure drop. If below the 3 % threshold, calculations are typically complete. If the inlet pressure drop is 3 % or higher, some additional work may be done ranging from increasing the inlet pipe diameter, and/or reducing the relieving pressure by the inlet pressure drop and re-running the calculations. This assumes a linear relationship between inlet pressure drop and PRV performance, which may not be the case. Or, choose a pilot operated PRV (POPRV) for the service, amongst other alternatives, design complete.

During the design phase of a project, it is not unusual for the piping design to be on the critical path of the project. Typically, PRV's and associated piping are sized during this phase. The attention paid to the PRV system design varies because of project cost and schedule pressures. Also, the experience of the designer and their awareness of operating the equipment are significant as well as the discipline of the project manager overseeing the work and who's also accountable for managing the cost and schedule pressures. Summing up, judgment is required on the part of the designer and the project manager during the design phase. This judgment could be enhanced with rigorous design tools, typically in the form of properly used computer programs.

A statistical analysis[xii] was done on 27,000 PRV's that most likely were completed with the sizing scenario, or a variation, outlined above. Their analyses showed about 64 % of the PRV population met the recommended practices. The rest were either missing PRV's, undersized or improperly installed including pressure drop issues. Assuming the 27,000 PRV's are representative of industry, one might say industry performance is less than impressive. And our increasingly litigious society is quick to hold operators accountable for any incident related to a failure of a pressure relief system to perform.

Some Comparisons

Table 1 shows some capacity comparisons using the equations above coupled with the inlet pipe pressure drop (Fanno equation) with an example:

Table 1

Capacities

API Equations (3) & (4)		Equations (5) & (6)	
Inlet ΔP as % of Setpoint	ΔW_{act} as % of Actual	Inlet ΔP as % of Setpoint	ΔW_{act} as % of Actual[1]
0	0	0	-0.16%
1.68 %	-1.67%	1.60 %	-2.86%
2.93 %	-2.91%	2.86 %	-4.87%
5.93 %	-5.84%	5.65 %	-8.90%

1. API equations (3) & (4) basis

Using API equations, (3) & (4), shows the inlet pressure drops may appear to be approximately linear in capacity reductions. However, the rigorous equations (5) and (6) give a somewhat different indication. Again, the difference between equations (3) & (4) with (5) & (6) is that (3) & (4) do not appear to include a complete kinetic energy analysis. The increasing ΔP's resulted from decreasing inlet pipe diameter with constant equivalent length. Note: using common industry practice, only the last data point exceeds 3 % and might get additional scrutiny.

A Word on Thermodynamics

A rigorous design tool, like a computer program, must have a solid thermodynamic basis for meaningful estimates of fluid behavior. The equation of state chosen for this work, i.e., coupling the inlet pressure drop with rigorous PRV analysis, is the Benedict-Webb-Rubin-Starling (BWRS) using the constants and interaction parameters published by Exxon[xiii]. This equation has been shown to represent the fluid behavior of hydrocarbon systems better than alternative equations[xiv]. Except in the vicinity of the critical conditions, the Peng-Robinson equation represents hydrocarbon fluid behavior adequately except for liquid densities – unless specifically tuned via constants and interaction parameters to match liquid densities.

Equations of state can yield erratic results, especially with the second derivative properties near critical conditions. Heat capacities and sonic velocity are second derivative properties. Most often equation of state constants are regressed from PVT (density) data. The Exxon BWRS constants were regressed using any available data[xv] at the time, including sonic velocities and heat capacities if available as well as PVT data.

Obviously, a good sonic velocity estimate is an important part of linking the inlet pipe pressure drop to the performance of the PRV in a rigorous calculation. Ideal gas sonic velocity estimate may differ sufficiently from actual sonic velocity as to introduce some error, though the magnitude –and significance– of this error is not known. This error can show up in at least a couple of areas, when using the Fanno equation for pressure drop estimates as well as the PRV throat. And a good estimate of the sonic velocity also plays an important role in awareness or prevention of vibration induced piping failures, e.g., acoustic or flow induced vibration. Thus, selection of the equation of state should not be arbitrary.

Future Explorations

The tail pipe should be added to the inlet pipe and PRV system. Currently, for tailpipe calculations, the API[xvi] suggests going to the known pressure, typically the tailpipe exit, then backing into the system. This is appealing because the low pressure typically results in larger piping. However, such an approach doesn't appear to address the possibility of choked flow. Additionally, larger piping may be more susceptible to vibration issues.

---

[i] API RP 520, Part II, Fifth Edition, August 2003, ¶ 4.2.3.1 *Inlet Pipe Loss*.
[ii] Section VIII – Division I, *Nonmandatory Appendix M Installation and Operation*, ASME Boiler and Pressure Vessel Code
[iii] OSHRC Docket No. 10-0637, Secretary of Labor v. BP Products North America & BP-Husky Refining, LLC.
[iv] API Standard 520, Part I, Eighth Edition, December 2008
[v] Ibid
[vi] Ibid
[vii] *Flanged Steel Pressure Relief Valves*, API Standard 526.
[viii] Haque, Richardson and Saville, *Blowdown of Pressure Vessels*, Trans IChemE, Vol 70, Part B, February 1992
[ix] Flow of Fluids Through Valves, Fittings, and Pipe, Technical Paper no. 410, Crane Co.
[x] Dale Beggs and James Brill, *A Study of Two-Phase Flow in Inclined Pipes*, Journal of Petroleum Technology, SPE, May 1973
[xi] C. Colebrook, *Turbulent Flow in Pipes, with Particular Reference to the Transition Region Between the Smooth and Rough Pipe Laws*, J. Inst. Civ. Eng., vol. 11, London, 1939
[xii] P. Berwanger, R. Kreder, Wai-Shan Lee, *Non-Conformance of Existing Pressure Relief Systems with Recommended Practices: A Statistical Analysis*, Berwanger, Inc., 2002.
[xiii] S. Hopke, C. Lin, *Application of the BWRS Equation to Natural Gas Systems*, 76th National AIChE Meeting, 1974.
[xiv] B. George, *Solving Unusual Design Problems Using Equations of State*, 61st Annual GPA Convention, 1982. And an internal ARCO study on the CGF of the North Slope coming to the same conclusion.
[xv] Personal communication, S. Hopke and B. George
[xvi] API RP 521