
# Pressure Safety Devices

Course No: C01-040
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## Introduction

BPVC requires pressure safety devices on pressure vessels under most scenarios. That word, "scenario" is very important to the understanding of PSD requirements. If you have a pressure vessel rated for 1 million psig [6900 MPag], and your most energetic pressure source is 100 psig [690 kPag] then you can apply engineering judgement and say that there is no need "in this scenario" to provide PSD, but there may be things that can happen (i.e., other scenarios) that do put the vessel at risk and require PSD.

The BPVC has generic PSD requirements that apply to every industry. For Oil & Gas the API has published API 521, Seventh Edition, 2020 to provide guidance in determining required relieving capacity. This standard draws heavily on the BPVC but is specific to Oil & Gas. The scope of API 521 requires that you **evaluate** any vessel that is built to a pressure vessel code for overpressure protection. Once the required capacity is determined, API 520, tenth edition, October 2020 provides the methods and limitations for calculating the relieving device expected flow rate.

For the most part API 521 only applies to things designated by code as "pressure vessels". API 521 looks to piping design codes (e.g., ASME B31.8 for gas piping) for pipelines and pipeline accessories. There are some notable devices that API 521 would exempt from an API 521 analysis that regulations have specified must be considered for PSD protection. The most common regulatory add-on is that in many jurisdictions, pig launchers and receivers must be evaluated for overpressure protection. Because of the potential for a launcher or receiver to be isolated while liquid full, this is a quite reasonable requirement.

## Credible Scenarios

The API and ASME started moving away from "Equipment-based sizing" to "System-based sizing" in about 1997. Equipment-based sizing looked at the size of a vessel and designated a relieving capacity based on that vessel size without considering what inflow rates were possible for a given location or service. Equipment-based sizing often resulted in overpressure-protection that was significantly oversized.

Credible-scenario based sizing is based on the capacities of the system where the vessel is installed. It looks at what can credibly happen and the volume requirement if that thing does in fact happen. API 521 lists 17 major categories that need consideration:

1. Closed outlets
2. Cooling-water failure to condenser
3. Top-tower reflux failure
4. Sidestream reflux failure
5. Lean-oil failure to absorber
6. Accumulation of non-condensables
7. Entrance of highly volatile material
    a. Water into hot oil
    b. Light hydrocarbons into hot oil
8. Overfilling
9. Failure of automatic controls
    a. Inlet control devices and bypasses
    b. Outlet control devices
    c. Fail-stationary valves
    d. Choke valves
10. Abnormal process heat or vapor
    a. Abnormal process heat input
    b. Inadvertent valve-opening
    c. Check-valve failure
11. Internal explosions or transient pressure surges (e.g., water, steam, or condensate hammer)
12. Chemical reaction
13. Hydraulic expansion
    a. Cold fluid shut in
    b. Lines outside process area shut in
14. Exterior fire
15. Heat transfer equipment failure
    a. Heat exchange tube rupture
    b. Double pipe
    c. Plate and frame
16. Power failure (steam, electric, or other)
    a. Fractionators
    b. Reactors
    c. Air-cooled exchangers
    d. Surge vessels
17. Maintenance

Many of these categories are never credible on a well site (e.g., #3 Top tower reflux failure), some of them are always credible (e.g., #1 Closed outlets). One useful approach is to evaluate a site and select the categories that have a reasonable chance of being credible. Virtually every well site needs to be evaluated for #1 Closed outlets, #9 Failure of automatic controls (the sub categories of that category are not often useful for well sites), #13 Hydraulic expansion (mostly "a. Cold fluid shut-in"); and #14 Exterior fire. An increasing number of well sites also require you to analyze #16 Power failure. **Error! Reference source not found.** is an extract from a credible scenario analysis for a specific field.

Table 1: Example Credible Scenario Analysis

| | | |
|---|---|---|
| PSV No: 4 | Equip type: Suction Scrubber | MAWP: 500 psig |
| PSV Manufacturer: Mercer | PSV Part No: 91-43F51V07P21 | PSV Size: 2x2 "F" orifice |
| PSV Press Setting: 100 psig | PSV calc flow: 972 MSCF/day | Max credible flow: 116 MSCF/day |
| Recommended Changes | None | |
| Possible Scenarios | Closed outlet | A blocked outlet could put reservoir pressure on the vessel in excess of PSV setpoint (116 MSCF/day required capacity) |
| | Fail Auto cntl | No credible scenario |
| | Hyd Exp | Not credible. While a plugged drain line could result in the vessel becoming liquid full, someone would have to manually shut the inlet and outlet valves after the vessel filled, requires too many unrelated activities. |
| | External fire | Not credible because there is no method of getting the vessel liquid full and isolated without failures that would be unrelated to the fire |

The "PSV Part Number" in **Error! Reference source not found.** is a manufacturer pneumonic code. The "91" says it is a Mercer 9100 series. The "43" says it is a 2-inch Male NPT threaded inlet and a 2-inch Female NPT threaded outlet. The "F" refers to the API orifice designation (see **Error! Reference source not found.**). The "5" says that it is made of carbon steel. The "1" says it is a closed cap. The "V" specifies a Floroelastomer seat material designated FKM. The "07" is Mercer's spring code. The "P" specifies the trim material. The "2" specifies the o-ring material is FKM 75 Duro. The final "1" is a special code assigned by Mercer. Other manufacturers have different pneumonic codes, but all provide basically the same information.

The fire case seems to be the most controversial. The coefficient of thermal expansion of water is about 100 psi/R [1.2 MPa/K], so if a vessel is liquid full, even a low-grade external fire can rapidly raise the pressure within the vessel. Gas is a lot more forgiving. Raising the temperature of a vessel filled with gas (no liquid) from 60 °F [15.6 °C] at 20 psia [137.9 kPaa] to 61 °F [16.1 °C] would raise the pressure to 20.038 psia [138.2 kPaa], so it is clear that an external fire does not have the potential to rapidly increase pressure in a gas-filled vessel. If there is some liquid, it will boil off which will raise vessel pressure, but when you do the volume/pressure calculations they rarely work out to having enough mass of liquid in the vessel to raise the pressure significantly.

The fire case is limited to pool fires (i.e., a standing pool of liquid is burning) since a jet fire directed at a vessel would cause the vessel to fail long before pressure could rise significantly from expanding fluids.

Not considering the fire case will often draw loud and extensive criticism of your work and your overall intelligence. This prejudice is as silly as any other pre-judgement, but it is extremely common in our industry. Most of the "you always have a fire case" proponents will accept a reasoned analysis of the actual conditions, but not always. I've seen analysis where it was perfectly clear that the fire case was not a credible overpressure scenario because there were not even any flammable liquids stored on the location (to say nothing of it being impossible to isolate the vessel liquid full) be rejected because "the fire case is always credible". When confronted with that sort of unreasoning non-logic it is best to pretend that there is a credible fire case and set the required capacity less than the highest flow of actual credible scenarios.

When your inflow source is a reservoir, people frequently do an Absolute Open Flow calculation and develop a flow rate based on the rate the reservoir would flow into atmospheric pressure. This is not what the reservoir is actually seeing. In the example in **Error! Reference source not found.**, the PSV is set at 100 psig [690 kPag] and the well was flowing into a 30 psig [207 kPag] compressor suction. If you slam the outlet valve shut, the well will want to flow at 30 psig. As pressure builds up, the flow rate will drop off at a rate proportional to the difference between the square of reservoir pressure and the square of bottom-hole pressure. The proper flow rate to use is based on the reservoir flowing into the PSV set pressure.

For every category that the pre-analysis labeled as "possible", the designer must first determine if they are credible, and if so, what is the source of the overpressure and how quickly that source could replenish mass lost through the PSD. Then select the largest flow rate to check against the PSD flow rate. It is common for two (or more) scenarios to be credible with vastly different required flow rates. In that case it is reasonable to size a PSV for the lower case and put its setpoint at a bit less than Maximum Allowable Working Pressure (MAWP). Then you can size a second PSV (or rupture disk) for the higher requirement set at 105% of MAWP. A common reason for two very different scenarios is "13. a) Cold fluid" shut in which can generate substantial pressure with a small change in temperature and ""1 Closed outlet" which can put full reservoir pressure on the vessel. A PSV designed for cold fluid would be very close to the size and complexity of the pressure relief on a domestic water heater while one designed for closed outlet might be huge.

## Double Jeopardy

The design of pressure safety devices is centered around "credible scenarios". "Credible" is a very important concept. While it is often credible that an operator might shut the outlet of a pressure vessel with a pressure source still able to come in through the inlet, it probably isn't credible that at the same instant in time the displacer fell off the level controller—the two events are unrelated and it stretches credibility that they would happen simultaneously. You do not have to consider multiple, unrelated events occurring simultaneously as long as they really are unrelated.

On the other hand, cascading scenarios may be very important. For example, a pool fire may burn through the gas-supply tubing to the automation equipment which would cause fail-closed valves to go shut. If a site has an emergency shut down process then the "XV" would go shut on loss of control gas and now we have both an external fire and a closed outlet. That scenario is quite credible and should be considered in setting a required flow rate.

## Set Points

Vessel Maximum Allowable Working Pressure (MAWP) is set by the design and the design codes. Over pressure protection needs to allow adequate outflow to prevent the inflow from increasing the pressure to more than 110% of MAWP while the relief device is open (121 % of MAWP for fire case). These limitations on pressure accumulation are described in the BPVC not in the current version of API 521. For vessels with multiple credible scenarios (each with its own flow rate) it would be reasonable to install a device with a small relieving capacity (sized for the smaller credible scenario) set at MAWP and a larger device set at 105% of MAWP with a combined flow rate large enough to prevent the pressure from building up. In the example in **Error! Reference source not found.**, the vessel MAWP is 500 psig [3.45 MPag] and the device set point is 100 psig [690 kPag] with a flow rate 8 times larger than the maximum credible flow rate—certainly adequate to keep the vessel below 550 psig [3.79 MPag]. The relief valve set point was set this low because the operators had elected to install a lower pressure accessory on the device which required re-rating the system. BPVC specifies a maximum pressure accumulation during an overpressure event, any combination of set point and flow rate that prevents exceeding this maximum pressure accumulation is acceptable.

Vessel designers frequently install multiple identical relieving devices in lieu of the customer supplying the result of a credible-scenario analysis. Multiple devices can be quite reasonable, but there needs to be separation between the device set points. Every type of pressure-relieving device has an uncertainty or dead-band value. These can be small, but they cannot be zero. Consequently, two parallel devices with the "same set point" will have one that falls lower in its dead-band than the other device. In an over-pressure scenario, the one with the lowest set point will lift, lowering system pressure. In most cases the second device will not open at all, but when it does it will tend to cause the other device to chatter between open and closed which can damage a device. Because of this dead-band issue, you can't take credit for the second device and each of the two devices must be capable of passing the entire inflow. In order to take credit for both devices their set point needs to be different by at least twice the magnitude of the dead band (i.e., if the dead band is 5 psi [34.5 kPa] then the set points need to be separated by at least 10 psi [69 kPa]).

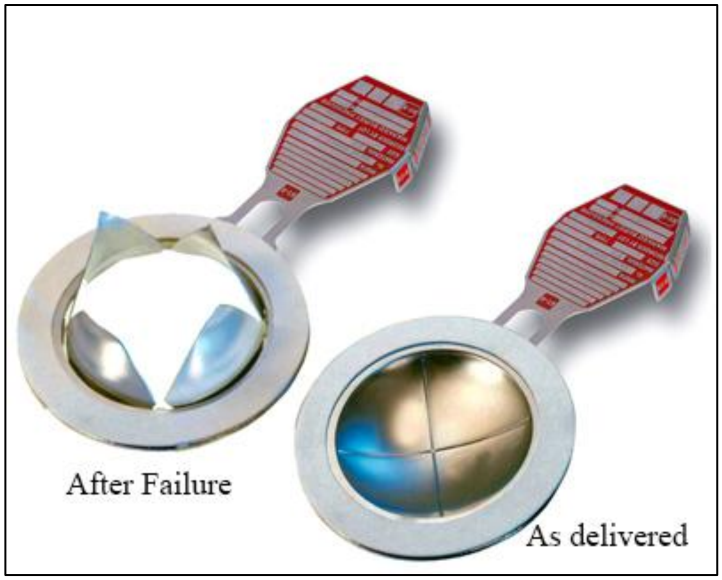



Figure 1: Rupture Disk

## Devices

There is a range of equipment designated as "pressure safety devices". These range from one-time use devices that fail catastrophically and cannot be reset to devices designed for such a small operating range that it is very common for them to lift and reseat (e.g., tank pressure/valve vents). In between these two ranges are devices which are intended to operate very rarely and their operation constitutes a significant safety event (as opposed to a normal "control event").

It is important to note that any device controlled by field-automation is a "control device" not a "PSD". All pressure safety devices must be able to operate independently from program logic. Having a blowdown valve that is part of a process control may be a perfectly appropriate control feature, but it is not part of the over-pressure protection.

### Rupture disk

At the lower range of operational complexity is the "rupture disk". These devices are designed to have a ductile failure at a particular differential pressure. The "ductile failure" indicates that the metal will rip instead of shattering (and potentially sending projectiles downrange). When a rupture disk fails, it is unable to later close and stop flow, it simply becomes an open pipe that must be physically replaced prior to returning to service.

Rupture disks (**Error! Reference source not found.**) are very subtle engineering creations. Ductile failure of a metal doesn't usually happen at a precisely defined stress. If you take one hundred "identical" pieces of steel and test them to failure, the failure point will exhibit a range of values. For rupture disks, that range must be exceedingly small. Consequently, rupture disks must be handled with great care. Any scratch or dent should obviously cause them to be discarded, but even the oil from a worker's hands can significantly change the failure point. The frequency of failures caused by mishandling of rupture disks has soured many organizations on their use and some companies have banned their use, therefore in cases where it seems that rupture disks would be a good choice it is useful to check company polices before committing much effort towards using them.

Rupture disks can have a secondary function of protecting direct or pilot operated PSVs from corrosive or toxic fluids. In that service, the failure point of the rupture disk is largely irrelevant (as long as it is below the setting of the PSV), and its reason for being installed is to protect the internal mechanism for the PSV from becoming fouled by process fluids. Installing a rupture disk below a PSV changes the predicted flow rate of the PSV which is accounted for in the flow calculation below.

## Direct-operated pressure safety valve (PSV)

Direct-operated PSV (**Error! Reference source not found.**) use spring tension to hold a disk on a sealing surface. When the net differential forces on the disk are greater from the process fluid than from the spring and exhaust pressure, the disk comes off the seat. These net differential forces are made up of: (1) process pressure applied over the (fairly large) surface area below the disk; (2) exhaust pressure applied over the (smaller) surface area above the disk; and (3) spring force above the disk. If the PSV exhausts into local atmosphere (i.e., the exhaust piping is short, large, and unobstructed) the set point of the PSV is determined with local atmospheric pressure above the disk (which is why you do PSV calculations in gauge pressure instead of absolute pressure). When a PSV is installed to exhaust into a flare header, imposed backpressure (say from other PSV lifting at the same time) shifts the set point upward while reducing the flow rate.

A direct-operated PSV is not a “snap acting” device (i.e., it does not pop from shut to fully open), and a slight overpressure will just crack the stem head off the seat. It is common for a PSV to leak after an overpressure event, and these valves should never be a part of a system control scheme—they are perfectly acceptable as part of safety system, not a control system.

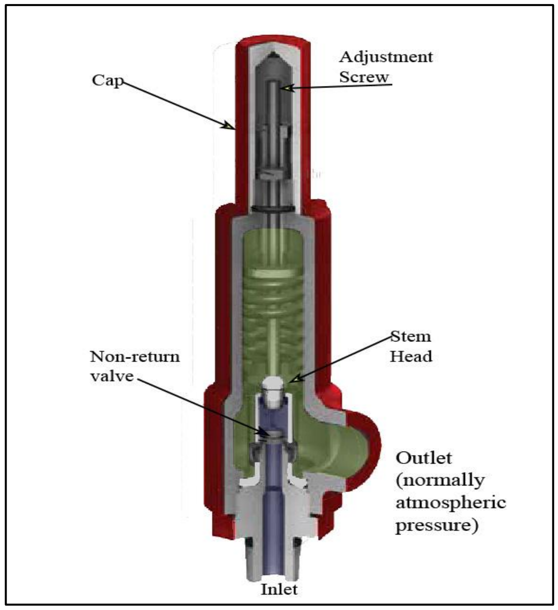



Figure 2: Direct Operated Pressure Relief Valve

In vacuum operations, all the net forces are pushing the disk towards the seat, but the valves are designed to prevent backflow and when exhaust pressure is higher than process pressure the PSV will pass exhaust pressure into the process. Conventional PSV can be used in vacuum service with rupture disks placed under the valve.

## Pilot-operated PSV

In a pilot operated PSV (**Error! Reference source not found.**), there is a “balance piston” that is held shut by a (non-adjustable) weak spring and process pressure above the piston. The area above the balance piston is connected to process pressure through a secondary device. The secondary device has a much stronger (adjustable) spring that is adjusted to hold process pressure up to its set point. Above the set point, the pilot plug opens and vents the area above the balance piston, opening the PSV.

The pilot plug is very small, and so is the volume being vented. A slight over pressure will crack the pilot plug and begin to lower the pressure above the balance piston, when the pressure above the balance pressure (plus the downforce of the weak spring) is lower than process pressure, the balance piston moves up and the relief process begins. This

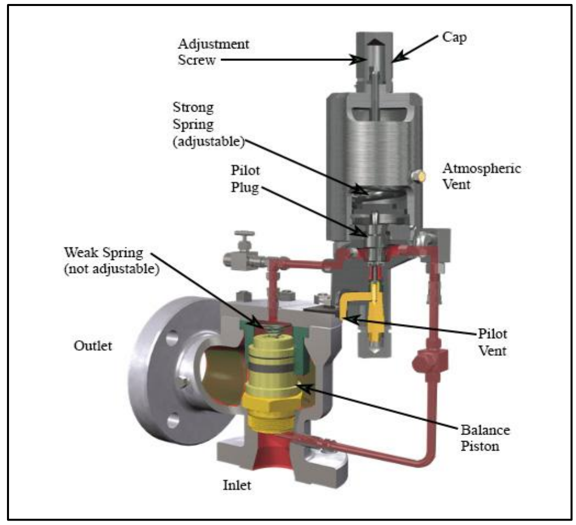



Figure 3: Pilot Operated Pressure Relief Valve

can result in a very slight flow through the main valve (and the pilot acts like a "continuous bleed controller", which is described in "Wellsite Pneumatic Control", Course No. __). If process pressure continues to increase, the pilot will eventually (actually, within a few seconds) come fully open causing the balance piston to come fully open.

Again, the set point is a differential pressure between process pressure and atmospheric pressure and flow calculations are done in gauge pressure. Since the pilot exhausts into local atmospheric pressure, relief piping length, size, and obstructions do not impact the set point, but outlet configuration does impact the relief flow rate. This makes Pilot-Operated valves the preferred choice for processes that must exhaust into a flare header.

Manufacturers often market Pilot-Operated Pressure Relief Valves as "control devices", and they do work in this service but the shape of the Balance Piston and its seat do not lend themselves to precise control and there are other backpressure and pressure regulator valves that do a better job. A Pilot Operated Pressure Relief Valve in control service cannot be considered part of any safety system.

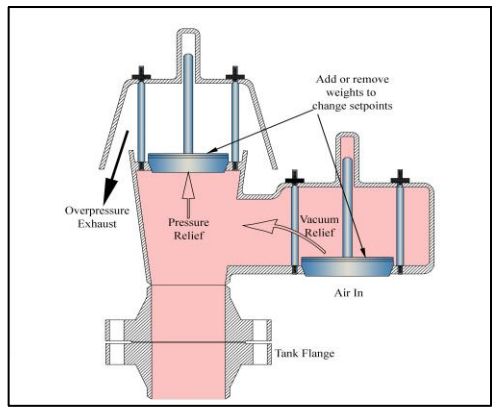



Figure 1: Tank Pressure/Vacuum Vent

Pilot-operated PSV do not add any appreciable operational-value when installed with the PSV exhausting to atmosphere with a very short tail pipe and were historically very rare on well site equipment. With regulatory proscriptions on venting raw gas to atmosphere, flare headers have become much more common on well sites and consequently the need for pilot operated PSV has increased.

Liquid-storage tanks are quite unable to operate with any significant pressure or vacuum. Consequently, equipment designed to protect tanks ( Figure 1) tends to operate far more often than other PSD equipment. Outgassing puts increasing mass into the vapor space, increasing pressure. Throughout the day, even an unheated tank will release pressure through the pressure/vacuum vent, and every time atmospheric pressure drops, more gas will leave the liquid. Throughout the night, most tanks will have some amount of condensation, and without the vacuum relief would risk collapsing the tank. Both valve paths on this device regularly have flow.

## Inspection and Testing

When a Pressure Safety Valve is purchased, the purchaser specifies a set pressure, flow rate, backpressure, and allowed over pressure. Then the valve spring is adjusted, a tag (**Error! Reference source not found.**) is created, and a calibration report is prepared. The entity selling the device (usually a local distributer) will calibrate the spring tension to cause the valve to lift at a specified pressure and will provide a calibration certificate. That certificate has a very definite shelf-life. API RP 576 Inspection of Pressure-relieving Devices (Fourth Edition, April, 2017) provides excellent guidance on how to inspect, what to look for, how to verify set point (both “as received” and “as reconditioned”), and how to transport, but it does not provide anything about how often to perform these tests. RP 576 says that it must be done “periodically”, but not on what period.

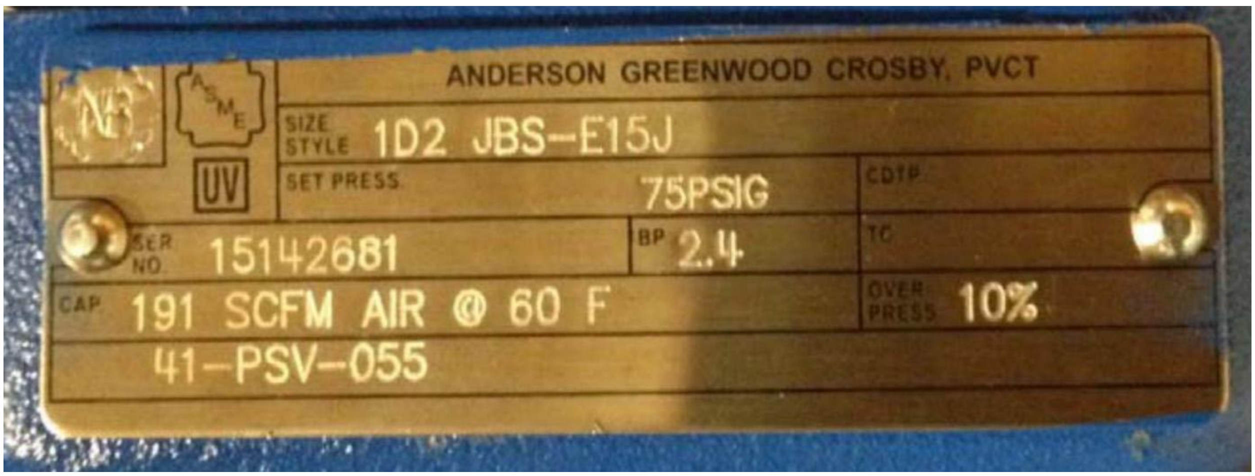


Figure 5: Mandatory Identification Tag

Such things as process-fluid quality, corrosion potential, fouling, and leakage must be taken into account when setting inspection frequency. Each company, each regulating body, and even each facility management have a say in determining inspection details and frequency. If a regulating body says “must be inspected every 5 years”, and Company Policy says “must be inspected every 3 years”, and field management says “must be inspected every year unless granted a documented exemption” then an annual inspection satisfies all three.

It is common on well sites for the answer to be “annually”, but there are many exceptions that lead to a shorter frequency. PSV in a field with corrosive gas will have a shorter frequency. PSV in a field with fluids subject to fouling (e.g., paraffin accumulation) will have a shorter frequency. PSV in vibrating service will often have a shorter frequency.

It is also common to track the results of the inspection of each device and if the device does not show signs of fouling, corrosion, or leakage, and the “as received” set point is equal to the “as reconditioned” set point for some number of iterations (typically three), then the inspection frequency can be extended. For example, a PSV received a clean bill of health for three annual inspections in a row, the facility allowed the inspection frequency to change to bi-annual (with proper documentation of the decision in the file). Six years later it has three more perfect reports, the facility allowed it to go to a three-year cycle. At the second three-year inspection the valve was found to have corrosion on the spring, the valve was replaced and the site went back to an annual inspection cycle on that valve. Record keeping is crucial to the success of that sort of plan.

For the most part, regulators don't care what your inspection frequency is, as long as the methodology is documented and followed, and records are kept proving the inspections were proper and done at the proper time. Often on well sites the field will go for years, even decades, without anyone ever pulling the records from the file. And then one day an incident that injures someone can be traced to an over-pressure event. At that point the musty files can easily be the difference between "normal wear and tear" and "criminal negligence".

It is beginning to become common for a site to institute "Risk-based Inspection (RBI) Assessments". This analysis is intended to evaluate the operating conditions of each pressure safety device and establish an inspection frequency for that device that is consistent with the conditions where it is operating. The benefits of RBI include:

a) systematic and well-developed technical methods and tools for evaluating pressure-relieving devices, which have a wide array of considerations;
b) focus on risk management by addressing critical concerns to protect equipment overpressure;
c) organized approach to improving pressure-relief device performance;
d) incorporation of operational history and inspection servicing records;
e) cost-effective risk mitigation task identification to address safety/health/environmental consequences and lessen economic loss.

# Flow rate determination

The methodology for calculating flow rate through a pressure safety device is dependent upon the process fluid that will be flowing through it. Single phase gas flow rate is a function of the speed of sound while single-phase liquid is dependent on the liquid specific gravity at flowing temperature and pressure. Multi-phase, flashing liquids, non-Newtonian fluids, and condensing gases are extremely complex, and there is not a generally accepted method for estimating the relieving rate of these fluids.

## Rate calculation for gas

All gas PSV are "critical flow devices" (i.e., they satisfy the requirements of Equation 1). As long as the downstream pressure ($P_2$) is less than or equal to the right side of Equation 1 the velocity (but not the mass flow rate) through the PSV will be constant.

$$\left(P_2 + P_{atm}\right) \le \left(P_1 + P_{atm}\right) \cdot \left(\frac{2}{k+1}\right)^{\frac{k}{k-1}} \qquad \text{Equation 1}$$

When the flow satisfies the critical pressure, the fluid velocity is equal to "sonic velocity" (Equation 2). Equation 2 units work out to distance/unit time (e.g., ft/s or m/s), and for a specific gas at a specific temperature, this calculation represents the speed of sound within the fluid described by the adiabatic constant and specific gravity. Notice that there is no pressure term in Equation 2, that is because while the downstream pressure is less than or equal to the critical pressure, changing either upstream or downstream pressure has no effect on the velocity (but it does change the mass flow rate).

$$v_{sonic} = \sqrt{k \cdot R_{gas} \cdot T_1} = \sqrt{k \cdot \frac{R_{air}}{\gamma} \cdot T_1} \qquad \text{Equation 2}$$

The requirement to maintain critical flow makes the exhaust piping length and size to be of vital importance. If the friction losses in the flare-header piping or the pressure drop across the flare nozzle causes the pressure on the downstream side of the PSV to exceed the pressure allowed by Equation 1 then the flow rate will be significantly less than the flow rate calculated below. Confirming the friction loss in the piping when each PSV is flowing alone is only the first step. You also must confirm that if all connected PSV lift concurrently, the flare header and the flare nozzle will allow enough flow to keep all the PSV in critical flow. This calculation is a major reason that too many well site flare headers are inadequate. It is often done properly in plants and rarely done adequately on well sites.

When the pressure of a gas is quickly reduced (e.g., through a pressure regulator), most gases will cool through a process known as "Joule-Thomson Cooling". This effect is upstream-pressure dependent and can result in gas temperatures dropping to very low values, which in turn changes the density of the gas, sometimes dramatically. For any credible scenario involving a safety valve downstream of a pressure regulator, you must estimate cooling effect of the pressure drop when calculating mass flow rate.

Mass flow rate is a function of the upstream (process) pressure and temperature, exhaust pressure (assumed to be less than Equation 1), sonic velocity, gas characteristics, and the size of the opening within the valve. The source of pressure may be significantly higher than the setpoint of the pressure safety valve (e.g., reservoir pressure is 10,000 psig and valve set point is 600 psig), but it must be assumed that the safety valve is sized to allow fluid friction to bleed off the bulk of the overpressure. When required flow rate is provided as volume flow rate in standard units (i.e., SCF, MSCF, MMSCF, kSCm, etc.), then multiply the flow rate times the density at standard conditions to get to a mass and then adjust the time units.

$$C_{units} = \frac{\sqrt{lbm \cdot R}}{lbf \cdot hr} \qquad \text{for SI} \qquad C_{units} = \frac{\sqrt{kg \cdot k}}{mm^2 \cdot kPa \cdot hr}$$

$$C = 520 \cdot C_{units} \cdot \sqrt{k \cdot \left(\frac{2}{k+1}\right)^{\frac{k+1}{k-1}}} \quad \text{for SI} \quad C = 0.03948 \cdot C_{units} \cdot \sqrt{k \cdot \left(\frac{2}{k+1}\right)^{\frac{k+1}{k-1}}} \qquad \text{Equation 3}$$

$$A_{orifice} = \frac{\dot{m}_{required}}{C \cdot P_1 \cdot K_d \cdot K_b \cdot K_c} \cdot \sqrt{\frac{T_1 \cdot Z_1}{MW}}$$

The constants are:

- $K_d$ → Discharge constant, from manufacturer (for valve selection use 0.975)
- $K_b$ → Backpressure constant, from manufacturer (for valve selection use 1.0)
- $K_C$ → Rupture disk constant (use 0.9 if a rupture disk is present under the PSV, 1.0 if absent)

Once you've calculated the area required, pick the next larger value in **Error! Reference source not found.** values ( the first number in the column headers is PSV inlet pipe size and the second number is the PSV outlet pipe size).

Table 2: PSV Available Orifices Sizes

| | Area ($in^2$) | Area ($cm^2$) | 1×2 | 2×2* | 2×3 | 3×3* | 3×4 | 4×6 | 6×8 | 6×10 | 8×10 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| D | 0.110 | 0.710 | X | | | | | | | | |
| E | 0.196 | 1.265 | X | X | | | | | | | |
| F | 0.307 | 1.981 | X | X | | | | | | | |
| G | 0.503 | 3.245 | | X | X | | | | | | |
| H | 0.785 | 5.065 | | X | X | | | | | | |
| J | 1.287 | 8.303 | | X | X | X | X | | | | |
| K | 1.838 | 11.858 | | | | X | X | | | | |
| L | 2.853 | 18.406 | | | | | X | X | | | |
| M | 3.600 | 23.226 | | | | | | X | | | |
| N | 4.340 | 28.000 | | | | | | X | | | |
| P | 6.380 | 41.161 | | | | | | X | | | |
| Q | 11.050 | 71.290 | | | | | | | X | | |
| R | 16.000 | 103.226 | | | | | | | X | X | |
| T | 26.000 | 167.742 | | | | | | | | | X |
| * Equal inlet/outlet combinations only available in threaded connections, flanged valves always have a larger outlet than inlet | | | | | | | | | | | |

With a selected PSV, you can calculate a valve capacity with the actual orifice area and actual valve-specific constants from the manufacturer using Equation 4.

$$\dot{m}_{design} = A_{orifice} \cdot C \cdot P_1 \cdot K_d \cdot K_b \cdot K_c \cdot \sqrt{\frac{MW}{T_1 \cdot Z_1}} \qquad \text{Equation 4}$$

## Rate calculation for non-flashing liquid

The calculations in this section assume that the liquid will not change state through the PSV. For volatile liquids that are near their flash point or might condense if quickly cooled, a process simulator is required to calculate relieving rates.

$$A_{design} = \frac{Q}{38 \cdot K_d \cdot K_b \cdot K_c \cdot K_v} \cdot \sqrt{\frac{\gamma_{liq}}{P_1 - P_2}}$$

$$N_{re} = \frac{Q \cdot \left(2800 \cdot \gamma_{liq}\right)}{\mu \cdot \sqrt{A_{orifice}}} \qquad \text{Equation 5}$$

$$K_V = \left(\frac{170}{N_{re}} + 1\right)^{-0.5} \quad \text{valid for } N_{re} \geq 80\text{, for lower flows } K_v = 1.0$$

The specific gravity of liquids is not an intrinsic property and must be evaluated at flowing temperatures.

## Process Slam Valves

With an effectively infinite source of gas (e.g., a long, high-pressure pipeline or a gas reservoir), a PSV will "never" lower the pressure enough to reseat and eventually something will come apart in a manner that is not easily controlled (e.g., the Joule-Thomson cooling effect can put the piping immediately upstream of the PSV into the brittle fracture temperature zone and the PSV can physically break off the piping). The credible scenario analysis should show this risk. When the credible over-pressure source is effectively infinite (i.e., the design flow rate of the pressure safety device will not reduce the overpressure before the pressure in the line exceeds 110% of MAWP), then it is common to require a valve in the process line to slam shut at some pressure below 110% of MAWP. These slam valves isolate the protected piping from the credible pressure source and cannot be reopened without a manual intervention. These valves go by several names including High/Low Operation Valves, Emergency Shutdown (ESD) valves, XV, etc. They must be locally controlled (i.e., not controlled by an on-site process logic controller), and must require manual intervention to reopen. In conjunction with proper pressure safety devices, Slam Valves can prevent excessive waste and reduce risk.

## Exhaust forces

If a rocket can lift a mass into space, it stands to reason that a jet exiting a pipe at sonic velocity will exert some measurable force on the piping. For a system exhausting gas, that force is:

$$F_g = \left( \frac{K_D \cdot A_{orifice} \cdot \left(P_1 + P_{atm}\right) \cdot K_f}{1.383 \cdot A_{outlet}} - P_{atm} \right) \cdot A_{outlet}$$

$$\text{If } F_g < 0 \Rightarrow F_g = 0 \qquad \text{Equation 6}$$

$$F_r = \left( \frac{C}{332.7 \cdot C_{units}} \right) \cdot K_D \cdot A_{orifice} \cdot \left(P_1 + P_{atm}\right) \cdot \sqrt{\frac{k}{k+1}} + F_g$$

The new constant ($K_f$) is a function of the adiabatic constant of the process gas (see **Error! Reference source not f ound.**).

The results of this calculation can predict a considerable force that must be countered with adequate support. It is common for the (unsupported) tail pipe on threaded PSV to unscrew from the valve and create a flying-debris hazard. Flanged PSV don't unscrew, but they can experience significant bending forces (**Error! Reference source not found.**).

## Example

A processed natural gas pipeline is approaching a population center and code requires over pressure protection in that vicinity. The conditions are:

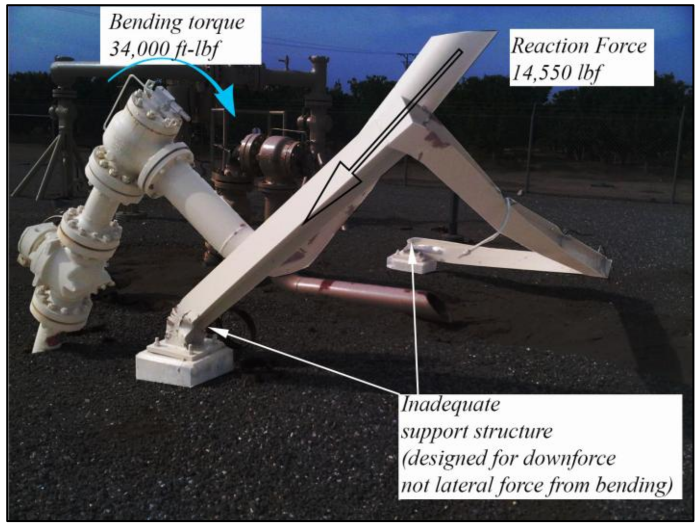
- PSV set pressure → 1550 psig
- PSV description → Pilot operated, 4X6 with "P" orifice (flow area 6.380 in$^2$)
- PSV outlet piping → Heavy wall 6-inch (ID 5.761 in$^2$)
- Site elevation → 2400 ft (atmospheric pressure 13.5 psia)
- Gas → Pipeline quality natural gas (specific gravity 0.5539, adiabatic constant 1.31)
- Distance between the centerlines of two vertical pipes → 2.3367ft

- Table 1:API 520-2 Reactive force table 05-A1

Table 2:API 520-2 Reactive force table 05-A1

| $k$ | $K_f$ | | $k$ | $K_f$ |
|---|---|---|---|---|
| 1.01 | 1.15 | | 1.55 | 0.95 |
| 1.05 | 1.13 | | 1.6 | 0.94 |
| 1.1 | 1.11 | | 1.65 | 0.93 |
| 1.15 | 1.09 | | 1.7 | 0.91 |
| 1.2 | 1.07 | | 1.75 | 0.9 |
| 1.25 | 1.05 | | 1.8 | 0.89 |
| 1.3 | 1.03 | | 1.85 | 0.87 |
| 1.35 | 1.02 | | 1.9 | 0.86 |
| 1.40 (air) | 1 | | 1.95 | 0.85 |
| 1.45 | 0.98 | | 2 | 0.84 |
| 1.5 | 0.97 | | | |

Figure 6: PSV Reaction Forces



## Block Valves

The PSV in **Error! Reference source not found.** has a block valve between it and the main process flow. Those valves are very controversial. On the one hand, having the valve in that position allows technicians to remove the valve without having to depressurize an extensive, high-pressure gas line. On the other hand, if the valve ever gets closed then the system being protected by the valve is at risk. There is not a fantastic answer to this decision. Code, regulations, and company policies often allow them on the condition that means be provided to both prevent their unauthorized operation and that they are inspected on a frequent (and documented) schedule to ensure that the block valves have not been repositioned and that the locking mechanism has not been tampered with.

There are some installations where a side- valve has been installed between the block valve and the PSV to allow the PSV to have its set-point tested without removing it from the line. This does not take the place of a physical inspection, but in many cases, it will allow the physical inspection frequency to be extended while still doing an annual certification of set point.

## Conclusion

Too often, field engineers have simply relied on vessel fabricators to specify pressure safety devices and don't check their work. This can lead to some very substandard results. If you haven't provided a credible-scenario analysis, then what volume are they using to size the valves? They typically won't provide the rationale behind their selection and sizing of the valves/rupture disks. If you ask for it, most commonly they will provide a graph of flow rate that was copied from the valve-manufacturer's brochure. At that point no one in the whole wide world knows if the flow rate is too high, too low, perfect, or good enough. Having investigated several incidents related to vessel over pressure accidents, I can assure you that if the accident investigation team is qualified to participate in the investigation, then their very first question will be "can you provide the credible scenario analysis?". The rest of the investigation has a very different tone if the answer is "Of course, here it is" as opposed to "What is that?" Without knowing what is credible, it is not possible to determine if the over-pressure protection is adequate.

## Nomenclature

| Variable | Description | FPS Units | SI Units |
|---|---|---|---|
| $A_{orifice}$ | Flow area of PSV orifice | $in^2$ | $cm^2$ |
| $A_{outlet}$ | Flow area of PSV discharge pipe | $in^2$ | $cm^2$ |
| $C$ | Convenient accumulation of parameters in Equation 3Equation 1 | $\frac{\sqrt{lbm \cdot R}}{lbf \cdot hr}$ | $\frac{\sqrt{kg \cdot K}}{mm^2 \cdot hr \cdot kPa}$ |
| $F_g$ | Force at outlet of exhaust pipe | lbf | N |
| $F_r$ | Reactive force on PSV | lbf | N |
| $k$ | Adiabatic “constant” | None | None |
| $K_d$ | Discharge constant | None | None |
| $K_b$ | Backpressure constant | None | None |
| $K_C$ | Rupture Disk Constant | None | None |
| $K_f$ | Reactive force constant | None | None |
| $K_v$ | Viscosity constant | None | None |
| $\dot{m}_{\text{required}}$ | Required mass flow rate at actual conditions | $lbm/ft^3$ | $kg/m^3$ |
| $\dot{m}_{\text{design}}$ | Design mass flow rate at actual conditions | $lbm/ft^3$ | $kg/m^3$ |
| $MW$ | Molecular weight | lbm/lb-mole | gm/gm-mole |
| $P_{\text{atm}}$ | Local atmospheric pressure | psia | kPaa |
| $P_1$ | Pressure upstream of some pressure drop | psig | kPag |
| $P_2$ | Pressure downstream of some pressure drop | psig | psig |
| $P_{\text{std}}$ | Pressure designated by proper authority to represent the standard pressure to be used for aggregating volumes | psia | kPaa |
| $Q$ | Liquid volume flow rate | gpm | Lpm |
| $R_{air}$ | Specific gas constant for air | 53.355 ft*lbf/R/lbm | 287.068 J/K/kg |
| $R_{gas}$ | Specific gas constant for any specified gas | $R_{air}/\gamma_{gas}$ | $R_{air}/\gamma_{gas}$ |
| $T_1$ | Temperature of gas upstream of a pressure drop | R | K |

| **Variable** | **Description** | **FPS Units** | **SI Units** |
|---|---|---|---|
| $T_{\mathrm{std}}$ | Temperature designated by proper authority to represent the standard pressure to be used for aggregating volumes | R | K |
| $v_{sonic}$ | Sonic velocity | ft/s | m/s |
| $Z_1$ | Compressibility upstream of some pressure drop | None | None |
| $\gamma$ | Process gas specific gravity (relative to air = 1.0) | None | None |
| $\mu$ | Dynamic viscosity | cP | cP |