PRV sizing for exchanger tube rupture

A comprehensive step-by-step approach to calculating required pressure relief valve (PRV) load

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Heat exchanger tube rupture is one of the common overpressure scenarios of pressure relief valve (PRV) design. ASME Code VIII-1[1] indicates that heat exchangers shall be protected with a relieving device of sufficient capacity to avoid overpressure in case of an internal failure. But it does not provide any guidance on how to size a PRV and how to define the required relief load. API RP 520 Part I[2] and API RP 521[3] do offer some guidelines for heat exchanger tube rupture in PRV design, but they are too general to be used to perform a detailed calculation or to do a relief system analysis. Furthermore, API 520's two-phase flow calculation concepts contradict its own assumptions,[6] which makes its design approach infeasible. A detailed PRV design procedure for heat exchanger tube rupture will be discussed.

The validity of a heat exchanger tube rupture case. API RP 521, Sec. 3.15.2[3] recommends that complete tube failure be considered a viable contingency when the design pressure of the low pressure side is less than two-thirds of the design pressure of the high pressure side. Here the design pressure is equal to or less than maximum allowable working pressure (MAWP).

Basically this standard has been widely accepted by the oil and chemical industries and proven effective at quantifying the requirements of PRVs for heat exchanger tube rupture. It is a practical and conservative standard.

One should avoid the tendency to use operating pressure instead of design pressure (MAWP) for the high pressure side in the two-thirds rule recommended by API, even though the modification does create certain room for eliminating some originally required PRVs, such as the following heat exchanger case:

Low pressure (L.P.) side: design pressure (D.P.) = 180 psig
High pressure (H.P.) side: design pressure (D.P.) = 300 psig
operating pressure = 250 psig

According to API's standard:

D.P. of L.P./D.P. of H.P. = 180/300 = 0.6 < 2/3 and tube rupture is a valid case.

But based on the modified standard:

D.P. of L.P./O.P. of H.P. = 180/250 = 0.72 > 2/3 and tube rupture can be ignored.

Quite obviously, it is an economical design by using operating pressure to replace the design pressure to manipulate the design calculation. However, the danger in most refinery or chemical processes is that the operating pressure of a process may change. It may rise to an unexpected, undesirable level due to process upset, environmental factors or human errors. Where the low pressure side of a heat exchanger is not fully protected when a tube rupture does occur, the potential consequences may never be offset by the savings gained from the economical design.

From a mechanical viewpoint, it is evident that the tube and shell design pressures are reliable data for measuring the reliability of a heat exchanger. Once a heat exchanger is manufactured, its reliability is measurable and fixed by its design pressures and temperatures. From that viewpoint, API's standard is logical.

It appears that using a variable (the operating pressure of a process) to judge a predefined datum (the strength of equipment) is not rational. The modified API guideline should not be used unless a HAZOP of the process is conducted, indicating that overpressure would never occur in the system. Undoubtedly, the logical API standard should be strictly followed for quantifying the overpressure protection.

Flow patterns of fluid flowing across a tube rupture. For possible heat exchanger tube rupture, there are basically four different scenarios of fluid in the high pressure side flowing through a sharp break to the low pressure side:

- Vapor flowing through break without phase change
- Liquid flowing through break without phase change
- Liquid flowing through break with phase change
- Two-phase (vapor and liquid) flowing through break.

Assumptions. A sophisticated flow model is not necessary, but some assumptions must be made for simplifying the calculation of a fluid flowing across a tube break:

a.) A tube rupture is considered as a sharp break in only one tube, with the high pressure fluid flowing through both sides of the break.

b.) Each side of the break is treated as a sharp-edged orifice having the cross-sectional area of a tube. That means the rupture opening equals twice the cross-sectional area of one tube.

c.) The fluid flowing through a sharp-edged orifice is an isenthalpic (adiabatic) expansion.

d.) The incremental flashing across the tube break is ignored.

e.) Two-phase fluid in either side is treated as a homogeneous mixed-phase fluid, which means that the phase slip is negligible.

f.) The effect of auto-refrigeration arising from the flashing of the fluid is not included. If it is a significant effect for some specific cases, special considerations must be included.

Based on the above assumptions, Crane[4] formulas for fluid flowing through orifices can be applied to calculate the flowrate across a tube rupture.

The analysis of pressure profile in tube rupture flow mechanism. In case of tube rupture in a heat exchanger, the process fluid will flow from the high pressure side across the tube rupture into the low pressure side. If the low pressure side cannot absorb the flowrate coming from the high pressure side at 10% accumulation overpressure of the design pressure at the low pressure side of the heat exchanger, the surplus flowrate should be relieved from the low pressure side across a PRV orifice into a flare header, a vessel or atmosphere.

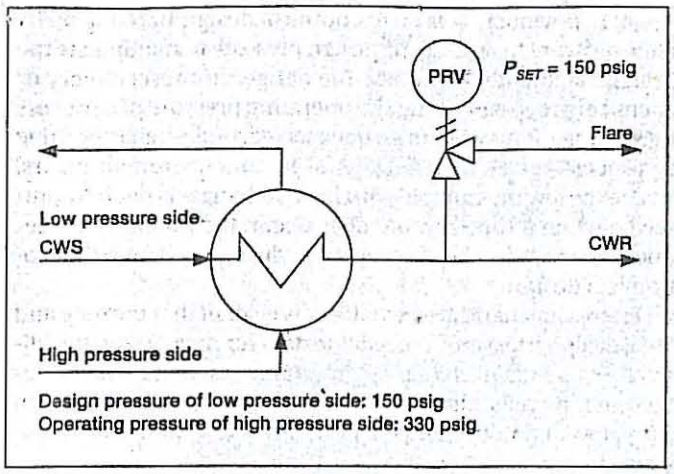

Fig. 1—Process flow diagram.

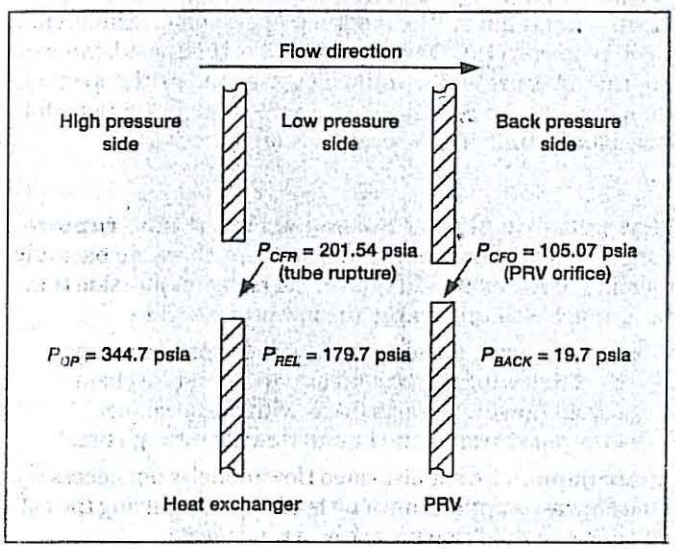

Fig. 2—Pressure profile of tube rupture.

The whole flow mechanism can be handled as two pressure relief processes in a series as shown in Figs. 1 and 2.

The pressure profile for the two relief processes is illustrated in the following example:

1. Pressure relief through tube rupture. When a tube ruptures, the process fluid at 330 psig or 344.7 psia will flow across the tube rupture into the low pressure side. With a PRV installed at the low pressure side with a relief pressure of 110% of its design pressure, i.e., P_{REL} = 150 psig (110%)+ 14.7 psi = 179.7 psia, the low pressure side will be under 179.7 psia. But the flow may be controlled by the critical flow pressure at the tube rupture for vapor phase, if the critical flow pressure is higher than the relieving pressure at the low pressure side.

The critical flow pressure at the tube rupture can be expressed as the following, as recommended by API 520 Eq. 1:

$$P_{CFR} = P_1\left(\frac{2}{K+1}\right)^{K/(K-1)} \qquad (1)$$

where

P_1 = Operating pressure at the high pressure side
= 344.7 psia

K = 1.1, ratio of specific heats (from simulation results)

$$P_{CFR} = 344.7\left(\frac{2}{2.1}\right)^{1.1/(1.1-1)}$$

= 201.54 psia

Since $P_{CFR} > P_{REL}$, the pressure difference of $dp = P_1 - P_{CFR}$ will be the driving force of process fluid flowing across the tube rupture.[6]

2. Pressure relief through PRV. The second pressure relief is the process fluid coming from the high pressure side, at the low pressure side flowing across the PRV orifice into the back pressure side.

Its flow mechanism is similar to the pressure relief through tube rupture discussed previously. Again, the critical flow condition at the orifice must be checked for vapor phase. Eq. 1 can be rewritten as:

$$P_{CFO} = P_{REL}\left(\frac{2}{K+1}\right)^{K/(K-1)} \qquad (2)$$

where

P_{REL} = Relieving pressure at the low pressure side
= 150 psig (110%) + 14.7 psi
= 179.7 psia

K = 1.1, rate of specific heats (from computer simulation results)

$$P_{CFO} = 179.7\left(\frac{2}{2.1}\right)^{1.1/(1.1-1)}$$

= 105.07 psia

Since $P_{CFO} > P_{BACK}$, the pressure difference, $dp = P_{REL} - P_{CFO}$, will be the driving force of process fluid flowing across the PRV orifice.[6]

Defining the relieving flowrates for tube rupture cases.

Vapor flowing through tube break without phase change. Crane Eq. 3-22[4] can be used for compressible fluids flowing through a sharp-edged orifice based on the assumption mentioned previously.

$$W_V = 1{,}891\, Y d^2\, C\,(dP/V)^{0.5}$$

Only one cross-sectional area included.

where $Y = 1 - 0.317\, dP/P_1$

The equation is summarized from the figures on page A-21 of Crane[4] for simplifying the lengthy calculations.

C = 0.6, flow coefficient for square-edged orifices (from Crane figures on page A-20)

$d^2 = 3.1416\, A_V/4$

It should be noted that critical pressure must be checked since under critical flow condition the actual pressure at a rupture opening (throat) cannot fall below the critical flow pressure even if a much lower pressure exists downstream.[6]

Incorrect use of the pressure at the low pressure side may cause oversizing the PRV. This common mistake should be avoided.

Combining all the previously given variables, the previous equation can be rewritten as:

$$W_V = 1{,}444.6\, A_V\,(1 - 0.317\, dP/P_1)\,(dP/LO_V)^{0.5} \qquad (3)$$

Liquid flowing through tube break without phase change. Critical condition does not need to be considered for liquid flowing through the tube break without phase change. Crane Eq. 3-21[4] can be used for liquid flowing through a sharp-edged orifice:

$$W_L = 1{,}891\, d^2\, C\,(dP\, LO_L)^{0.5}$$

Only one cross-sectional area included.

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where $C = 0.6$ (from Crane figures on page A-20)
$d^2 = 3.1416\, A_L/4$

Combining all the above given variables Eq. 4 can be rewritten as:

$$W_L = 1{,}444.6\, A_L\, (dP\, LO_L)^{0.5} \quad (4)$$

Liquid flowing through tube break with phase change. A simulation may be required to define the vapor ratio at the vena contracta by an isenthalpic (adiabatic) expansion from the relieving condition either to the downstream critical pressure of the flash vapor or to the downstream relieving pressure, whichever is greater.[6] The detailed calculation procedures are similar to the ones discussed next.

Two-phase flowing through tube break. Design basis:

- Vapor flow and liquid flow of a two-phase fluid to pass a tube breakage opening area are calculated individually.
- The total individual areas equal twice the cross-sectional area of one tube.

Calculate tube breakage flows. Vapor flowrate from Eq. 3:

$$W_V = 1{,}444.6\, A_V\, (1 - 0.317\, dP/P_1)(dP/LO_V)^{0.5} \quad (3)$$

where dP should be applied.

Liquid flowrate from Eq. 4:

$$W_L = 1{,}444.6\, A_L\, (dP\, LO_L)^{0.5} \quad (4)$$

where dP should be applied.

Total breakage opening area (assumption b):

$$A_{TOTAL} = A_L + A_V = 2\,(3.1416 d^2)/4 \quad (5)$$

Vapor ratio can be written as:

$$R = W_V/(W_V + W_L) \quad (6)$$

Where R can be obtained by the flash simulation of the bulk fluid at the tube break. An important point is that the flash pressure should be selected between the downstream relieving pressure and critical pressure, whichever is greater.[6]

The liquid and vapor flowrates, W_L and W_V, can be calculated by solving Eqs. 1 to 6 simultaneously. The examples shown later will illustrate how to calculate both the flowrates.

Notes. One important point is that the pressure used for calculating W_L, W_V and R should be consistent.

Currently, as recommended by API 520,[2] most oil companies adopt the following approaches for sizing a PRV at two-phase flow condition: while its vapor flowrate is calculated under the critical pressure condition, its liquid flowrate is calculated under the downstream relieving condition. They are inconsistent.

Obviously, the vapor flow and liquid flow at a tube break must be controlled by only one pressure, either the critical pressure at the throat or the downstream relieving pressure. As mentioned, the controlling pressure should be the greater of the two.

The second point is pertinent to the vapor/liquid ratio across the tube break. Some design engineers assume that the vapor/liquid ratio is constant across the tube break as recommended by many company manuals. In fact, the vapor/liquid ratio downstream is always higher than at the upstream high pressure side across the tube break, since more liquid should flash into vapor when pressure drops.

This procedure might be an acceptable simplified approach for the same vapor/liquid ratio crossing a break when the pressure drop across the break is small, but it seldom is the case as a heat exchanger tube ruptures. If a heat exchanger tube rupture is valid according to API 521's two-thirds rule, the vapor/liquid ratio of the bulk fluid would never remain unchanged after it flashed into the low pressure side from the high pressure side. A simple simulation can show the difference.

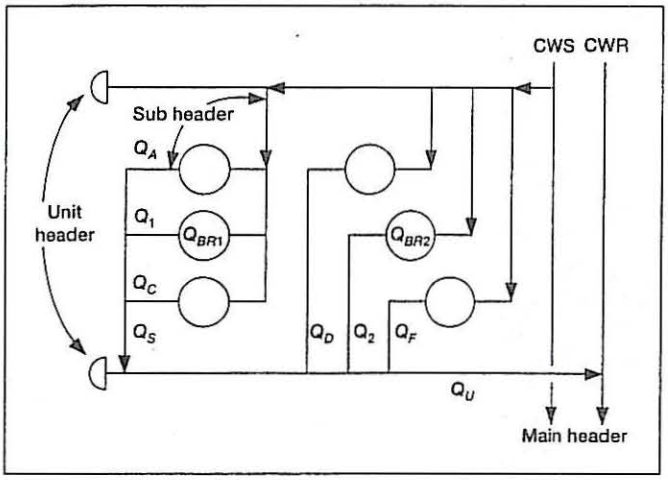

Fig. 3—Cooling water network.

Secondary effect of tube rupture. Once a tube is ruptured, the high pressure side fluid flows into the low pressure side. Flow of the low pressure side may be stopped by the pressure rise in the system. At the same time the function provided by the low pressure side system may also stop. From Fig. 3 one can see that when Q_{BR1} displaced Q_1, the high pressure at the return lines of the subheader might cause Q_A and Q_C to stop. If Q_{BR1} is big enough, even unit header services could be affected. Thus, if the low pressure system is for example cooling water, a tube rupture may induce a loss of cooling water relief, possibly involving several other services. Sometimes the consequence may be very serious. But, no matter how serious this could be, this is a secondary effect which should be handled as a consequence rather than a double contingency.

If a heat exchanger is part of a preheat network, the secondary effect could be very complicated and should be treated with extreme caution.

Minimum protection — thermal relief valve. Quite often, a lengthy calculation may conclude that the low pressure side is capable of absorbing the total process fluid flowrate coming from the high pressure side caused by a tube rupture. Does this mean that no pressure relief valve is required? Some think yes, others no. It is my understanding that a thermal relief valve is a good economical investment which should be installed on the low pressure side of a heat exchanger, even though calculations may show that the low pressure side provides more displacement credits than required relief capacity. The reason is simple. Once a tube rupture is detected, the heat exchanger may risk thermal expansion when all the inlet and outlet block valves of both the high pressure and the low pressure sides of the heat exchanger are closed. Any leakage through the high pressure side block valves can cause the pressure at the low pressure side to rise to that of the high pressure side, which might result in major failure of the heat exchanger.

For more information regarding thermal relief valves please see reference 5.

Volumetric capacity credit. Often the low pressure side is capable of absorbing the high pressure side flow across a tube rupture.

Example 1. Liquid fluid flashing into two-phase relief is controlled by critical flow condition.

A heat exchanger operates with hydrocarbon fluid on the

high pressure side and cooling water on the low pressure side (Fig. 1). The questions are whether a PRV is required at the low pressure side and if it is required, what is the relieving capacity and how is it sized?

1. Check the validity of tube rupture. Based on API RP 521 Sec. 3.15.2 two-thirds rule for the heat exchanger, the design pressure ratio of the low pressure side versus the high pressure side is as follows:

Pressure ratio = 150/330 = 0.45 < 2/3

Therefore, tube rupture is a valid case.

2. Check the fluid phase status after flash. From computer simulation, the results show that at 150 psig (110%) + 14.7 = 179.7 psia, the fluid flashes into two phases: streams V2 and L2. This is a case of liquid flashing into two-phase case.

3. Check critical flow condition. From Eq. 1:

$$P_{CFR} = P_1\left(\frac{2}{K+1}\right)^{K/(K-1)}$$

Where $K = 1.1$ (based on stream V2: $MW = 48.1$ and $T = 134.1°F$ from the simulation results).

$P_1 = 330$ psig + 14.7
$= 344.7$ psia

Critical flow pressure:

$$P_{CFR} = 344.7\left(\frac{2}{1.1+1}\right)^{1.1/(1.1-1)}$$

$= 201.54$ psia

Relieving pressure:

$P_{REL} = 150$ psig (110%) + 14.7 = 179.7 psia (10% accumulation)

Comparing the critical flow pressure and the relieving pressure, since $P_{CF} > P_{REL}$, the relieving is controlled by the critical flow condition.

4. Find the vapor ratio of two-phase fluid crossing the tube rupture. Calculations are based on the simulation results of fluid flashing at the critical flow conditions.

From simulation:

Vapor stream, VC: $W_{VC} = 266$ lb/hr
$LO_{VC} = 1.91$ lb/ft^3

Liquid stream, LC: $W_{LC} = 734$ lb/hr
$LO_{LC} = 29.88$ lb/ft^3

$R = W_{VC}/(W_{VC} + W_{LC}) = 266/(266 + 734) = 26.6\%$

5. Find the required relieving capacities. From Eq. 3:

$$W_V = 1{,}444.6 A_V\left(1 - 0.317\frac{dP}{P_1}\right)(dP\ LO_V)^{0.5}$$

where $dP = P_1 - P_{CF}$
$= 344.7 - 201.54$
$= 143.16$ psi

$$W_V = 1{,}444.6 A_V\left(1 - 0.317\frac{143.16}{344.7}\right)(143.16)^{0.5}(1.91)^{0.5}$$

$= 20{,}742.8\ A_V$ (E-1)

From Eq. 4:

$W_L = 1{,}444.6\ A_L\ (dP\ LO_L)^{0.5}$
$= 1{,}444.6\ A_L\ (143.16)^{0.5}\ (29.88)^{0.5}$
$= 94{,}481.9\ A_L$ (E-2)

Exchanger tubes are 3/4 in. OD 12 BWG, 0.532 in. ID.

From Eq. 5:

$A_{TOTAL} = A_V + A_L = 2\ (3.1416\ d^2)/4$ (E-3)

Thus $A_V + A_L = 2\ (3.1416)\ (0.532^2)/4 = 0.4446$ in.^{2} (E-3)

From simulation, the above vapor ratio is 26.6%. That means: $W_V/(W_V + W_L) = 26.6\%$ (E-4)

Equating Eqs. E-1 through E-4 and solving:

$W_V = 20{,}742.8\ A_V$ (E-1)
$W_L = 94{,}481.9\ A_L$ (E-2)
$A_V + A_L = 0.4446$ (E-3)
$R = W_V/(W_V + W_L) = 26.6\%$ (E-4)

The individual phase flowrates can be easily obtained:

$W_V = 5{,}744$ lb/hr or
$Q_V = 5{,}744$ lb/hr/1.91 lb/ft^3 = 3,007 ft^3/hr
$W_L = 15{,}847$ lb/hr or
$Q_L = 15{,}847$ lb/hr/29.88 lb/ft^3 = 530 ft^3/hr

6. Find the actual required relief loads for tube rupture. First, calculate the actual volumetric ratio at critical flow condition, from simulation:

$Q_{VC} = 139$ ft^3/hr (stream VC)
$Q_{LC} = 24.573$ ft^3/hr (stream LC)

Ratio = $Q_{VC}/(Q_{VC} + Q_{LC}) = 139/(139 + 24.573) = 85\%$

Assume the available volumetric capacity credit is 200 ft^3/hr the available vapor phase volumetric capacity credit is:

200 ft^3/hr (85%) = 170 ft^3/hr

The available liquid phase volumetric capacity credit is:

200 − 170 = 30 ft^3/hr

The actual required relief load for vapor is:

3,007 − 170 = 2,837 ft^3/hr or

2,837 ft^3/hr (1.91 lb/ft^3) = 5,418.7 lb/hr

The actual required relief load for liquid is:
530 − 30 = 500 ft^3/hr or
500 ft^3/hr (7.48052 gal/ft^3)/60 min/hr = 62.34 gpm

7. Sizing PRV. Check critical flow condition at relief valve orifice from Eq. 2, where:

$P_1 = P_{REL} = P_{SET}$ (110%) = 150 psig (110%) + 14.7
$= 179.7$ psia
$M = 48.1$ (stream V2)
$K = 1.1$

$$P_{CFO} = 179.7\left(\frac{2}{2.1}\right)^{1.1/(1.1-1)} = 105.07 \text{ psia}$$

Since $P_{CF} > P_{BACK}$, the flow is controlled by critical flow condition.

Calculate required relieving area for vapor phase at critical flow condition:

$$A_V = \frac{W_V}{C_o K_d P_1 K_b}\left(\frac{TZ}{M}\right)^{0.5} \quad \text{[API RP 520 I, 5 Ed., Eq. 2]}$$

where

$W = 5{,}418.7$ lb/hr
$P_1 = 179.7$ psia
$T = 97°F = 557°R$ (stream VC1)
$M = 48.19$ (stream VC1)
$Z = 0.8678$ (stream VC1)
$C_o = 327$
$K_d = 0.975$
K_b = Capacity correction factor due to back pressure

Say, $P_{BACK} = 5$ psig (back pressure at flare)
$P_{BACK}/P_{SET} = 5/150 = 3.33\%$

Continued

Therefore $K_b = 1.0$

$$A_V = \frac{5{,}418.7}{327(0.975)(179.7)(1.0)}\left(\frac{557(0.8678)}{48.19}\right)^{0.5}$$

$$= 0.2995 \text{ in.}^2$$

Calculate the required relief area for liquid phase at critical flow condition:

$$A_L = \frac{Q}{38 K_d K_W K_V}\left(\frac{G}{P_1 - P_2}\right)^{0.5} \quad [\text{API RP 520 I, 5 Ed., Eq. 9}]$$

where

Q = 62.34 gpm
K_d = 0.65
K_W = 1
K_V = 1
G = 0.5527 (stream VL1)
P_1 = 179.7 psia
P_2 = P_{CFO} = 105.07 psia

$$A_L = \frac{62.34}{38(0.65)(1)1}\left(\frac{0.5527}{179.7 - 105.07}\right)^{0.5}$$

$$= 0.2172 \text{ in.}^2$$

The total required relief area:

$A_{TOTAL} = A_V + A_L = 0.2995 + 0.2172 = 0.5167$ in.^{2}

An "H" type orifice with 0.785 in.^{2} is required.

Example 2. Liquid flashing into two-phase relief is controlled by the set pressure of a PRV. From Example 1, if the shell side design pressure is 180 psig, other conditions remain unchanged:

1. Validity of tube rupture case. Pressure ratio = 180/400 = 0.45 < 2/3. Therefore, tube rupture is still valid.

2. Check the fluid phase status after flash. From simulation, the liquid hydrocarbon flashes into V1 and L1 product streams. Thus, this is also a liquid flashing into two-phase case.

3. Check critical flow condition.

$$P_{CF} = P_1\left(\frac{2}{K+1}\right)^{K/(K-1)}$$

Where $K = 1.1$ (based on stream L1: $M = 50.9$, $T = 147.0$ °F)
$P_1 = P_{OP} = 344.7$ psia

Thus, critical flow pressure:

$P_{CFR} = 201.54$ psia

Relieving pressure:

$P_{REL} = 180$ psig (110%) (14.7 psi) = 212.7 psia

Since $P_{REL} > P_{CF}$, the downstream relieving conditions should be controlled by P_{REL}.

Steps 4 to 7 are similar to the ones of Example 1, but the driving force, $dP = P_1 - P_{REL} = 344.7 - 212.7 = 132$ psi should be used in all the related equations.

Example 3. The example of two-phase fluid from the high pressure side flowing across a tube rupture into the low pressure side is not included here. The detailed procedures are similar to Examples 1 and 2. The critical point is this: never assume the vapor ratio across a tube rupture or a PRV orifice remains unchanged. Always do a computer simulation for the process fluid to find out the actual vapor ratio after the process fluid flashes across a tube rupture or PRV orifice. When sizing a PRV at tube rupture, the actual vapor ratio should be applied.

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NOMENCLATURE

A = required effective discharge area of the valve, in.^{2}
A_{TOTAL} = required total effective discharge area of the valve for liquid and vapor phases, in.^{2}
C = flow coefficient for square-edged orifices from Crane A-20.
C_o = coefficient determined from an expression of the ratio of the specific heats of the vapor at standard conditions. This can be obtained from API RP 520 I, 5 Edition, Table 9.
d = tube inside diameter of a heat exchanger, in.
dp = pressure difference, psi.
G = specific gravity of the liquid at the flowing temperature referred to water = 1.0 at 70°F.
k = ratio of the specific heats.
K_b = capacity correction factor due to back pressure. This can be obtained from the manufacturer's literature or estimated from Fig. 27 of API RP 520-I, 5 Edition.
K_d = effective coefficient of discharge.
= 0.975 for vapor phase and 0.65 for liquid phase as recommended by API RP 520 I, 5 Edition.
K_v = correction factor due to viscosity as determined from Fig. 32 of API RP 520 I, 5 Edition.
K_w = correction factor due to back pressure.
= 1.0 if the back pressure is atmospheric. Balanced bellows valves in back pressure service will require the correction factor determined in Fig. 31 of API RP 520 I, 5 Edition.
LO = density, lb/ft^3.
M = molecular weight of the vapor.
P_1 = upstream relieving pressure, psia.
P_{BACK} = back pressure, psia.
P_{CF} = critical flow throat pressure, psia.
P_{OP} = operating pressure, psia.
P_{SET} = set pressure of a PRV, psia.
P_{REL} = relieving pressure of a PRV, psia. This is the set pressure plus the allowable overpressure plus atmospheric pressure.
Q = flowrate, gpm.
R = vapor ratio.
T = relieving temperature of the inlet gas or vapor, °R (°F + 460).
W = flowrate, lb/hr.
Y = net expansion factor for compressible fluid flowing through orifice.
Z = compressibility factor for the deviation of the actual gas from a perfect gas, a factor evaluated at relieving inlet conditions.

Subscripts

L = liquid phase
V = vapor phase
C = critical flow condition
VC = vapor phase at critical flow condition
LC = liquid phase at critical flow condition
CFO = critical flow condition at PRV orifice
CFR = critical flow condition at tube rupture

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