A Critical Review of ISO 4126-10

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G. A. Melhem, Ph.D., FAIChE

IOMOSAIC® CORPORATION

A Critical Review of ISO 4126-10

Process Safety and Risk Management Practices

authored by
G. A. Melhem, Ph.D., FAIChE

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1 Introduction

This paper provides a critical review of the second edition of ISO [1] 4126-10, an international standard titled "Safety devices for protection against excessive pressure". Part 10 of ISO 4126 focuses on "Sizing of safety valves and bursting discs for gas/liquid twophase flow". The second edition of ISO 4126-10 [1] was published in February of 2024. The first edition of ISO 4126-10 was published in October of 2010 [2]. This review is primarily focused on items that significantly deviate from other recognized and generally accepted good engineering practices (RAGAGEP).

2 Applicability

A single parameter ω method is used to calculate the relief flow requirements for different types of applicable scenarios. Both the first and second editions of ISO 4126-10 place strict limitations on the applicability of the single parameter ω method [3, 4, 5] (section 5) used to establish the relief flow requirements [6, 7]. These limitations are summarized in Table 1.

ISO 4126-10 indicates that the use of the sizing method outside the designated limits can lead to "unacceptable errors" and/or in general the "oversizing" of the relief device. In many cases, oversizing a pressure relief valve is just as "unacceptable" because oversizing can induce twophase flow, add more safe discharge location challenges, and/or lead to destructive chatter.

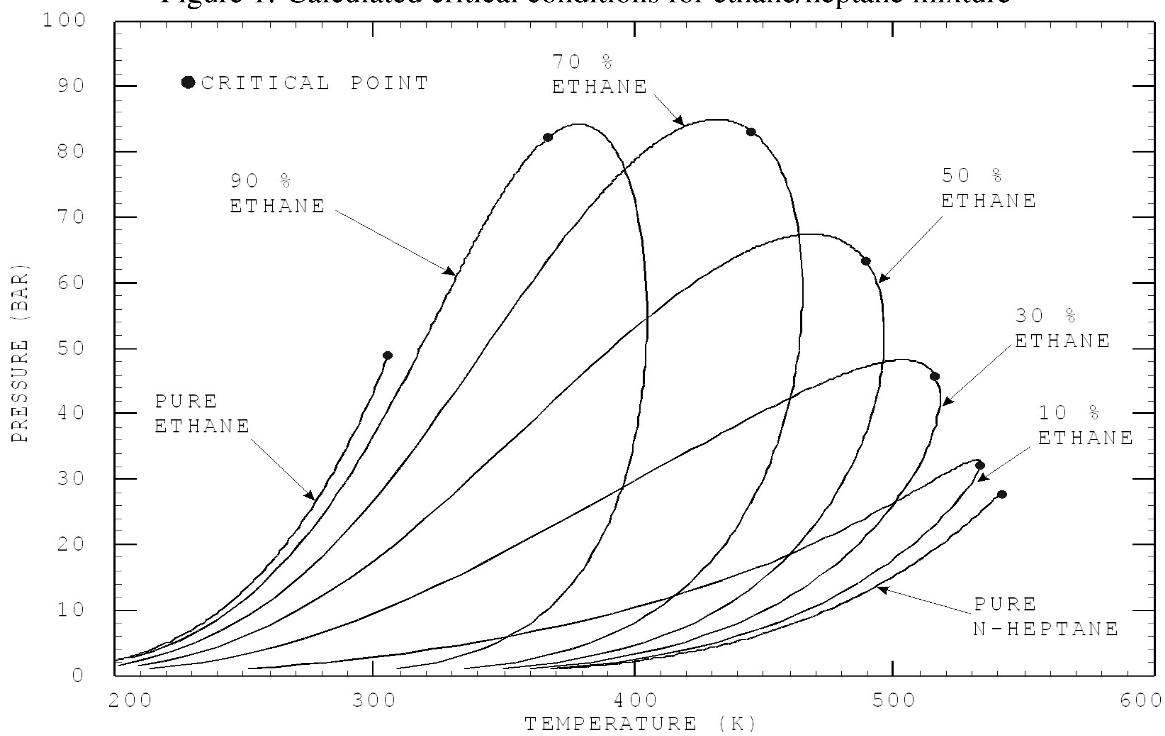
We note that critical properties for mixtures are not simply linear averages of the individual mixture components critical properties (see Figure 1). ISO 4126-10 does not provide a method for the calculation of mixture critical properties. An equation of state is required to generate mixture critical properties. If an equation of state is needed to generate critical mixture properties then the same equation of state can easily be used to generate suitable properties data sets for direct numerical vdP integration [6, 7].

ISO 4126-10 methods also require the evaluation of mixture latent heat of vaporization, mixture surface tension, mixture liquid heat capacity, mixture liquid and vapor mass densities, mixture bubble point or saturation temperatures, and mixture viscosity [9, 10, 11]. During pressure relief caused by a runaway chemical reaction the mixture properties will dynamically and continuously change. Providing a consistent set of these thermodynamic and transport properties already dictates the use of detailed thermophysical and transport properties models. This important fact contradicts the use of the simplified single parameter ω method for establishing the mass flow rate and choking conditions. The ω method is a simplified equation of state with a limited range of applicability which has an implicit imputed value of the mixture speed of sound [7, 8] for nozzle and pipe flow calculations.

The limitations shown in Table 1 render the sizing based on the single parameter ω method outlined in ISO 4126-10 essentially useless (or at best minimally useful) for many industrial systems and applications [7, 12]. The use of the ω method has introduced the need to perform iterative complex calculations that would otherwise be much simpler and more accurate using direct numerical vdP integration for different types of flow and where condensation, degassing, body bowl choking,

[1] International Organization for Standardization

Figure 1: Calculated critical conditions for ethane/heptane mixture



Source: SuperChems Expert

and/or multiple chokes are possible.

3 Onset and Disengagement of Twophase Flow

A method for determining if twophase flow can occur is outlined in ISO 4126-10 [1]. ISO 4126-10 points out in section 6.3 that it is important to determine if twophase flow is going to occur at the inlet of the relief line.

However, it is also important to determine the vapor/liquid ratio (slip in the vessel or vapor quality) that enters the vent line (see Sections 3.1 and 3.2). The correct slip ratio entering the vent line must yield the same flow rate using the American Institute of Chemical Engineers (AIChE) Design Institute for Emergency Relief Systems (DIERS) coupling equation (see Equation 1) and the vent line / nozzle flow equation(s). This important step is completely missing from ISO 4126-10. One cannot just use an arbitrary vapor to liquid ratio entering the vent line that does not satisfy the DIERS coupling equation.

In addition, there is no practical reason to recast the well known DIERS Coupling equation α vs. ψ chart and limit its applicability as shown in Figure 4 in ISO 4126-10 [1]. As mentioned earlier, mixture surface tension and viscosity properties are required and will influence the DIERS vapor/liquid disengagement estimates.

The DIERS coupling equation is widely used and has been shown to reproduce large scale test data [13, 14]. It can be applied to quench tanks, systems with non-boiling height considerations, and systems with wall heating considerations [13].

The DIERS α vs. ψ curve is shown in Figure 2. To determine if a particular venting rate will result in twophase flow, one can simply locate the associated ψ and void fraction point on the chart. If the point is above the selected flow regime curve, then all vapor flow is predicted. If the point is below the curve, then twophase flow will occur.

3.1 Vapor Quality Entering Vent Line

If twophase flow conditions are predicted, the weight fraction of vapor entering the relief device or vent line, ($\mathcal{Y}$), is the vapor weight fraction which satisfies the following relation:

$$\frac{\mathcal{Y} G_m A_h}{\epsilon \zeta u_\infty \rho_v A} = \frac{1}{1 - C_0 \epsilon \frac{\rho_v}{\rho_l} \frac{1-\mathcal{Y}}{\mathcal{Y}}} \tag{1}$$

where C_0 is a bubble rise velocity correlating parameter (k_∞ in ISO-4126-10), ϵ and ζ are flow dependent parameters given as function of the vessel average void fraction. G_m is the mixture calculated mass flux typically using vdP direct integration, A is the vessel cross sectional flow area, A_h is the relief device flow area, u_∞ is the bubble rise velocity, ρ_v is the vapor mass density, and ρ_l is the liquid mass density . This equation is often referred to as the DIERS coupling equation.

For **bubbly flow**:

$$\epsilon = \frac{\alpha}{1 - C_0 \alpha} \text{ and } \zeta = \frac{(1-\alpha)^2}{1-\alpha^3} \tag{2}$$

For **churn flow**:

$$\epsilon = \frac{2\alpha}{1 - C_0 \alpha} \text{ and } \zeta = 1 \tag{3}$$

Calculations involving partial vapor-liquid disengagement can be computationally intensive as they require calculation of G_m at each estimate of $\mathcal{Y}$. Note that at very large superficial vapor velocities (large vents), the disengagement will occur at a vessel liquid fill fraction equal to $\left(\frac{C_0 - 1}{C_0}\right) = 1 - \frac{1}{C_0}$. C_0 best values recommended by DIERS and ISO 4126-10 are practically equivalent, $k_\infty = 1.18$ for bubbly flow ($C_0 = 1.2$), and $k_\infty = 1.53$ for churn turbulent flow ($C_0 = 1.5$).

3.2 Solving the DIERS Coupling Equation

The solution of the DIERS coupling equation requires trial and error. The form represented by Equation 1 has to be rearranged in order to produce a solution without numerical discontinuities as shown by Melhem [15] (also see [16]). The preferred form for a numerical solution is:

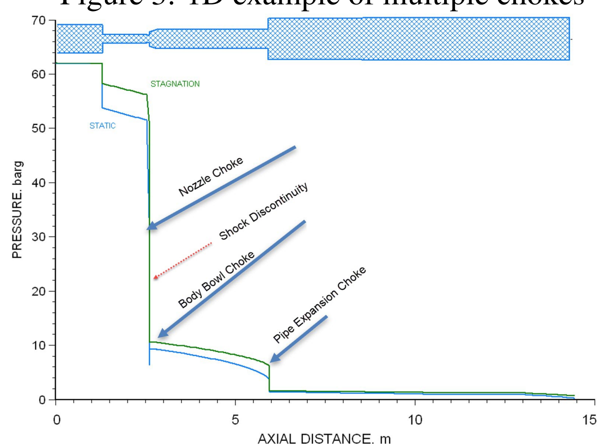
$$f(\mathcal{Y}) \quad = \quad G_m \left(\frac{A_h}{A}\right)\left(\mathcal{Y} - C_0 \epsilon \frac{\rho_v}{\rho_l}(1-\mathcal{Y})\right) - \epsilon \zeta u_\infty \rho_v = 0 \tag{4}$$

The solution begins by guessing the vapor quality entering the vent line, $\mathcal{Y}$, and then by estimating the mass flow through the vent line, G_m, using an appropriate twophase flow model (often homogeneous or slip equilibrium). The calculated value of G_m is inserted in Equation 4 and $f(\mathcal{Y})$ is then evaluated. With this form of the DIERS coupling equation the actual solution of $\mathcal{Y}$ will always be bounded between 0 and 1.

Once a relief requirement has been established, it is a common practice to select the next size up for a pressure relief valve or a rupture disk. These calculations have to be repeated using the actual selected pressure relief device because large enough relief devices can induce twophase flow even though the initial established flow phase may have been all vapor.

4 Twophase Discharge Coefficient

Figure 3: 1D example of multiple chokes



Source: Process Safety Office® SuperChems

Section 6.5.2 of ISO 4126-10 provides equation 49 for calculation of the twophase discharge coefficient. This equation yields a twophase discharge coefficient as a volumetric average at relief conditions of the certified all liquid and all vapor discharge coefficients with a liquid viscosity correction factor for high viscosity liquid flow:

$$K_{\text{dr,2ph}} \quad = \quad K_{\text{dr,g}} \alpha_{\text{seat}} + (1-\alpha_{\text{seat}}) K_{\text{dr,l}} K_{\text{v}} \tag{5}$$

where $K_{\text{dr,g}}$ is the certified vapor discharge coefficient, $K_{\text{dr,l}}$ is the liquid certified discharge coefficient, $K_{\text{dr,2ph}}$ is the twophase discharge coefficient, α_{seat} is the void fraction at relief conditions at the flow limiting area of the safety relief device, and K_{v} is liquid viscosity corrector factor to be used if the liquid viscosity exceeds 100 cp.

We first point out that $K_{\text{dr,l}}$ is measured or obtained under subsonic flow conditions while $K_{\text{dr,g}}$ is measured or obtained under choked (sonic flow) conditions [2]. It is well known that $K_{\text{dr,g}}$ will always be greater than $K_{\text{dr,l}}$ because the relief device geometry downstream of the nozzle does

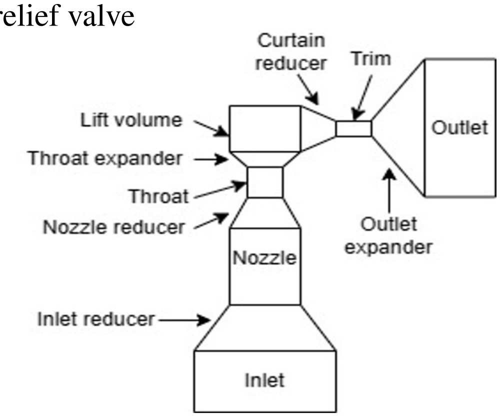
[2] The terms “choked flow”, “sonic flow”, and “critical flow” are equivalent.

not influence the mass flux for choked flow. For subsonic flow the relief device geometry causes additional pressure loss which is why $K_{\mathrm{dr,l}}$ is less than $K_{\mathrm{dr,g}}$.

The twophase discharge coefficient $K_{\mathrm{dr,2ph}}$ should be determined based on whether the flow is choked or not [18, 19]. The twophase flow choking boundary is typically larger than that of all vapor flow. If the starting pressure is large enough and there are flow area increases downstream of the relief device, multiple chokes can occur [20] (see Figure 3). Even if there are no flow area increases downstream of the relief device, body bowl choking can also occur.

Vapor quality has a significant influence on the location of the choke point and choking conditions and therefore mass flow. For nozzles less than 100 mm in length, non equilibrium flow can occur [21, 22].

Figure 4: 1D representation of a pressure relief valve



Source: [17]

For subcooled flow with a large degree of subcooling, the vapor quality at the choke point will be 0 and the flow is determined using the pressure differential between the starting pressure and the choke pressure as all liquid flow. For subcooled flow with a small degree of subcooling, the choke point can shift from the relief device nozzle (throat) to trim to outlet. This was recently demonstrated by [17] using a 1D representation of a pressure relief valve (see Figure 4) and Process Safety Enterprise® SuperChems for DIERS Lite.

Regardless of flow type, it is recommended to use the gas flow discharge coefficient for choked flow and the liquid discharge coefficient for subsonic flow. There is no physical basis for interpolating between sonic (choked) and subsonic (unchoked) flow conditions. Sonic flow conditions are followed by a shock discontinuity downstream of the choke point (see Figure 3) and it is physically meaningless to interpolate flow conditions through that discontinuity [23, 17, 24] as outlined in Section 6.5.2 of ISO 4126-10. We acknowledge that there is a discontinuity when switching discharge coefficients, especially for small degrees of subcooling [3]. Under no circumstance should the calculated twophase flow rate exceed the calculated flow rate for all liquid flow [17].

5 Mass Flow and Pressure Change in Vent Line System

ISO 4126-10 [4] indicates that "the pressure change through the vent line shall be simultaneously calculated and the built-up back pressure evaluated". This requirement cannot be properly satisfied without iterative trial and error calculations for mass flow and pressure.

The use of a single parameter ω method to properly represent the inlet line, the safety relief device, and the discharge line is unduly complex and riddled with traps. It is like trying to fit a square peg

[3] See May 22, 2024 "DIERS Update" by G. A. Melhem to the API Subcommittee of Pressure Relief Systems.
[4] See ISO 4126-10 Section 6.7

in a round hole. The use of a different method called the HNE-CSE [5] method is also discussed for mass flow calculation.

Both of these methods do not properly address the entrance velocity kinetic energy flow contribution to the safety relief device, body bowl choking in the pressure relief valve, and/or how to properly calculate multiple chokes and their associated pressure discontinuities in the discharge line. ISO 4126-10 indicates that the single parameter ω can be fit using 80 % or 90 % of the initial source pressure. However, this is not sufficient for typical plants relief line installations, especially for high pressure systems. Both of these methods will require numerous single parameter fits in order to properly represent the pressure profile in the entire relief line.

A single parameter ω equation is just a reduced analytical model or a simple equation of state with limited applicability range. The method was introduced many years ago by Leung [3] to simplify the integration of nozzle flow. This method has numerous shortcomings as discussed by ISO 4126-10 (see Table 1) and Melhem [6, 7].

$$\frac{v_m}{v_o} = \frac{\rho_o}{\rho_m} = 1 + \omega \left[\left(\frac{P_o}{P} \right) - 1 \right] \tag{6}$$

Where v is the specific volume, ρ is the mass density. The subscripts m and o refer to mixture and initial source conditions.

For nozzle flow, it is far better and simpler to directly integrate vdP using a real equation of state and several pressure points along a specified thermodynamic path to locate the throat pressure, P_t that leads to a maximum value in mass flux:

$$G^2 = \rho_{m,t}^2 \left(2\Delta h + u_o^2\right) = 2\rho_{m,t}^2 \int_{P_t}^{P_o} v_m dP + \rho_{m,t}^2 u_o^2 = \frac{-2\int_{P_o}^{P_t} v_m dP + u_o^2}{\underbrace{\left[\frac{x_t}{\rho_{g,t} S_t} + \frac{1-x_t}{\rho_{l,t}}\right]^2 \left[x_t S_t^2 + 1 - x_t\right]}_{\left(\frac{1}{\rho_{m,t}}\right)^2}} \tag{7}$$

Where h is the fluid specific enthalpy, $S_t = \frac{u_g}{u_l}$ is the velocity slip ratio at the throat pressure and temperature conditions, and u_o is the nozzle entrance velocity, typically 0 for vessel flow.

For the simple single parameter ω equation, given a nozzle throat pressure, P_t, the mass flux can be calculated as follows:

$$G^2 = -2\rho_m^2 \left[v_o (1-\omega)(P_t - P_o) + v_o \omega P_o \ln\left(\frac{P_t}{P_o}\right) \right] + \rho_m^2 u_o^2 \tag{8}$$

For all liquid flow, $\omega \rightarrow 0$. For all vapor flow, $\omega \rightarrow 1$. $0 < \omega < 1$ for non-flashing flow and $\omega > 1$ for flashing flow. The value of P_t that maximizes G has to also be found by trial and error.

It is very important to note that the single parameter ω shown by equation 6 has an implicit speed of sound, $c_{s,t}$. This implicit speed of sound has a strong effect on use of the ω method for pipe flow

[5] Homogeneous Non-Equilibrium Consistent Sizing Equations.

as it will essentially dictate the choke point in a constant flow area pipe:

$$c_{s,t} = v_m \sqrt{\frac{P_t^2}{\omega v_o P_o}} \tag{9}$$

Generalized reduced analytical models provide both speed and solution stability advantages for complex pipe flows. Since the thermodynamic path is already specified, the only variable that requires integration is pressure [7]:

$$\frac{\partial P}{\partial x} = \frac{\rho u^2 \frac{1}{A(x)} \frac{\partial A(x)}{\partial x} - \left[\frac{\partial P}{\partial x}\right]_{\mathrm{F}}}{1 - \frac{u^2}{c^2}} = \frac{\rho u^2 \frac{1}{A(x)} \frac{\partial A(x)}{\partial x} - \left[\underbrace{\rho g \sin\theta + \rho \frac{f u|u|}{\sqrt{A(x)/\pi}} + \frac{\rho}{2} u|u| \frac{K}{L}}_{\text{Flow Resistance}}\right]}{1 - \frac{u^2}{c^2}} \tag{10}$$

where u is the flow velocity, K is the number of velocity head losses due to fittings, L is the total length of the pipe segment, f is the friction factor, $A(x)$ is the pipe flow area as a function of axial distance x, g is the gravitational constant, θ is the pipe segment angle with respect to horizontal, and c is the speed of sound implied by the reduced analytical model.

Since $\frac{\partial \rho}{\partial P}$ can be directly obtained from the generalized reduced analytical model for single and multiphase systems where slip between the phases can be considered, we can obtain c from $\frac{\partial \rho}{\partial P} = \frac{1}{c^2}$.

It is interesting to note from equation 10, that when sonic or choked flow is reached ($u = c$), the change of pressure with respect to axial distance tends to infinity, $\frac{\partial P}{\partial x} \to \infty$. Alternatively, the $\frac{\partial x}{\partial P} \to 0$ at the choke point and $\frac{\partial x}{\partial P}$ becomes < 0 for supersonic flow.

Because plant piping systems can be complex with varying flow areas and orientation leading to pressure recovery in some piping segments, the solution is typically integrated in time instead of axial distance:

$$u = \frac{\dot{m}}{\rho A(x)} \tag{11}$$

$$\frac{\partial x}{\partial t} = u \tag{12}$$

$$\frac{\partial P}{\partial t} = u \frac{\partial P}{\partial x} \tag{13}$$

where $\dot{m}$ is a specified mass flow rate that is a trial and error value. $\dot{m}$ is increased until sonic flow is achieved and the entire pipe length is traversed by the flow after resolution of all shock pressure discontinuities. For subsonic flow, the flow has to reach the end of the pipe and the pressure has to equal the ambient backpressure.

ρ is calculated directly from the generalized reduced analytical model as pressure is changing along the axial distance. Temperature is also calculated directly from the generalized reduced analytical model as a function of pressure. Viscosity is then calculated as a function of temperature and pressure.

Mixture viscosity is required to calculate frictional pressure drop for pipe flow. When suspended solids are present in the mixture (for example polymer systems), the mixture viscosity can be adjusted using the Thomas multiplier [25] to account for the impact of suspended solids on overall viscosity:

$$C_{vs} = (1-\alpha)\,\beta_S \tag{14}$$

$$\frac{\mu_{ms}}{\mu_m} = 1 + 2.5\,C_{vs} + 10.05\,C_{vs}^2 + 0.00273\,\exp\left(16.6 C_{vs}\right) \tag{15}$$

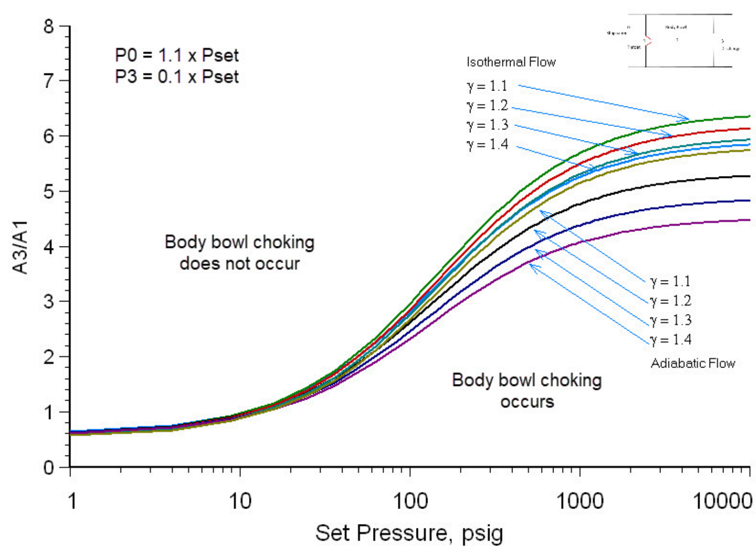
where β_S is the local volume fraction of solids in the condensed phase, C_{vs} is the overall solids volume fraction and μ_{ms} is the pseudo homogeneous mixture-solids viscosity. Equation 15 applies to flows with suspended small solids particles where solids particles settling is negligible. Suspended solids are not addressed by ISO 4126-10.

Because of their inherent shortcomings, ω based methods should be removed from ISO 4126-10 in favor of direct numerical solutions of Equations 7 and 10. Alternatively, users of ISO 4126-10 can easily qualify other methods such as Process Safety Office® SuperChems which provides complete and detailed solutions of the entire vent line systems as well as coupled vent lines and vessel dynamics (with and without chemical reactions) to be used to comply with ISO 4126-10. This is mentioned in ISO 4126-10 section 6.5: “Alternative methods are available, however, it is important to ensure that any method selected is relevant to the particular application and is correctly applied by those appropriately qualified and experienced”.

6 Body Bowl Choking in Gas Systems

Body bowl choking can be an issue for both vapor and twophase flow. ISO 4126-10 does not provide a method for the identification and detection of body bowl choking and for proper calculation of the body bowl choke pressure discontinuity.

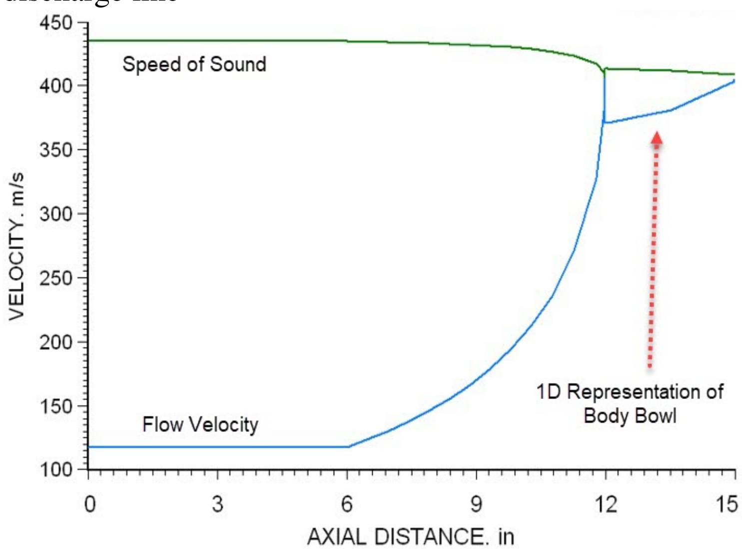
Figure 5: Body bowl choking regions for an ideal gas



Source: [26]

This is particularly important for twophase flow as the body bowl choking pressure becomes the back pressure impacting the force balance on the PRV disk and can negatively impact PRV stability, especially for conventional spring loaded pressure relief valves. Choking pressure ratios are wider for twophase flow than for gas or vapor flow which is why pressure relief valves used in twophase flow service are almost always of the balanced type. Body bowl choking becomes more pronounced at higher pressure relief valve set pressures (see Figure 5).

Figure 6: Body bowl choking for a gas system without a discharge line

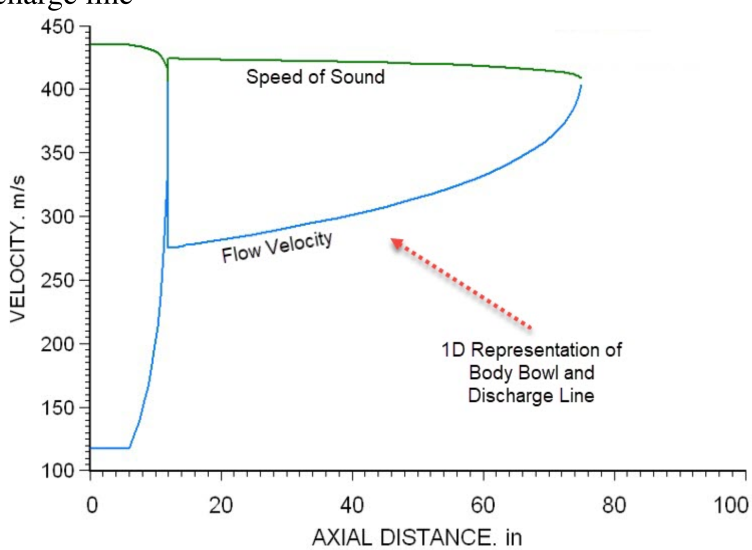


Source: Process Safety Office® SuperChems

Pressure relief valves with large nozzle flow areas relative to discharge flange connections [6] flow areas are more susceptible to body bowl choking. This includes pressure relief valves with J, L, P, Q, R, and T flow orifice areas designations [27]. To properly capture body bowl choking, a small section of pipe should be added to the 1D piping representation of the relief line [8, 26]. It is recognized that the flow patterns in the valve body can be complex [28] and that a 1D representation is only useful to capture a reasonable representative pressure value that can be used as the back pressure on the valve disc [29].

We illustrate some of the concepts of body bowl choking using all gas flow for simplicity. Figure 6 illustrates body bowl choking for a gas system with a ratio of outlet flange flow area to nozzle flow area of 2. Two choke points are identified: (a) a nozzle choke point and (b) a body bowl choke point.

Figure 7: Body bowl choking for a gas system with a discharge line



Source: Process Safety Office® SuperChems

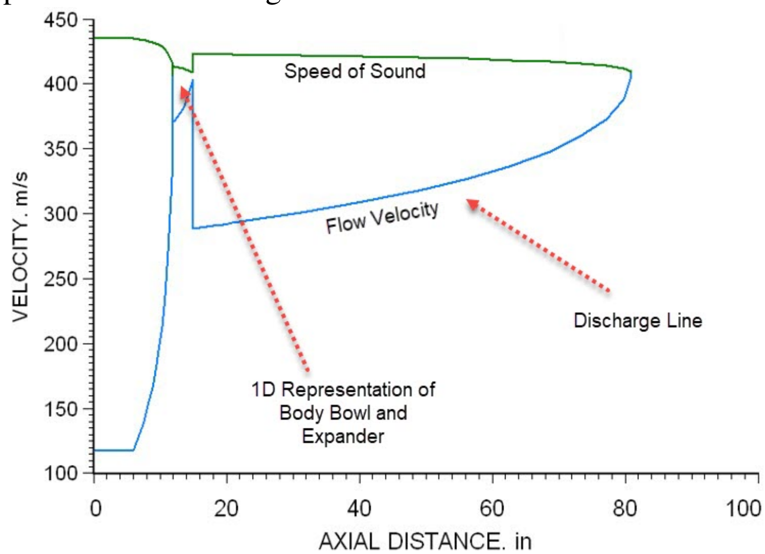
In Figure 7 we add a discharge line that has the same diameter as the PRV outlet flange. Adding a discharge line with the same diameter as the PRV outlet flange shifts the choke point to the end of the line. Longer discharge line segments can cause the choke point to shift to different locations/segments downstream of the first choke point that is regulating the flow. A very long discharge line segment can cause significant loss of pressure leading to subsonic flow in upstream locations and can ultimately become the flow

[6] Often referred to as β ratios.

regulating piping segment. This is common in relief lines where the relief device is only a rupture disk where the flow is regulated by the entire relief line.

In Figure 8 we add an expander to the PRV outlet flange to enlarge the diameter of the discharge line. We notice that we now have three choke locations: (a) one choke at the PRV nozzle, (b) one choke at the PRV body bowl, and (c) one choke at the end of the discharge line. The first choke regulates the flow, the second choke regulates the backpressure to the PRV and the third choke regulates the end of pipe exit pressure.

Figure 8: Body bowl choking for a gas system with an expander and a discharge line



Source: Process Safety Office® SuperChems

7 Body Bowl Choking in Twophase Systems

The same body bowl choking can occur in two phase flow with added complexities due to phase change, non-equilibrium flow, and slip between the two phases. The choke point and conditions are highly influenced by the initial vapor quality. Body bowl choking can be more severe for for slightly subcooled flow or flows with low levels of inlet vapor content (see Figure 9). The difference between equilibrium and non-equilibrium nozzle flow becomes less important as the inlet vapor content or quality increases [21, 22, 29].

For twophase flow, homogeneous equilibrium flow results in higher body bowl choking pressures than slip equilibrium flow as demonstrated by Huff [29]. This high level of body bowl pressure for homogeneous equilibrium flow may be unreasonably restrictive. Operating experience with twophase flow systems indicates that slip equilibrium occurs is the body bowl and discharge line.

The considerations discussed above make it practically impossible to account for body bowl choking in twophase flow systems without the aid of a computer code such as Process Safety Office® SuperChems.

8 Multiple Chokes

Multiple choke point identification and detection and the calculation of the associated pressure discontinuities are well beyond the practicality of using a method such as the one outlined in

ISO 4126-10. ISO 4126-10 does not provide an adequate/practical method for the calculation of multiple chokes and their associated pressure discontinuities in the relief line.

Temperatures associated with pressure discontinuities in the discharge line cannot be properly calculated for gas/vapor flow using an ω like method because the thermodynamic path for a discontinuity is different than the thermodynamic flow path for a nozzle. Temperature values can be bounded however, by producing two sets of calculations with bounding thermodynamic flow paths, isentropic and isenthalpic (constant stagnation enthalpy). Temperature values associated with pressure discontinuities are reasonably represented by generalized reduced analytical models as long as the flow is twophase.

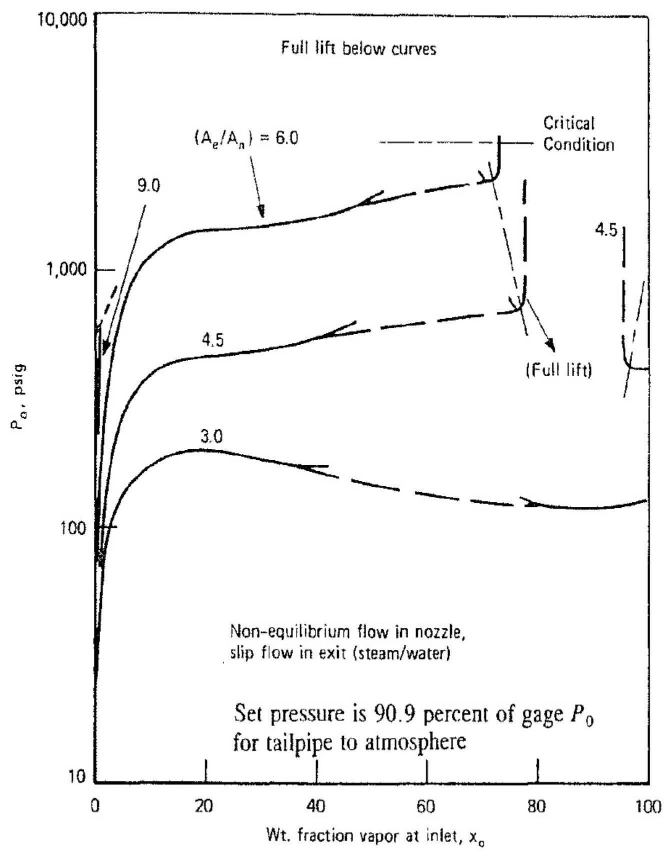
The identification of multiple chokes [20] requires iterative calculations and resolution of irreversible shock discontinuities. If the starting pressure is high enough, multiple chokes can occur in relief line downstream piping. Multiple chokes in relief piping should be avoided because they increase acoustic induced vibration failure risks.

9 PRV Stability

Section 6.6 discusses PRV stability. Regardless of flow phase, it is now well known that the 3 % irrecoverable pressure loss rule is not sufficient to guarantee PRV stability [30, 31, 32, 33, 34] (also see [35, 36, 37, 38, 39]).

Recent work by numerous organizations and researchers have highlighted the importance of the pipe/fluid speed of sound [8, 40, 41], force balance [42], and critical line length [43, 44] on PRV stability.

Figure 9: Body bowl back pressure limitation vs. inlet vapor quality at 10 % overpressure for steam/water system



Source: [29]

The current 3 % inlet line pressure loss rule should not be used unless the inlet line is shorter than the critical length (80 % of critical length). The critical line length is significantly dependent on the fluid/pipe speed of sound [8, 40, 41]. When using 3 % irrecoverable pressure loss as the sole criterion for PRV stability, the inlet line length must be less than the critical line length and the backpressure must be within tolerable limits. If the critical line length is not used, then the total percent pressure drop (frictional and dynamic) must be less than the percent blowdown minus 1 or 2 percent.

The 3 % rule should be replaced with the API force balance coupled with critical line length for simple piping geometries, where the inlet line length is less than the critical line length criterion [42, 44]. Detailed 1D dynamics [30] should be used for complex piping geometries, especially where the inlet line length is greater than the critical line length.

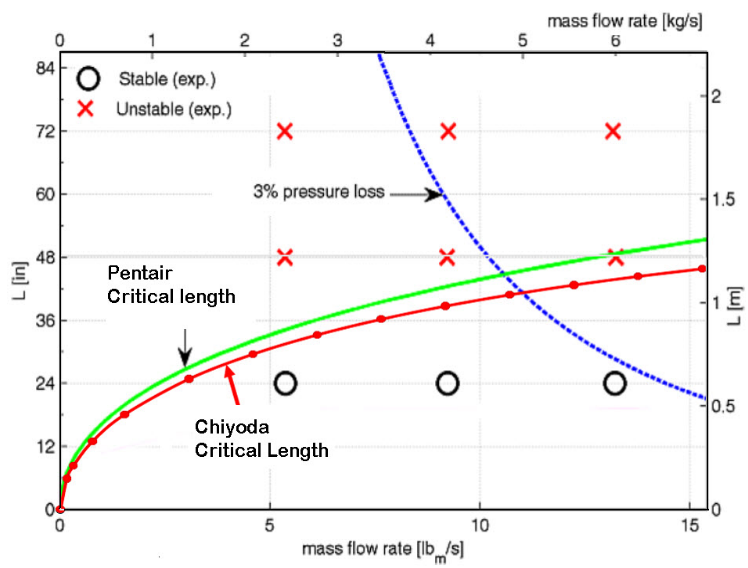
ISO 4126-10 recognizes that PRV stability has to be assessed by the user but does not recommend a specific engineering analysis. Such engineering analysis is addressed by API 520-II [45] and Melhem [30, 42].

10 Recommendations

It is highly recommended that the working group of ISO/TC 185 considers the critical items identified in this paper in the 2030 revision/update of ISO 4126-10 before expanding its scope [28].

In particular, the use of the ω method is outdated and renders the calculation methods in ISO 4126-10 essentially useless. The use of the ω method has introduced the need to perform iterative complex calculations that would be much simpler using direct numerical vdP integration for different types of flow and where condensation, degassing, body bowl choking, runaway reactions, and/or multiple chokes are possible.

Figure 10: Inlet line length stability limit as a function of disk lift ratio or mass flow rate



Source: [43]

Because of their inherent shortcomings, ω based methods should be removed from ISO 4126-10 in favor of direct numerical solutions of Equations 7 and 10 or well established and validated computer codes such as Process Safety Office® SuperChems. We note that SuperChems for DIERS Lite is provided as a companion software to the 2nd Edition of the CCPS Guidelines for Pressure Relief Systems and Effluent Handling [46].

Table 1: ISO 4126-10 Twophase flow method limitations

Criterion	Limit	Comments
Flashing flow	$\frac{T_{\text{over}}}{T_c} < 0.9$ or $\frac{P_{\text{over}}}{P_c} < 0.5$	T_c and P_c are the critical temperature and pressure of the single component or the mixture.
Condensing flow	Do not use	Typically encountered in high pressure ethylene and propylene systems or systems where vapor is at conditions close to the dew point.
Flashing flow for mixtures	Boiling point range < 100 K	The difference between components having the highest and lowest bubble point temperatures at relief conditions.
Dissolved gases	Do not use	Small amounts of dissolved gases (such as nitrogen) in liquids can drastically change the speed of sound [8] of the mixture, change the mixture properties and the mass flow rate through the relief device. Speed of sound is a critical factor for PRV stability.
Omega value	$0 \leq \omega \leq 100$	ω is a measure of the fluid compressibility. The value of ω ranges from 0 for all liquid flow, 1 for all vapor/gas flow, greater than 1 for flashing flow and between 0 and 1 for non-flashing flow.
Immiscible liquids	Do not use	Typical of systems involving emulsion polymerization for example.
Viscosity	< 100 cp	If more than 100 cp, assume homogeneous equilibrium venting occurs and the void fraction entering the relief device is the same as the vessel average void fraction.
Runaway reaction	$\frac{dT}{dt} < 120\ \mathrm{C/min}$ and $\frac{dP}{dt} < 12\ \mathrm{bar/min}$	Limiting values are at relief conditions.

T_{over} and P_{over} are overtemperature and overpressure during relief.

Source: ioMosaic Corporation and ISO 4126-10 Section 5 [1, 2]

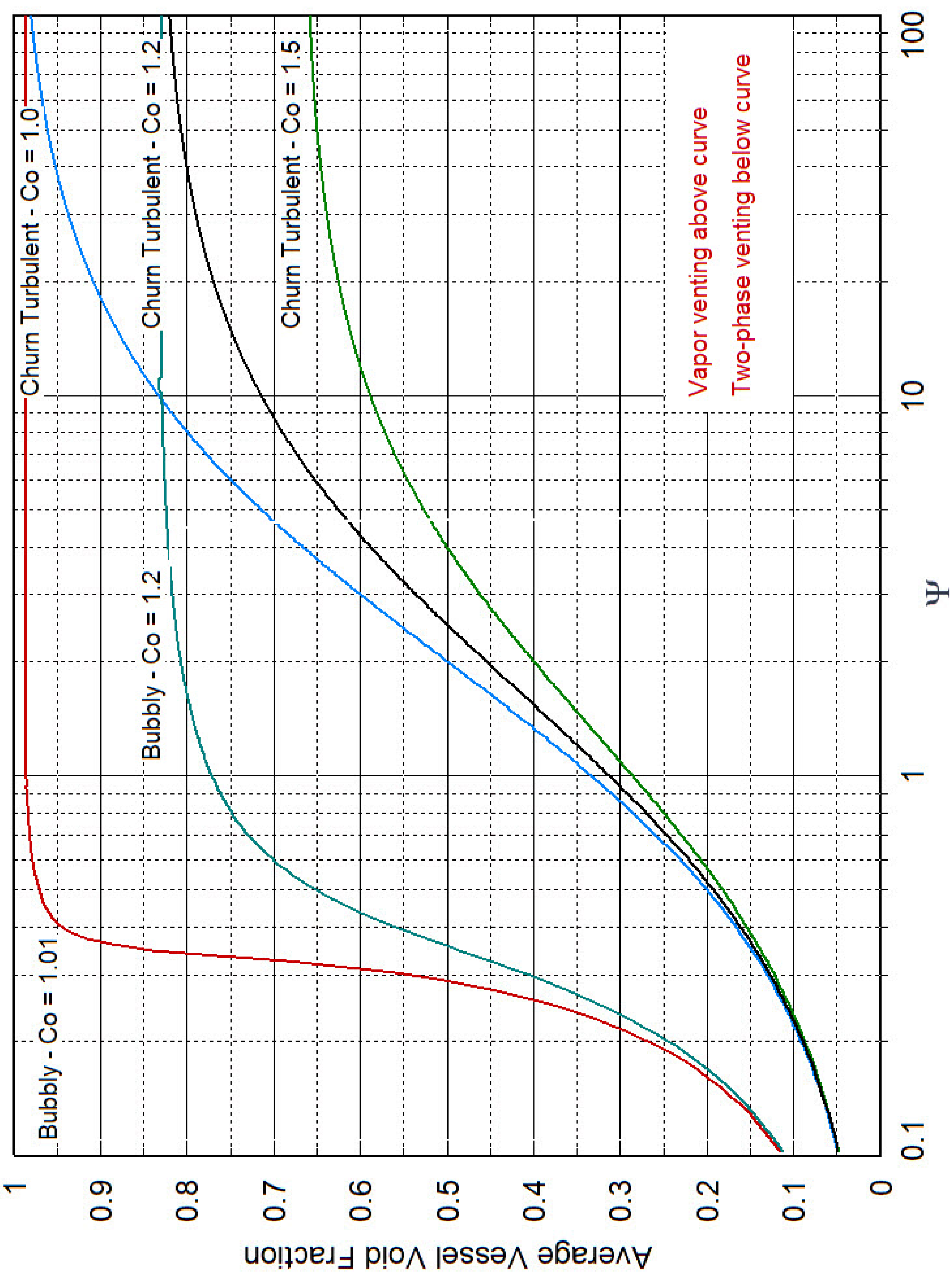


Figure 2: DIERS α (vessel average void fraction) vs. $\psi = \frac{u_{sv}}{u_\infty}$ (dimensionless superficial vapor velocity due to flow) curve

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About the Authors

Dr. Melhem is an internationally known pressure relief and flare systems, chemical reaction systems, process safety, and risk analysis expert. In this regard he has provided consulting, design services, expert testimony, incident investigation, and incident reconstruction for a large number of clients. Since 1988, he has conducted and participated in numerous studies focused on the risks associated with process industries fixed facilities, facility siting, business interruption, and transportation.

Prior to founding ioMosaic Corporation, Dr. Melhem was president of Pyxsys Corporation; a technology subsidiary of Arthur D. Little Inc. Prior to Pyxsys and during his twelve years tenure at Arthur D. Little, Dr. Melhem was a vice president of Arthur D. Little and managing director of its Global Safety and Risk Management Practice and Process Safety and Reaction Engineering Laboratories.

Dr. Melhem holds a Ph.D. and an M.S. in Chemical Engineering, as well as a B.S. in Chemical Engineering with a minor in Industrial Engineering, all from Northeastern University. In addition, he has completed executive training in the areas of Finance and Strategic Sales Management at the Harvard Business School. Dr. Melhem is a Fellow of the American Institute of Chemical Engineers (AIChE) and Vice Chair of the AIChE Design Institute for Emergency Relief Systems (DiERS).

Contact Information

Georges. A. Melhem, Ph.D., FAIChE
E-mail. melhem@iomosaic.com

ioMosaic Corporation
93 Stiles Road
Salem, New Hampshire 03079
Tel. 603.893.7009, x 1001
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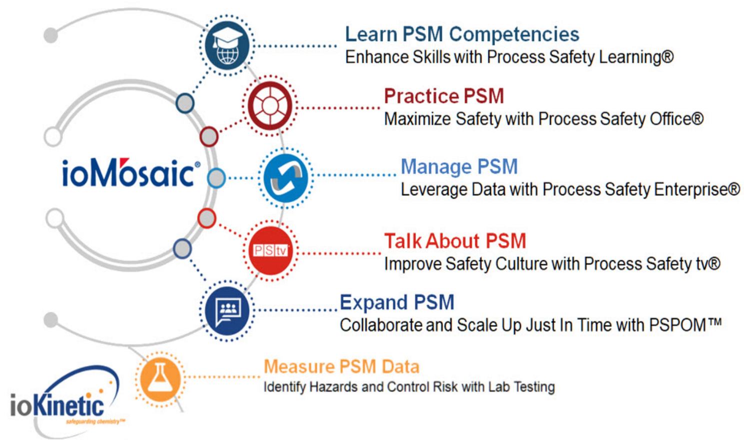

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Contact us

www.ioMosaic.com
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1.844.ioMosaic