Emergency Relief Systems—Sizing and Scale-Up

A practical and safe, but not overly conservative, sizing method is detailed.

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The sizing of emergency relief systems to assure adequate venting in connection with runaway chemical reactions generally involves considerations of scale-up [1, 2]. For any new reaction, scaling-up to production reactor sizes from laboratory-scale tests may be necessary and the question arises as to whether a simple scale-up rule can be used. One such rule which has seen frequent application is to scale on the basis of a fixed ratio of vent area to reactor volume [1] or vent area to heat release [2]. However, since most large-scale applications are likely to involve liquid-vapor venting [1-5], a safe application of this simple rule would appear to require that the scale-up remain insensitive to the phase ratio of vapor and liquid entering the vent line. In this paper we shall therefore explore the sensitivity of the vapor-disengagement assumption as it effects the sizing of safety relief valves for several typical chemical reaction systems.

VAPOR DISENGAGEMENT PHENOMENOLOGY

Within a process vessel, friction and momentum effects are generally not important. The drift-flux theory [6] is therefore particularly useful where the relative motion is governed by a balance between buoyancy and drag forces. The drift flux, j_{gl}, can be written in terms of the individual fluxes j_g and j_l (equivalent to superficial velocities) as

$$j_{gl} = j_g(1 - \alpha) - \alpha\, j_l \tag{1}$$

where α is the local void fraction. As demonstrated by data, including liquid-liquid and gas-liquid systems, the usefulness of the drift-flux concept is that it can be expressed in terms of basic quantities as follows:

$$j_{gl} = \alpha(1 - \alpha)^n\, V_\infty \tag{2}$$

where V_∞ is the terminal velocity of a single particle in an infinite fluid and n is an exponent which varies principally with the flow regimes.

In the case of a bottom-injected, gas-sparged vessel, the local void fraction is approximately equal to the average void fraction ($\bar{\alpha}$), i.e., j_g is invariant with the height of the process vessel. *On the other hand, in the case of a volumetrically heated system such as a chemically reacting system, j_g varies with height.* For a constant-pressure, adiabatic, volumetrically heated boiling pool of constant cross-section, the one-dimensional vapor mass-balance equation can be written as

$$\frac{dj_g}{dz} = \frac{q(1 - \alpha)\rho_l}{h_{fg}\,\rho_g} \tag{3}$$

where q is the equivalent heat generation in the liquid phase, h_{fg} is the latent heat of vaporization, and ρ_g and ρ_l are the vapor density and the liquid density, respectively. The above set of equations, together with an appropriate selection of flow regimes, can be used to indicate 1) the boundary between all-vapor venting and two-phase venting, and 2), in the case of two-phase venting, the vapor-to-liquid phase ratio entering the vent line.

Vapor Venting

To assure all-vapor venting (complete vapor disengagement) the following inequality must be satisfied

$$\bar{\alpha} < \text{vessel free-board volume } (\bar{\alpha}_o) \tag{4}$$

where $\bar{\alpha}$ is determined by setting $j_l = 0$ in Equation 1. This condition, together with Equation 2, results in

$$j_g = \alpha(1 - \alpha)^{n-1}\, V_\infty \tag{5}$$

Equation 5 in differential form can be combined with Equation 3 and integrated to yield the necessary relationship between j_g and $\bar{\alpha}$.

Assuming ideal-churn turbulent flow ($n = 0$) [6], the following simple result is obtained

$$j_{g\infty}/V_\infty = \frac{2\bar{\alpha}}{1 = \alpha} \tag{6}$$

For a vertical cylindrical vessel $j_{g\infty}$ is related to known quantities by

$$j_{g\infty} = \frac{H(1 - \bar{\alpha}_o)\rho_l q}{h_{fg}\,\rho_g} \tag{7}$$

where H is the vessel height.

A number of applicable volume-heated boiling-pool experiments have been reported recently using water and aqueous electrically conducting solutions. Disregarding the data which clearly show extended bubbly and foaming behavior [7, 8], the non-foaming data are reasonably consistent, despite the obvious scatter in results. Typical data, obtained from both microwave (water) and electrically heated apparatus (water/zinc sulfate) are reasonably well characterized by the ideal-churn turbulent drift-flux model Equation 6 for average void-fraction values up to 0.5 (see Figure 1). For values larger than 0.5, the data (not shown in Figure 1) start to deviate significantly from the ideal-churn turbulent model and is better described by including a distribution parameter $C_o = 1.35$ in the above drift-flux formulation [8]*.

*In this case Equation 5 would have the form

$$j_g = \frac{\alpha(1 - \alpha)^n}{1 - C_o\alpha}\, V_\infty$$

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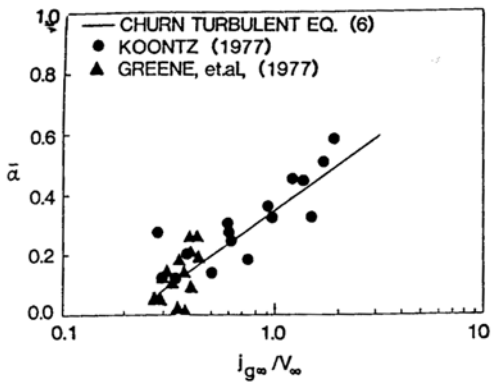

Figure 1. Some typical non-foaming pool-average void fraction-measurements and comparison with the churn turbulent drift flux function ($j_l = 0$).

Two-Phase Venting: Liquid-Vapor Phase Ratio

To determine the effect of partial vapor disengagement, the vapor-to-liquid phase ratio entering the vent line must be specified. In this case, j_l must be retained in Equation 1. Again assuming churn turbulent flow, j_g and j_l can be related to α as follows

$$j_g = \frac{\alpha}{1-\alpha}(V_\infty + j_l) \tag{8}$$

j_g and j_l can also be related to the vapor-liquid phase ratio, x, entering the relief line by writing

$$j_g \rho_g A_v = Gx\, A_{\text{vent}} \tag{9}$$

$$j_l \rho_l A_v = G(1-x)\, A_{\text{vent}} \tag{10}$$

where G is the total mass flow rate, A_v is the cross-sectional area of the reactor vessel, and A_{vent} is the equivalent cross-sectional area of the vent line, resulting in

$$j_l = \frac{\rho_g}{\rho_l}\frac{1-x}{x} j_g \tag{11}$$

j_l can be eliminated by substituting Equation 11 into Equation 8

$$j_g = V_\infty \Big/ \left(\frac{1-\alpha}{\alpha}\frac{\rho_g}{\rho_l}\frac{1-x}{x}\right) \tag{12}$$

At pressure equilibrium, i.e., turn around in pressure, we can write*

$$\frac{A_v H(1-\bar{\alpha})\rho_l q}{h_{fg}\,\rho_g} = G\, A_{\text{vent}}\,[v_g x + v_l(1-x)] \tag{13}$$

where v_g and v_l are the specific volumes of vapor and liquid, respectively. Combining Equation 13 with Equations 9 and 12 produces the following expression for the vapor-to-liquid phase ratio entering the relief line

$$x = \frac{1 + V_\infty}{\left(1 - \frac{\rho_l}{\rho_g}\right)\psi + \left(\frac{\rho_l}{\rho_g}\,\frac{1-\alpha}{\alpha} + 1\right)} \tag{14}$$

where ψ is a dimensionless velocity defined as $V_\infty/[H(1-\bar{\alpha})\rho_l q/h_{fg}\rho_g]$. Equation 14 is consistent with $x = 1.0$ ($j_l = 0$) when

$$\frac{1-\bar{\alpha}}{\alpha} = \frac{1-\bar{\alpha}}{2\bar{\alpha}} \text{ where } (\psi = V_\infty/j_g) \tag{15}$$

*Equation 13 ignores the change in liquid volume due to vaporization, i.e., Equation 13 is a good approximation when $\rho_l >> \rho_g$.

Assuming that this is approximately valid when $\psi < V_\infty/j_g$, we obtain

$$x \cong \frac{1 + V_\infty}{\left(1 - \frac{\rho_l}{\rho_g}\right)\psi + \left(\frac{\rho_l}{\rho_g}\,\frac{1-\bar{\alpha}}{2\bar{\alpha}} + 1\right)} \tag{16}$$

SCALING CONSIDERATIONS

It is clear from Equation 7 that the liquid-to-vapor phase ratio entering the vent line is quite sensitive to vessel size ($j_{g\infty}$ is directly proportional to vessel height). It follows that a small test vessel may experience all-vapor venting (the swell height is less than the free-board volume) while a typical size reactor vessel will vent a two-phase, liquid-vapor mixture (swell height exceeds free-board volume) for the same heat release per unit volume. A quantitative illustration of the case is provided in Figure 2 by using Equations 6 and 7 and considering a scale-up from a small test vessel (10 gallons) to a field-size vessel (1250 gallons). The illustration reflects a system possessing organic-like properties (typical of polymerization reactors [*1-4*]), simulated by Freon-12, a free-board volume of 20%, and an energy-release rate of 1 W/g at a set pressure of 689 kPa. Direct application of the simple scale-up rule in this case suggests a significant overpressurization for the process vessel, as illustrated in Figure 3. The two-phase flow is characterized by the Henry-Fauske non-equilibrium flow model [*11*] assuming churn-type turbulent flow behavior ($C_o = 1.0$) within the vessel, while the vent area (0.008 m²)

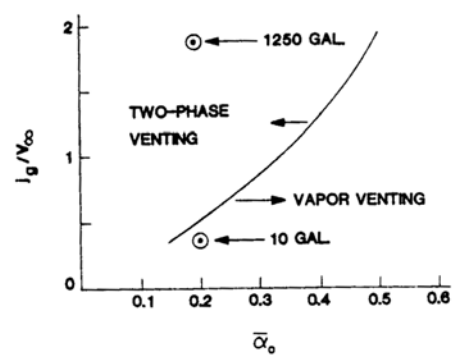

Figure 2. Illustration of boundary between all-vapor and two-phase venting for a non-viscous, non-foaming system in a straight vessel.

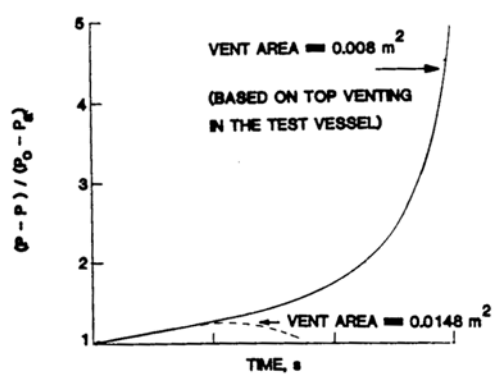

Figure 3. Solid line illustrates pressure runaway in the field application based upon the simple scale-up rule using top venting data from the test vessel. Dotted line illustrates safe venting based upon simple scale-up rule using bottom venting data.

is based upon all-vapor venting as suggested by the simple scaling rule.

Proceeding on the basis that liquid-vapor flow conditions must be considered in sizing emergency relief systems for large process vessels, it is particularly interesting to note that the sensitivity to the asumption made for the inlet quality to the vent line can be shown to be quite small.

Again using the Henry-Fauske critical-flow model to characterize anticipated non-equilibrium flow conditions through safety relief valves, Table 1 shows less than a 10% variation in vent diameter between assuming zero inlet quality (all-liquid) throughout the venting period and inlet qualities described by churn-type turbulent-flow characteristics.

The noted insensitivity to the inlet-quality assumption can be directly related to the large differences in the average vessel void fraction at pressure turn-around for the various venting assumptions (see Table 2). It is this large difference in predicted mass loss (i.e., loss of reactants) which largely compensates for the decreasing volumetric discharge rate going from the churn-type turbulent vapor disengagement assumption to all-liquid venting.

A similar weak sensitivity is noted in Table 3 for systems which possess water-like properties (typical of phenolic-resin reactors [5]*. In summary, the above considerations clearly suggest that, in cases where liquid-vapor venting must be considered, the simple scale-up rule on the basis of constant vent-area to vessel-volume ratio can still be used by replacing small-test vessel data based upon top venting with bottom venting as illustrated in Figure 2. This procedure will assure a safe but not overly conservative design for large process vessels.

*For this illustration a vessel (H/D = 1.5) with a volume of 3.47 m^3 and 80% filled with water with an equivalent heat-generation rate of 2 W/g at a relief-system pressure corresponding to 1000 kPa was used.

TABLE 1. VENT-SIZE VARIATIONS AND INLET-QUALITY ASSUMPTIONS; CALCULATIONS BASED UPON 20% OVERPRESSURE*

Venting Assumption	Vent Size, cm
Churn-type turbulent behavior	12.4
Homogeneous behavior	13.6
All-liquid flow	13.7

*In addition to the equations describing vapor disengagement noted in this paper, the standard mass- and energy-conservation equations were satisfied over the time interval Δt corresponding to the turn-around in pressure. The energy-release rate is assumed to double every 10°C.

TABLE 2. VENTING ASSUMPTIONS AND VOID FRACTIONS AT PRESSURE TURN-AROUND

Venting Assumption	Void Fraction
Churn turbulent	0.49
Homogeneous	0.69
All-liquid flow	0.83

TABLE 3. VENT-SIZE VARIATIONS; AND INLET-QUALITY ASSUMPTIONS BASED UPON WATER-LIKE PROPERTIES AND 20% OVERPRESSURE

Venting Assumption	Vent Size, cm
Churn turbulent flow	6.8
Homogeneous flow	8.0
All-liquid flow	8.0

CONCLUDING REMARKS

Since the liquid-to-vapor phase ratio entering the vent line has been shown to be rather sensitive to the vessel size, it follows that direct application of small-test vessel data (indicating all-vapor venting) to the field-size vessel by scaling on the basis of constant vent-area to vessel-volume ratio may lead to unsafe design practices. However, on the basis that the Henry-Fauske critical-flow model correctly describes flashing vapor-liquid flows through safety relief valves of interest* it is demonstrated that the relief vent size is rather insensitive to variation in the inlet quality, ranging from churn-type turbulent-flow behavior within the vessel to all-liquid flow entering the vent line. On this basis, a practical and safe but not overly conservative sizing method can still be provided by the simple scale-up rule by replacing top-venting with bottom-venting data**.

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DISCUSSION

IAN SWIFT, Fauske & Assoc. Inc.: The area: volume-scaling technique is commonly used and relies on venting tests carried out in rather small vessels typically 400 cm^3 in

*Efforts to further substantiate the validity of the Henry-Fauske critical-flow model to applications of interest to the chemical process industry are included in the program of an organization called the Design Institute for Emergency Relief Systems (DIERS) sanctioned by the American Institute of Chemical Engineers (AIChE) with broad international industry support.

**In this paper, illustrations have been given only of reacting systems in which system pressure is directly related to energy release through a vapor-pressure-temperature relation such as polymerization and phenolic-resin reactors. For reactors in which a decrease in pressure does not always lead to a reduction in reaction rate (such as a runaway peroxide decomposition [12]), application of the vapor-venting method as compared to two-phase venting may result in the largest vent-size requirement. In this case the top-venting data along with the simple scale-up rule will assure a safe vent design.

volume. Could you comment on the validity of this approach in scaling vent areas to full-size reactors?

MIKE GROLMES, Fauske & Assoc. Inc.: As the paper suggests results of small scale *top-vented* tests cannot be reliably scaled directly to full-size reactors. This would be reliable only if the reacting system behaved as a homogeneous mixture (no vapor disengagement). The point of the paper is that direct area: volume scaling may be more reliable if bottom venting is used as the basis in small-scale tests. However, other factors such as relative heat loss effect should always be considered in the interpretation of results and scaling methods.

ALAN DUXBURY, ICI: You mentioned that it may be alright to calculate vent area on the basis of vapor venting and double it? That surprised me. I hadn't actually heard of that practice being recommended where liquid flow could occur although I do know of Boyle's original recommendation that you should size for liquid and double or triple that area.

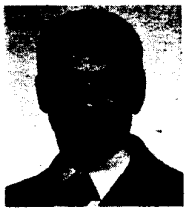
GROLMES: We have found that often a vent area that is twice as large as that required for all vapor venting at the set pressure will be adequate for two-phase venting with allowance for overpressure. The point was not made in the paper since the paper focuses on more technical approach to scaling for two-phase venting conditions.

DUXBURY: I don't think that was my experience. One of the points you made was that the vent size required was not very sensitive to the phase ratio. If you had a long line it could be very different, because if you've got liquid flashing in the line you may have a much smaller volume flow than for vapor of nonflashing liquid. So, you must be very careful.

GROLMES: The paper does not address long vent conditions. We would like to provide a suggested approach for the basic concept of scaling. Accommodation of the added resistance for long tail pipe is a feature that can be included in a rather straightforward manner either empirically or analytically. But, this obviously deserves further elaborations.

DUXBURY: I don't disagree with you. I don't think my comment invalidates your suggestion that one can start with bottom venting to draw conclusions for top venting. I was trying to say that there can probably be a much bigger difference between the sizes you'd need for vapor venting and for two-phase venting if you have a long vent line.

Hans K. Fauske earned the D.Sc. degree from the Norwegian Institute of Technology in 1963. He has been a visiting professor in nuclear and chemical engineering at Northwestern University and was most recently director of the Fast Reactor Safety Technology Management Center at the Argonne National Laboratory. He has published more than 100 scientific articles in the areas of reactor safety and multi-phase flow application and served as a consultant to the President's Commission on the accident at Three Mile Island. Fauske received one of the first University of Chicago awards for Distinguished Performance at the Argonne National Laboratory and is a fellow of the American Nuclear Society.



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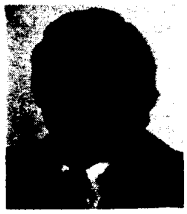

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