SIZING OF RELIEF VALVES FOR TWO-PHASE FLOW IN THE BAYER PROCESS

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ABSTRACT

This paper is to review the methods currently used in engineering design calculations for predicting the relieving capacity of a safety relief valve under various entering flow conditions. The methods considered include the Recommended Practice (RP) 520 of the American Petroleum Institute (API), the Homogeneous Equilibrium Model (HEM) and various published empirical Slip Models.

Recent research conducted by the Design Institute for Emergency Relief System (DIERS) has indicated that the API method leads to undersized relief valves in comparison with HEM under certain conditions. Researchers have found that the experimentally observed relief discharge rates are a factor of three times higher than discharge rates predicted by HEM, especially for low pressure fluids. The Slip Models give results close to experimental data, however there are several correlations from which the slip ratio must be carefully selected to obtain appropriately conservative results.

KEY WORDS

Pressure Relief Valves, API, Homogeneous, Slip

NOTATIONS

A - Area (m^2).
dP - Pressure increment (Pa).
G - Mass flux ($kg/m^2.s$).
k - Ratio of specific heats.
K_b - Correction factor due to back pressure for gas.
K_d - Coefficient of discharge.
K_v - Correction factor due to viscosity.
K_w - Correction factor due to back pressure for liquid.
m - Mass flow rate (kg/s).
M_w - Molecular weight (kg-mole).
P - Upstream relieving pressure (Pa).
T - Temperature (Kelvin).
x - quality in mass fraction.
Z = 1 for ideal gas.
μ - Viscosity (Pa.s).
ν - specific volume (m^3/kg).
ρ - Density (kg/m^3).

Subscripts

c - Critical.
f - Liquid.

P_b - Back pressure (Pa).
R - Gas constant (Pa.m^3/K.kg-mole).
S - Slip ratio = gas velocity / liquid velocity

g - Gas or Vapour.
o - Stagnation.
t - Throat.

1.0 INTRODUCTION

The aim of safety systems in processing plants is to prevent damage to equipment, avoid injury to personnel and to eliminate any risks of compromising the welfare of the community at large and the environment.

Relief systems are therefore designed to passively protect against a predetermined set of “worst case” conditions and are installed to react to these conditions regardless of daily operational activities. In many circumstances relief valves open during operation due to a process upset or as vessels are returned to service after maintenance. There can however be instances where relief valves open and the plant operator is unsure of the reasons why. The cause of these unexplained relief scenarios should be studied to determine why they have occured and whether the relief system has been adequately sized to safely discharge the relieving fluid.

In recent years, methodologies originally used to determine relief valve areas have come under increasing scrutiny, particularly as operators review system capacity in relation to throughput increases or modifications to the existing design. Research into current design codes and practices for pressure vessel relief valves has shown that the commonly applied calculation methods can underestimate relief capacity. Newer more theoretically sound models are now being developed. Current Australian design codes are yet to incorporate these new models however in the interests of design integrity for both new and revised relief systems, these new models should now be considered.

In the Bayer process, safety relief valves on pressure vessels are potential sources of three-phase vapour-liquid-solid release. The flow occurring in the relief valve is therefore complex. In order to select an appropriate model a number of factors such as flow patterns, phase distribution, flow conditions and fluid properties must be considered with respect to the nature of the fluid. There are a wide variety of theoretical models which apply to two-phase flow. Each model has limitations and while a particular model may work well under certain conditions it may not be applicable in others.

2.0 REVIEW OF TWO PHASE FLOW METHODS

The models reviewed in this paper are, for simplification purposes, called “two phase” although the results shown in Figure 2 actually consider all three phases of solid, liquid and gas. Hereinafter reference to two phase calculations implies the solids and liquor form the first phase and the flashing vapour forms the second. This is especially significant to the Bayer and mineral processing industries where solid particles present in slurry mixtures must be accounted for in design calculations for relief valve sizing. The current design codes, in their treatment of two phase mixtures do not make specific reference to the solid phase. The solid phase should not be overlooked by the relief system designer.

2.1 API Method

According to continuity consideration for separated two-phase flow as outlined by Wallis (1969), we have the following general result for the overall mass flux.

$$G = \left[\frac{x}{m_g / A_g} + \frac{1 - x}{m_f / A_f} \right]^{-1} \tag{1}$$

and the total flow area:

$$A = A_g + A_f \tag{2}$$

The individual mass flux terms m_g / A_g and m_f / A_f in Equation (1) can be evaluated separately as follows:

The API formula for gas flow can be expressed as:

$$\frac{m_g}{A_g} = K_d K_b P \sqrt{\frac{M_W}{RTZ}} \left[k \left(\frac{2}{k+1} \right)^{(k+1)/(k-1)} \right]^{1/2} \tag{3}$$

The API for liquid flow can be written as:

$$\frac{m_f}{A_f} = K_d K_w K_v \sqrt{2 \rho_f \left(P - P_b \right)} \tag{4}$$

These above equations are the different forms of the Bernoulli equation.

2.2 Homogeneous Equilibrium Model (HEM)

DIERS recommends the homogeneous equilibrium model HEM as the appropriate flashing flow formulation (Leung, 1996). In this model, the flashing two phase flow mixture is treated much like a classical compressible gas while undergoing an adiabatic expansion with thermodynamic equilibrium in both phases. Among the many other flow models tested in the DIERS research program, the HEM yields conservative estimates of the flow capacity in a relief valve (DIERS Technology, 1992).

For flow in frictionless nozzles, The mass flux expression derived from the first law of thermodynamics is:

$$G^2 = \frac{\int_{p_o}^{p} -2v\,dp}{v^2} \qquad (5)$$

For two phase flow, the specific volume can be written as:

$$v = xv_g + (1-x)v_f \qquad (6)$$

At critical flow, the mass flow rate exhibits a maximum with respect to the throat pressure:

$$\left(\frac{dG_c}{dp}\right)_t = 0 \qquad (7)$$

The combination of Equations (5), (6) and (7) yields:

$$G_c^{\,2} = \left(\frac{\int_{p_o}^{p_t} -2\left[xv_g + (1-x)v_f\right]dp}{\left[xv_g + (1-x)v_f\right]_{@t}^2}\right)_{max} \qquad (8)$$

Methods for solving the above equation are via numerical integration (Perry, 1984), maximisation procedure (Wallis, 1969) or estimation method (Simpson, 1991). Detailed thermodynamic properties are required in these tedious calculations. Hence recently Leung (1996) proposed an analytical solution while employing an approximate equation of state (the ω-method) for the saturated two phase flashing region. The method gave good agreement with the experimental data.

2.3 Slip Flow Model

The analysis of critical flow rate in 2.2 above is based on the assumption that the flow is one-dimensional, homogeneous, in equilibrium and isentropic. In two phase flow, these assumptions may not necessarily be applicable. The co-existence of the two phases in various flow patterns makes homogeneous flow a near approximation only under the special condition of highly dispersed flow. When the two phases are separated, such as in annular or slug flow, the homogeneous assumption is no longer true. Also, when there is strong slip between phases, the homogeneous assumption may not be true even in the case of bubble or mist flow. When the flow is undergoing interfacial momentum transfer due to flow acceleration, the equilibrium and isentropic concepts are violated due to inconsistencies with the above assumptions. Therefore the slip flow model must be considered. Several of the available slip models can be cast in the general form (Butterworth, 1975; Grolmes and Coats, 1997):

$$S = C\left(\frac{1-x}{x}\right)^{p-1}\left(\frac{\rho_g}{\rho_f}\right)^{q-1}\left(\frac{\mu_f}{\mu_g}\right)^{r} \qquad (9)$$

The values of constants corresponding to the different models are as listed in Table 1.

Table 1: Slip Models

Correlation or Model	**C**	**p**	**q**	**r**
Homogenous model	1	1	1	0
Lockhart and Martinelli (1949)	0.28	0.64	0.36	0.07
Fauske (1962)	1	1	1/2	0
Thom (1964) correlation	1	1	0.89	0.18
Zivi (1964) model	1	1	0.67	0
Baroczy (1965) correlation	1	0.74	0.65	0.13
Moody (1965)	1	1	2/3	0
Wallis (1965) separate cylinder model	1	0.72	0.40	0.08

Based on the kinetic energy flux equation, Wallis (1969) determined the effective two phase specific volume in the presence of slip to be:

$$v = \left[xv_g + S(1-x)v_f\right]\left[x + \frac{1-x}{S^2}\right]^{1/2} \qquad (10)$$

If Equation (10) is used in combination with Equations (5) and (7), the expression for the critical mass flux is (Simpson, 1991):

$$G_c^{\,2} = \left(\frac{\int_{P_o}^{P_c} -2\left[xv_g + (1-x)v_f\right]dp}{\left\{\left[xv_g + (1-x)v_f\right]^2\left[xS^2 + 1 - x\right]\right\}_{@t}}\right)_{max} \qquad (11)$$

The required area of the relief valve is:

$$A_c = \frac{m}{K_d G_c} \qquad (12)$$

3.0 APPLICATION TO BAYER PROCESS

Pressure relief valves protect flash vessel integrity should a process condition occur which is outside the original mechanical design envelope. One such application of pressure vessel protection in the Bayer process is in the complex heat interchange system of Digestion. In this area of the Bayer plant the effective extraction of alumina is achieved by simultaneously heating process liquor/slurry as the digested slurry is depressurised in a flash train.

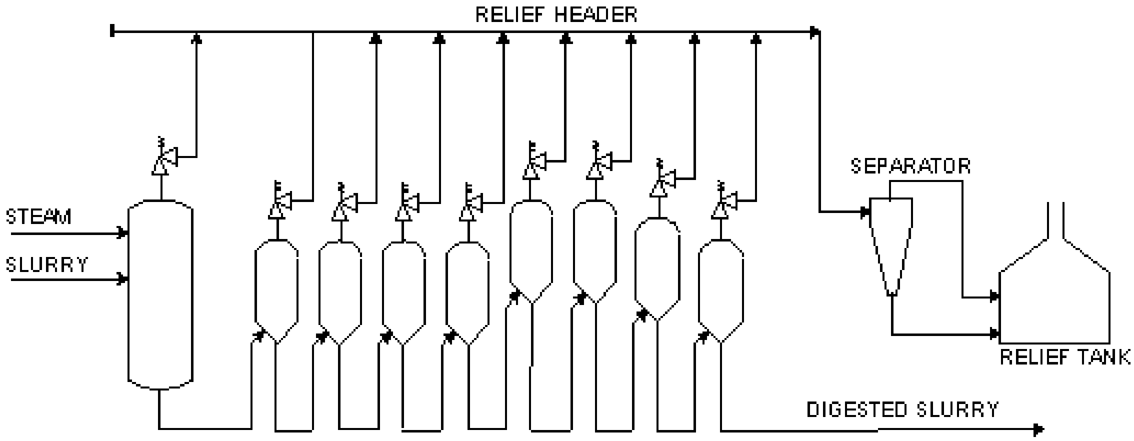


Figure 1: Typical Digestion Flash Train Relief System

Figure 1 above shows a typical relief system for a digestion flash train composed of eight flash vessels, (flash vapour piping and heat exchangers are not shown) which are connected to a relief header, separator and relief tank. Each flash vessel discharges into a sub header which then discharges into the main relief header. At the separator the slurry is further depressurised by cyclonic separation before finally discharging into the relief tank. Slurry can then be drained to a nearby sump from where it may be returned to the process.

3.1 Relief Capacity Scenarios

The relief valves in Digestion are sized to accommodate the maximum relief flows as determined by a rigorous dynamic simulation of different process scenarios. Ranging from “normal” operating conditions to more unusual process upsets, these scenarios involve hydraulic two phase modelling utilising two phase flashing slip models to determine the overall process effect. Many events could contribute to flash vessel levels and pressures approaching design limits, two of which are power failure and underflow blockages. These are discussed below:

3.1.1 Power Failure

Power failures can cause severe consequences for a number of areas in an alumina refinery. For Digestion, all pump controlled flows will cease as power is lost to pumps which supply and draw from the flash train, however uncontrolled flows such as those which cascade through the flash vessels, will be uninterrupted and continue to flow.

During the power failure the spent liquor or slurry flows, which are pumped through the shell and tube heat exchangers will stop and consequently the cold energy sink is no longer available to the flash vessels. The hot slurry in the flash vessels can no longer be cooled by countercurrent flashing heat exchange with the incoming spent liquor and the potential then exists for the hot slurry to flow onto downstream pressure vessels. Higher vapour pressures associated with the incoming hot slurry then introduces higher than "normal" pressure to vessels with lower design ratings. As a result the relief valves will then pop open and should be appropriately sized to handle these unusual circumstances.

3.1.1 Underflow Pipe Blockage

There may be several reasons for a blockage to occur downstream of a flash vessel including a scale blockage, control valve failure or more simply the incorrect closure of a process slurry valve. Dynamic process modelling of this upset scenario is based on the "worst case" condition whereby a complete blockage occurs. This methodology then predicts the minimum elapsed time period where levels and pressures in flash vessels may rise sufficiently to cause relief valves to open. At the moment the relief valve(s) open, the relief area required is a function of upstream vapour pressures driving the slurry through the valve inlet and back pressures generated in the relief header system downstream of the relief valve.

In this case flash vessels may gradually fill with slurry depending on the upstream vapour driving force and the relative elevations between the flash vessels. There may be sufficient pressure to entirely fill the vessel which then overflows into the heat exchanger. Depending on the time elapsed several upstream vessels could fill due to a downstream blockage.

3.2 System Back Pressure

The relief header, as shown in Figure 1 is connected to each individual flash vessel and collects the flashing slurry as it discharges from the relief valves. It is then safely transferred to a separator where the pressure is further reduced before flowing into the relief tank.

This relief header system must also be dynamically modelled to determine the system back pressures exerted on the relief valves as these pressures will affect the relief valve capacities. The relief header must be appropriately sized to transfer the discharging mixture safely away from the flash vessels.

3.3 Sizing Relief Valves

The dynamic hydraulic simulation of a relief system yields a data set of conditions for flow between flash vessels for which the relief valves must be sized. An example of a data set produced for a typical flash train as shown in Figure 1 is detailed in Table 1 below.

Table 1: Flash Train Relief Conditions

Vessel Data	**Digest**	**FV 1**	**FV 2**	**FV 3**	**FV 4**	**FV 5**	**FV 6**	**FV 7**	**FV 8**
Flowrate (m^3/h)	3211	3211	3136	3050	2949	2832	2694	2537	2360

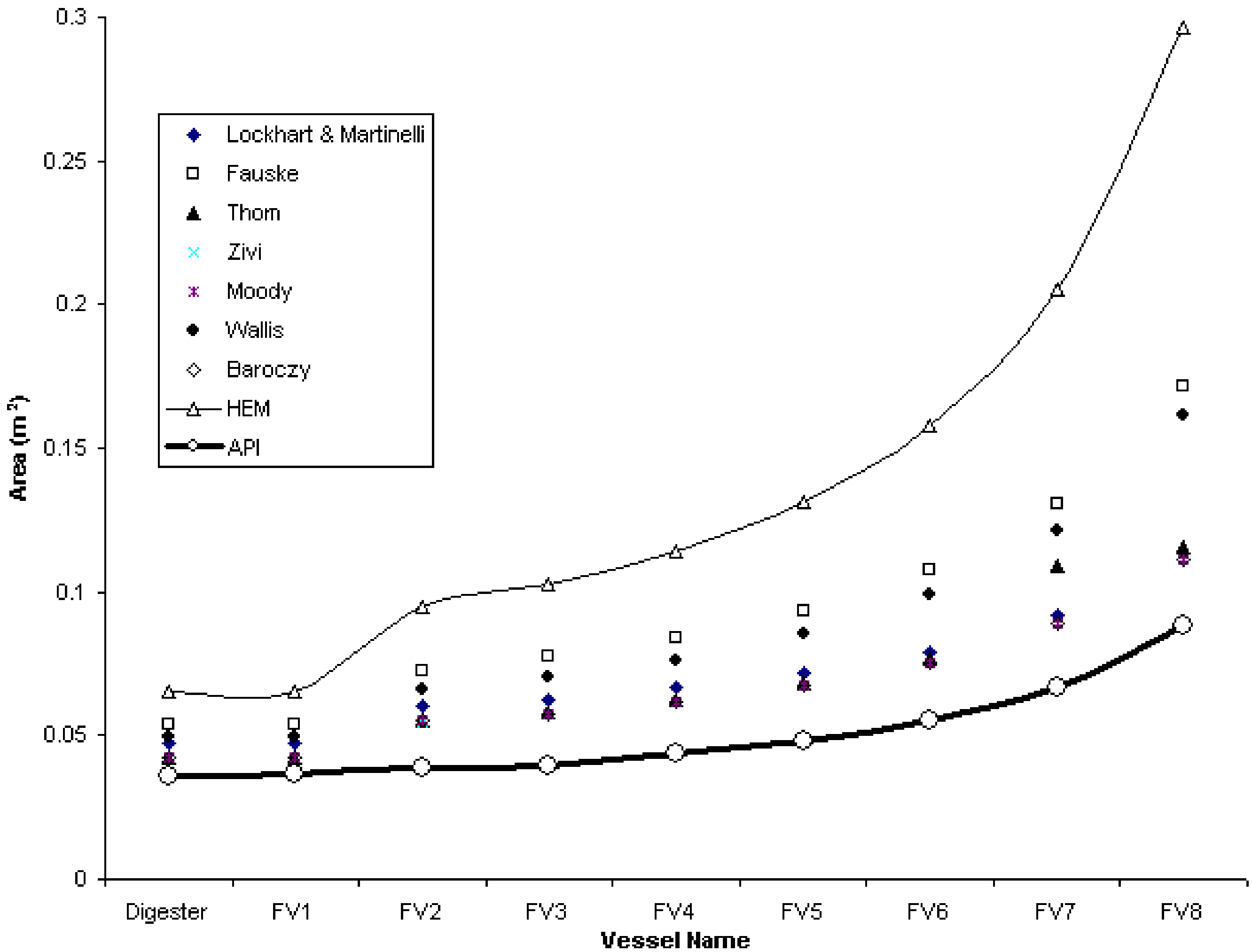
Solids Conc. (wt%)	3.3	3.3	3.4	3.4	3.5	3.6	3.7	3.8	4.0
Temperature (°C)	250	250	244	238	229	219	206	190	169
Pressure (kPa abs.)	5000	5000	2978	2630	2248	1834	1400	975	595
Boiling Point Elevation (°C)	10.5	10.5	10.7	10.8	10.8	10.8	10.8	10.8	10.8
Liquid SG @ Temp	1.122	1.122	1.133	1.146	1.162	1.181	1.205	1.236	1.273
Slurry SG @ Temp	1.149	1.149	1.160	1.173	1.190	1.210	1.236	1.267	1.307
Liquid Cp (kJ/kg°C)	3.439	3.439	3.460	3.402	3.434	3.477	3.536	3.618	3.735

From the results in Table 1, the required relief valve areas were calculated using the API method, HEM and the various empirical slip models. Required relief valve areas are shown in Figure 2.

4.0 DISCUSSION

From Figure 2, three distinct ranges for required relief area emerge. The greatest required relief areas are calculated using the homogeneous model (HEM), as recommended by DIERS. In comparison, the API method (currently adopted as a worldwide industry standard) results in the lowest required relief areas. In between these lie the various empirical slip model correlation results of Lockhart & Martinelli, Fauske, Thom, Zivi, Moody, Wallis and Baroczy.

The API method is a two step calculation procedure which allows for adiabatic two phase flashing to commence inside the relief valve, however it makes one overriding simplification which is not supported by the practical application of thermodynamic theory. This simplification involves the assumption that the liquid and vapour, which pass through the relief valve, do not interact. ie. they are a stratified or separated flow regime. The liquid phase passes through the valve based on a pressure gradient between the valve set pressure and relief header back pressure. The vapour phase, however, passes through the valve based on the pressure gradient between the valve inlet set pressure and the steam critical pressure. In effect, the API method allows two different pressures to exist at the same time inside the relief valve.



In contrast to the API method, the homogeneous equilibrium model (HEM) considers both liquid and vapour phases to pass through the valve as a mixture. This assumption is more realistic considering the relief valve will be a highly turbulent environment typified by flow regimes which are anything but stratified. Results of testwork conducted by DIERS and other researchers (Donaldson, 1993; Grolmes and Coates, 1997 and Leung et al., 1998) suggest the homogeneous model is conservative in estimating the required relief area and as Figure 2 demonstrates, this conservatism is accentuated at the low pressure end of the flash train.

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Figure 2: Flash Train Required Relief Areas

Between the results obtained for the API method and the homogeneous model are those generated using several published empirical Slip Models. These models have been developed by several researchers to account for inadequacies associated with two phase hydraulic modelling using the HEM method. The Slip Models account for the rapid expansion of the slurry mixture through the relief valve where the developing vapour phase has a lower density and is hence accelerated above the liquid phase velocity. The resultant temperature and energy differences between the phases (localised thermodynamic nonequilibrium) then induce interphase transfer of heat, mass and momentum. The various Slip Model correlations, as shown in Figure 2 were formulated by researchers under different test conditions and hence yield some variation in the required relief areas.

The Slip models show an approximation to the homogeneous model at high pressures since the specific volume of the developing vapour phase is low. Under these conditions, the Slip ratio, S is closer to unity and hence results obtained resemble near homogeneous flow. For lower pressures, the specific volume of the vapour fraction is much greater and can increase the Slip factor to values as high as 10. This explains the Slip Model’s ability to allow a greater mass flux or throughput to pass through a relief valve’s nozzle than predicted by the homogeneous model. ie. lesser relief area will be required.

Evidence of the Slip Model validity has been published by the DIERS Technology (1992) where experimentally observed relief valve discharge rates increased by a factor as great as three compared to rates predicted by the homogeneous equilibrium model.

While not specifically formulated for relief valve sizing, the Slip models allow the process designer to account for the inadequacies of the API method yet avoid potentially costly over design through implementation of the homogeneous (HEM) results. This translates as a potential saving in the cost of the relief valves that must be installed and hence will simplify the physical layout.

5.0 CONCLUSIONS AND RECOMMENDATIONS

The API method for sizing relief valves can lead to underpredicted required relief valve areas, however it should be understood that since actual relief valves purchased from manufacturers are always larger than the required area, a surplus of relief area is inevitably installed. Also, the relief conditions under which a relief system is designed are usually very conservative and would rarely occur, if at all, in plants with implemented procedures and suitably trained operators.

It is important to also note that the DIERS research involves the study of complex exothermic chemical reactions that do not apply to the flash train system of Digestion in a Bayer plant. The Digestion reactions are predominantly endothermic and do not contribute to the potential for a relief condition to be attained.

The adequacy of any safety relief system is subject to certain conditions, which are the principle basis for the original design. In time, with changes in plant personnel and operational policies and flow increases, the rationale associated with these conditions may be forgotten. The original principles adopted in a safety relief design must therefore be adhered to and any changes to operation and maintenance procedures must be thoroughly investigated for their effect on the integrity of the safety relief system.

For the process designer, increased knowledge in the field of two phase hydraulics, highlighted by test work and information published by groups such as DIERS, should be considered in any review of an existing relief system or in the design of a new one. Process hydraulic modelling conducted by Kaiser Engineers at several alumina refineries has enabled the development, verification and practical implementation of the published empirical Slip Flow models.

6.0 REFERENCES

American Petroleum Institute 520 (1993) Sizing, Selection and Installation of Pressure Relieving Devices in Refineries, Part I: Sizing and Selection, 6th Ed.

Butterworth D.A. (1975) A comparison of some void-fraction relationships for co-current gas-liquid flow, Int. J. Multiphase Flow, v1, pp 845-850.

Donaldson D.J. (1993) Digestion Relief System Design Factors, Light Metals, pp.105-110.

Emergency Relief System Design Using DIERS Technology (1992) Project Manual, AIChE.

Grolmes M.A. and Coates S.G. (1997) Progress Report on High Viscosity Two-Phase Flow Modelling, Part II, Presentation for 20th DIERS Users Group Meeting, Phoenix, AZ, April 1-15.

Leung J.C. (1996) Easily Size Relief Devices and Piping for Two-Phase Flow, Chemical Engineering Progress, December 1996, pp 28-50.

Leung J.C., Fitzsimons T.R., Kleine W.A. and Rudoll T.(1998) Phenolic Runaway Reaction: Pressure Relief and Containment, International Symposium on Runaway Reactions, Pressure Relief Design, and Effluent Handling, New Orlean, Louisiana, USA, March 11-13, pp.615-632.

Perry R.H. and Green D.(1984) Perry's Chemical Engineers' Handbook, 6th Ed.

Simpson L.L.(1991) Estimate Two-Phase Flow in Safety Devices, Chemical Engineering, August 1991, pp 98-102.

Wallis G.B. (1969) One-dimensional two-phase flow, New York, McGraw Hill.