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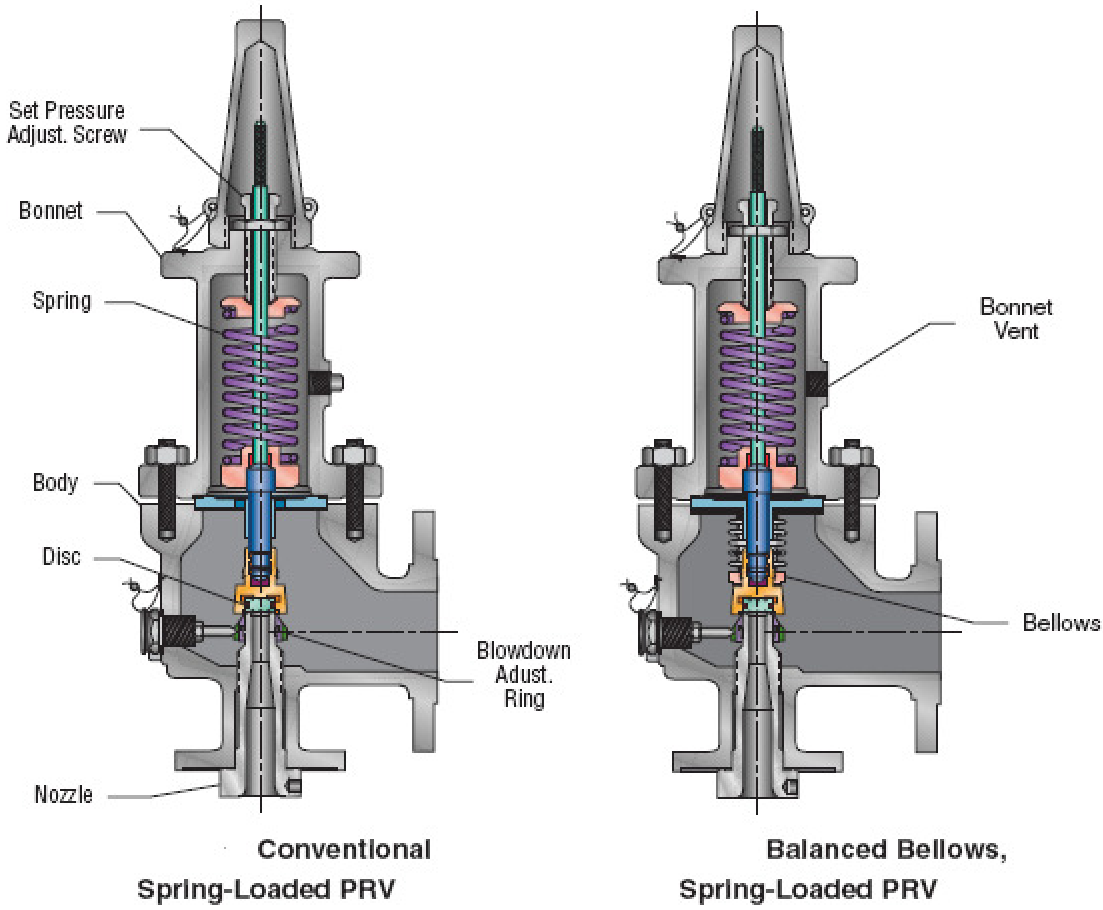

EMERGENCY RELIEF SYSTEM (ERS)

SIZING SOFTWARE

METHODS & PRACTICE

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1.0 Introduction

Emergency Relief System (ERS) design has been covered extensively by the Design Institute for Emergency Relief Systems (DIERS) [2] of the American Institute of Chemical Engineers (AIChemE), the American Petroleum Institute Standards (API) [3,4,5], the UK Health and Safety Executive (HSE) [1] and work by Leung [10], Fauske [7], Huff, Etchells, Wilday and many others, with research still ongoing.

The practising engineer is required to provide safe, practical and timely solutions to ERS design problems without having the opportunity to study the topic in depth. This technical note attempts to provide a comprehensive review of ERS fundamentals, definitions and design principles to allow the solution of cases encountered in normal practice. For more complex cases or when in doubt reference should be made to an appropriate authority or specialist in the field.

Chemstations have developed a menu driven Emergency Relief Sizing program supported by a component data base with comprehensive thermodynamic options. The design fundamentals presented in Sections 2, 3 and 4 have been used to validate the software by spreadsheet methods in Report MNL 044 [11]. The extensive features of the Chemstations sizing program cover most situations encountered in practice and are summarised below

- Batch or continuous process applications
- Reactive or non reactive systems
- Relief device: conventional relief valve, balanced relief valve, bursting disc or relief valve and bursting disc in combination
- Relief device: set pressure and allowable overpressure
- Design or rating for pressure vessel or atmospheric vessel
- Vessel model options: churn turbulent, bubbly or homogeneous equilibrium
- Single phase phase vent flow: liquid, gas / vapour or steam flow
- Prediction of two phase flow onset with relief sizing using Leung's method
- Two phase vent flow model options: homogeneous equilibrium (HEM), equilibrium rate (ERM) and Henry Fauske's homogeneous non-equilibrium (HNE)
- Heat model code options: API 520/521, NFPA 30, OSHA 1919.106 or API 2000
- Heat model options: specify heat rate, tempered runaway reaction or specify additional heat input
- Vessel geometry: cylindrical/spherical, vertical/horizontal
- Vessel end type: ellipsoidal, spherical, conical or flat
- Vessel parameters: diameter, tan-tan height and design pressure
- Vessel operating liquid level
- Vessel void volume calculated from geometry and liquid level or specify
- Height of bottom tan line from grade where fire can be sustained
- Options for fire fighting protection available and insulation factor
- Relief device stagnant or flowing input stream composition and condition set from CHEMCAD component database with appropriate thermodynamic options applied
- Relief device inlet and outlet diameter and length specified and pressure drop calculated and checked for design code compliance

2.0 Relief Sizing Fundamentals

2.1 Vessel Models(1, 2 p 8)

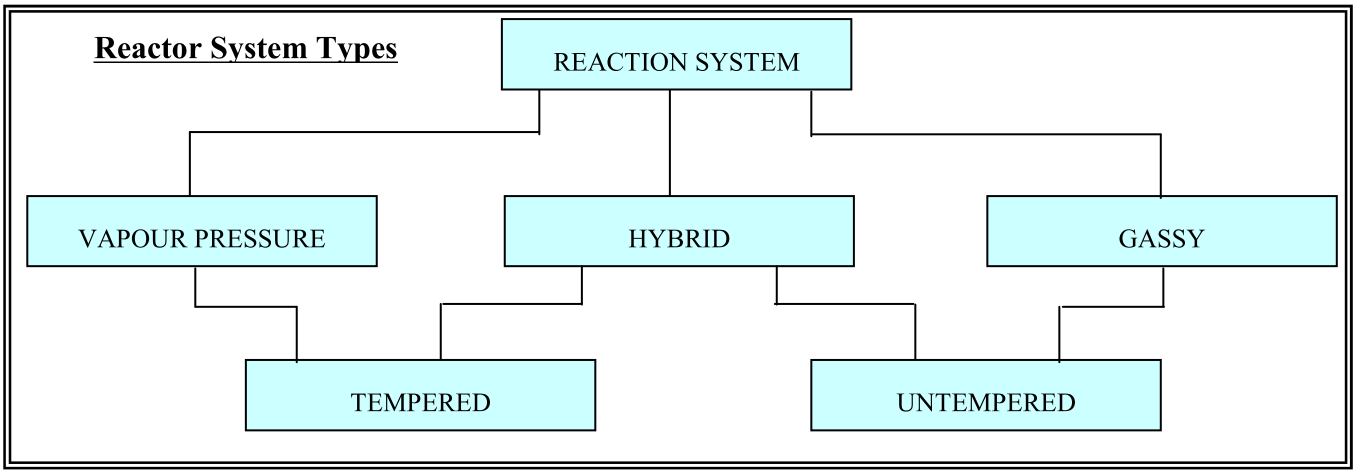
In relief system design involving chemical reactions there are three main types of process system:

- **Vapour pressure systems**
  The pressure generated by a runaway reaction is entirely due to the vapour pressure of the reacting mixture which rises as the temperature of the mixture increases during a thermal runaway. Vapour pressure systems are **tempered** in that the temperature and reaction rate are controlled during relief due to latent heat removal.
- **Gassy systems**
  The pressure generated by a runaway reaction is entirely due to a permanent gas which is evolved by the chemical reaction. Gassy systems are **untempered** in that pressure relief does not control the temperature or reaction rate.
- **Hybrid systems**
  The pressure is due to both the evolution of a permanent gas and increasing vapour pressure with increasing temperature.

Vessel flow models estimate the liquid swell (degree of vapour-liquid disengagement) as a function of vapour throughput. The vessel flow models are coupled with vent flow capacity models to determine the vapour mass fraction (onset of two phase-vapour-liquid venting) and the total mass flow rate entering the vent line. There are three flow models considered:

- **Homogeneous**
  Assumes zero vapour-liquid disengagement in the vessel where vapour bubble rise velocity relative to the liquid is zero. The vapour void fraction entering the vent line is the same as the vapour void fraction in the vessel. Typical of extremely viscous or foamy conditions or for when venting time is too short for an appreciable level rise.
- **Bubbly**
  Assumes uniform vapour generation throughout the liquid with limited vapour-liquid disengagement in the vessel. The liquid phase is continuous with discrete bubbles.
- **Churn-turbulent**
  Assumes uniform vapour generation throughout the liquid with considerable vapour-liquid disengagement in the vessel. The liquid phase is continuous with coalesced vapour regions of increased size relative to the vessel bubble model.

Two phase-vapour-liquid venting is typical of gassy(foamy) and hybrid reaction systems represented by the homogeneous and bubbly vessel models.

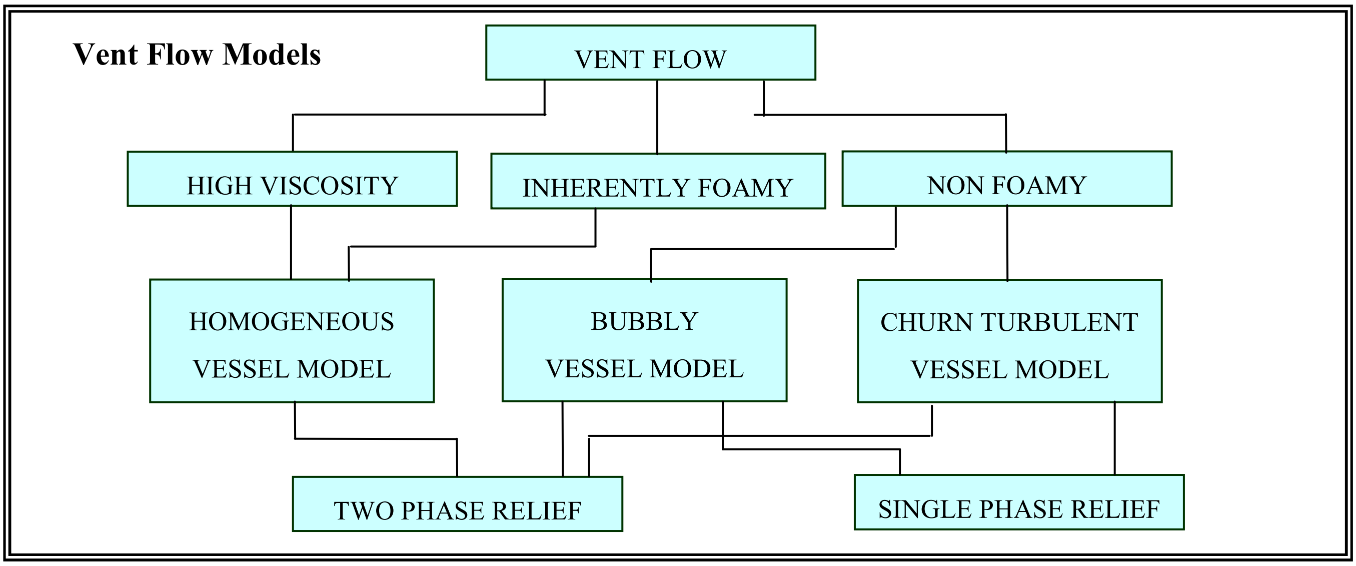


2.0 Relief Sizing Fundamentals

2.2 Vent Flow Models [1, 2]

Vent flow models for ERS sizing are based on single-phase or two-phase flow.
Single phase models considered are for Liquid, Vapor or Gas and Steam applications which are covered rigorously in the standards API 520 [3], API 521[4] and API 2000 [5].
ERS sizing involving chemical reactions invariably involves two-phase flow and three models are considered, namely:-

- **Homogeneous Equilibrium Model (HEM)** [1 p79]
  Applicable to both flashing and non-flashing two-phase flow, assumes uniform mixing of the phases across the pipe diameter with no phase slip(mechanical equilibrium), thermal equilibrium between the phases, and complete vapour/liquid equilibrium.
- **Henry-Fauske's Homogeneous Non-equilibrium(HNE) Model** [1 p165, 2 p64]
  Basis for family of models with varying extent of non-equilibrium behaviour, the rate of flashing at the choke point being a specified fraction (N) of the equilibrium value. Two-phase mixture treated as a single pseudo-component and vapour/liquid equilibrium is maintained in the reactor during the relief process. Assumes ideal vapour/liquid equilibrium, vapour phase is an ideal gas, frictionless nozzle, turbulent flow, no heat gain or loss and overpressure in the range 10-30% of absolute relief pressure.
- **Equilibrium Rate Model (ERM)** [2 p68]
  Special case of HNE family. Applicable to flashing two-phase flow only (vapour pressure systems), assumes saturated liquid entering the vent, no flashing (non-equilibrium flow) until the choke point and then flashing at equilibrium rate (N=1) at the choke point and the vapour phase is treated as an ideal gas.
  The following assumptions are considered conservative but may result in oversizing and an impractical vent system design requiring further rationalisation and optimisation.
- **For tempered systems**
  Two-phase rather than single phase vapour relief.
  Homogeneous vessel behaviour.
  Homogeneous equilibrium model (HEM) for relief system flow.
- **For untempered systems**
  Two phase relief at the point of maximum gas generation rate.
  Churn turbulent level swell vessel behaviour which minimises early loss of reactants.
  Homogeneous equilibrium model (HEM) for relief system flow.



3.0 Definitions and Assumptions

3.1 % Overpressure (Ref 3, p39, Tables 1, 2 and 3)

$$\textbf{\% Overpressure} = \left[\frac{P_{mG} \times CF - P_{RG}}{P_{RG}}\right] \times 100 \qquad (3.1)$$

where **CF** = contingency factor
= 1.1 for non-fire contingency (single valve case only)
= 1.21 for fire contingency (single and multiple valve cases)
P_{mG} = vessel maximum allowable working pressure (gauge) MAWP
P_{RG} = valve set pressure (gauge)

Note if P_{RG} is set at MAWP, then in accordance with ASME Code, Section VIII, Division 1, the accumulated pressure shall be limited to 110% of MAWP for the non- fire single valve operating contingency case and 21% for the fire contingency case. The set pressure of the device shall not exceed the MAWP. For multiple valve installations consult Reference 3. The following tables provide examples for the determination of the relieving pressure for a single valve case (operating contingencies).

Valve Set Pressure Less than MAWP - **CF = 1.1**	
Protected vessel MAWP, psig, P_{mG}	100.0
Maximum accumulated pressure, psig,	110.0
Valve set pressure, psig, P_{RG}	90.0
Allowable overpressure, psi	20.0
% Allowable overpressure	22.2
Relieving pressure, psia	124.7

Valve Set Pressure Equal to MAWP - **CF = 1.1**	
Protected vessel MAWP, psig, P_{mG}	100.0
Maximum accumulated pressure, psig	110.0
Valve set pressure, psig, P_{RG}	100.0
Allowable overpressure, psi	10.0
% Allowable overpressure	10.0
Relieving pressure, psia	124.7

Note if the overpressure is stated based on absolute pressure then this must be corrected to gauge pressure basis for entry into CHEMCAD as follows

$$\textbf{\% overpressure} = \left[\frac{P_m - P_R}{P_{RG}}\right] \times 100 \qquad (3.2):$$

The relief device set pressure or opening pressure is normally specified in gauge units and CHEMCAD requires the set pressure in absolute units. The maximum pressure is related to the set pressure or opening pressure as follows:

$$P_m = (P_R - P_{atm}) \times (1 + \%\,\text{overpressure}/100) + P_{atm} \qquad (3.3)$$

3.2 Friction Factors (6 - Memo 5/ 96)

A generalised Fanning friction factor is used for establishing line loss in the inlet to the relief device, with 0.005 being the default value. The following parameters are calculated at the relief device inlet:

Void fraction is given by $$\alpha_0 = \frac{X_0 v_0}{X_0 v_0 + (1 - X_0) v_f} \quad (3.4)$$

Two-phase(subscript $_{TP}$) viscosity $$\mu_{TP} = \alpha_0 \mu_g + (1 - \alpha_0) \mu_f \quad (3.5)$$

Reynolds Number $$Re = \frac{D\ G}{\mu_{TP}} \quad (3.6)$$

The following correlations are now used to calculate the Fanning friction factor based on the user specified roughness factor ε_R(ft) and all units in lb, ft, hr.

$$Re <= 2100 \qquad f = 16/Re \quad (3.7)$$

$$Re > 2000,\ f = 0.25\left[0.86859\ \ln\left(\frac{3.24\varepsilon_R}{D} + \left(\frac{7}{Re}\right)^{0.9}\right)\right]^{-2} \quad (3.8)$$

Material	Roughness Factor ε_R (ft)
Steel	0.00015
Cast Iron	0.00085
Plastic	0.000017
Glass	0.000005

3.3 Relief Device Discharge Coefficients (6 - Memo 12/ 96)

$$\text{Device factor} = C_d K_c \quad (3.9)$$

C_d = 0.975 for relief valve
= 0.953 for relief valve and rupture combination (6, memo 12/16/96)
= line factor for rupture disc alone with 0.625 as default
= 0.625*0.6 = 0.375 for rupture disc with vacuum support

K_c = 1 for relief valve alone
= 1 for rupture disc alone
= 0.9 for safety valve certified per the rules of ASME [3]
= 0.9 for relief valve and rupture disc combination

$$N = 4f(L/D) + 1.6 \quad (3.10)$$

Where 1.6 is resistance coefficient of rupture disc(1.1) + entrance(0.5)
And **f** is the Fanning friction factor, assume 0.005 as first trial.

3.4 Equivalent Lengths (6 - Memo 10/ 96)

3.4.1 Valves

Valve Type	L/D
Gate Valve	13
Ball Valve	3
Plug Valve	18
Swing Tilting Check Valve	100
Vertical Lift Check Valve	600
Angled Lift Check Valve	55

3.4.2 Fittings

Fitting Type	L/D
Std 90 degree elbow	30
Std 45 degree elbow	16
Long radius elbow	20
180 degree return	50
Std Tee (Run)	20
45 degree angled Tee (Run)	10
Std Tee (Branch)	60
45 degree angled Tee (Branch)	30

3.4.3 Pipe Parameters

Pipe Parameters	K
Entrance	0.78
Exit	1.0
Expansion	$(1 - \beta^2)^2$
Contraction	$0.5\,(1 - \beta^2)$

Where **β** is the ratio of the smaller pipe diameter to the larger pipe diameter

3.0 Definitions and Assumptions

3.5 Discharge Pipe Considerations [(9)]

All equivalent lengths can be referred to the same size by the resistance coefficient K Relationship [(9- p2-10 Equ 2-5)] as shown in Equation 3.11 below

$$K_a = K_b \left(\frac{d_a}{d_b} \right)^4$$

where 'a' defines K and d with reference to the pipe size to which all resistances are to be expressed, and where K is defined [(9- p2-8 Equ 2-4)] as shown in Equation 3.12 below

$$K = 4f \left(\frac{L}{D} \right)$$

The Fanning friction factor **f** is assumed identical in both pipes giving Equation 3.13 below

$$\left(\frac{L}{D} \right)_a = \left(\frac{L}{D} \right)_b \left(\frac{d_a}{d_b} \right)^4$$

3.6 Inlet / Outlet Pipe Sizing Rules [(6 - Memo 5/96)]

For compressible flow, including two phase vapour-liquid, the relief device inlet line pressure drop due to friction should not be more than 3% of the differential set pressure of the relief device protecting a vessel under the jurisdiction of ASME.
If the safety device is not oversized the nominal diameter of the inlet line should be at least equal to the nominal diameter of the inlet of the safety device.
If the safety device is a combination of rupture disc and relief valve, include 75D as the equivalent length of pipe contributed by the rupture disc for pressure drop calculation.

The following is a summary of the general rules:

Conventional Spring Loaded Safety Relief Valves

Differential set pressure	**ΔP_S = Set pressure – Constant Back Pressure**
Inlet pressure drop	**ΔP_I = 3% ΔP_S**
Outlet pressure drop	**ΔP_O = 10% ΔP_S**

Balanced Bellows Spring Loaded Safety Valve

ΔP_S = set pressure
ΔP_I = 3% ΔP_S
ΔP_O = 30% ΔP_S

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3.0 Definitions and Assumptions

3.7 Environment F Factor (3, 4 p17)

For ERS sizing based on fire exposure heat models the environment factor **F** is determined from the relevant standards whose key recommendations are shown below:

Environment	F Factor
Bare Vessel	1.0
Insulated Vessel	0.3
Sprinkler System	0.3 (Note 1)
Both Insulation and Sprinklers	0.15

Note 1: API 2000 does not recommend any allowance for sprinkler systems on bare vessels (facilities assumed failed) F = 1.0 is recommended.

For insulated vessels the following additional requirements are to be considered:

The insulation conductance in $Btu/hr.ft^2.°F$	F Factor
4	0.3
2	0.15
1	0.075
0.67	0.05
0.5	0.0376
0.4	0.03
0.33	0.026

3.8 API Valve Selection (6-memo)

Certified valves are open within 10% of set pressure for ASME Section VIII and 3% for ASME Section I with certified flows being 90% of the observed flow when operating at the previously stated overpressure.
The flow capacity of the valve body, at an acceptable back pressure, should be greater than the flow capacity of the nozzle. It is not unusual to have an outlet pressure rating lower than the inlet pressure rating of a relief valve.
Standard API valve type designations and nozzle sizes follow, where the first number indicates the inlet nominal size(in) and the number following the nozzle letter code is the outlet size (in).

Valve Type	Orifice Area(in^2)	Valve Type	Orifice Area(in^2)
1 x D x 2	>0.06 & <=0.110	3 x L x 4	2.853
1 x E x 2	0.196	4 x M x 6	3.6
1½ x F x 2	0.307	4 x N x 6	4.34
1½ x G x 2½	0.503	4 x P x 6	6.38
1½ x H x 3	0.785	6 x Q x 8	11.05
2 x J x 3	1.287	6 x R x 8	16.0
3 x K x 4	1.838	8 x T x 10	26.0

If vent area is <= 0.06 the relief valve is based on ¾ x 1 size body.

3.0 Definitions and Assumptions

3.9 Exposed Area Considerations

The heat transfer area to be used is determined by the agency governing the equipment operation and the criteria for selection is detailed in the following table.

EQUIPMENT	AGENCY GOVERNING THE EQUIPMENT OPERATION		
	NFPA-30 OSHA 1919.106	API-520[3] & API-521[4] Operating Pressure > 15 psig	API-2000[5] Operating Pressure ≤ 15 psig
1. Sphere	55% of total exposed area.	Area up to the maximum horizontal diameter or up to the height of 25 ft., whichever is greater.	55% of total exposed area. or surface area to a height of 30ft above grade whichever is greater.
2. Horizontal Tank	75% of total exposed area. If under 200 ft^2, use 100% of total exposed area.	Area equivalent to the average inventory level up to the height of 25ft.	75% of total exposed area or the surface area to a height of 30ft above grade, whichever is greater.
3. Vertical Tank	100% of total exposed area for the first 30 ft. Exclude bottom area if the bottom is flat and supported on ground.	Area equivalent to the average inventory level up to the height of 25 ft.	As in OSHA.
4. Process Vessel	-	Area equivalent to the average inventory level up to the height of 25 ft.	-
5. Fractionating Column	-	Area equivalent to liquid level in bottom, and reboiler if part of the column, plus liquid hold up from all trays up to a height of 25 ft.	-

For vertical tanks the area of the top head is excluded if the top head is more than 30ft above the ground for NFPA, OSHA and API 2000 codes.
However the top head area is included if there is two phase flow conditions. **(6, memo 4/97)**

3.0 Definitions and Assumptions

3.10 Methods for Evaluating Gas Density [8]

The gas density at relief device inlet conditions is calculated using the following general relationships. Non ideal behaviour is considered by determining the compressibility factor **Z** which can be estimated using the Compressibility Charts of Nelson and Obert [8,p376] provided the pseudocritical absolute temperature $\mathbf{T_c}$ and pressure $\mathbf{P_c}$ are known by evaluating the ordinates from the following

$$T_R = \frac{T_f}{T_c} \quad \text{and} \quad P_R = \frac{P_f}{P_c} \tag{3.14}$$

The gas density is calculated from the following

$$\rho_G = \frac{M_W P_f}{10.73 \;\; T_f Z} \text{lb / ft}^3 \quad \text{or} \quad = \frac{M_W}{22.415} \times \frac{P_f}{Z} \times \frac{273}{T_f} \text{kg / m}^3 \tag{3.15}$$

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4.0 Sizing Methods Basis

4.1 Relief System Sizing General

Figure 1 provides details of the general strategy used for ERS design and sizing. This procedure can be summarised as follows

- If relief from vessel decide if churn-turbulent, bubbly or homogeneous (2.1)
- Decide if system is vapour pressure, gassy or hybrid (2.1)
- Determine if single or two phase venting by checking onset conditions (2.2)
- Determine required relief rate **W** (mass flow/sec) based on worst case (4.2)
- Calculate proposed relief system capacity **G** (mass flow/unit area) (4.3)
- Calculated required relief device area **A** from **A=W/G**

The conditions for the onset of two phase flow (4.5.1) depend on the vessel geometry, the charge volume, the vessel flow model selected and the relief system vapour generation rate (mass flow/sec).
The phase split or quality of the venting stream is the key parameter, as defined by the mass fraction of gas/vapour **X** or the volume (void) fraction **α**, which depends on the flow regime in the venting vessel and are related by the following

$$X = \frac{\alpha\,\rho_g}{\alpha\,\rho_g + (1-\alpha)\rho_f} \qquad (4.1) \qquad\qquad \alpha = \frac{X/\rho_g}{(X/\rho_g) + ((1-X)/\rho_f)} \qquad (4.2)$$

If single phase flow has been used for initial sizing it should be confirmed that the onset conditions have not been established thus negating this assumption.

The required relief rate **W** is determined from the heat model selection (4.3) and whether we have single phase or two phase relief. The heat rate is determined from the relevant design codes for heat input $\mathbf{Q_H}$, the heat generated by the reaction $\mathbf{Q_T}$ or equipment failure modes.

The single phase gas flow capacity **G** is determined from the relevant sizing equation in the applicable standard and using relief device manufacturer's data for sizing coefficients. For bursting discs if friction is significant ($L_E > 40$) the Omega (4.5.3) method is used.

The two phase gas flow capacity **G** depends on the vent flow model selected and can be determined from the Omega method which is applicable to vapour pressure, gassy and hybrid systems.

4.0 Sizing Methods Basis

4.2 Relief System Sizing Flow

The required relief rate **W** can be determined from the following relationships depending on whether we have single or two phase vapour venting

For single phase vapour pressure systems, with no runaway reaction, we have

$$W = \frac{Q_H}{h_{fg}} \quad \text{(vapourisation rate)} \tag{4.3}$$

For runaway reactions the adiabatic temperature rise **(dT/dt)** and the peak rate of permanent gas evolution $\mathbf{Q_G}$ are determined experimentally. For single phase vapour venting the following apply [1, A6.2.2]

Tempered system $$W = \frac{dT}{dt}\frac{m\ C_f}{h_{fg}} + Q_G\rho_G \quad \text{(vapourisation rate + gas evolution rate)} \tag{4.4}$$

Untempered $$W = Q_G\rho_G + \frac{m\ C_f}{h_{fg}}\left(\frac{dT}{dt}\right)_{max} \quad \text{(gas evolution+corresponding vapour rate)} \tag{4.5}$$

For two phase vapour pressure systems Leung's method [1, A5.7] can be used where

$$W = \frac{m_R\, q_{av}}{\left[\left(\frac{V}{m_R}\frac{h_{fg}}{v_{fg}}\right)^{0.5} + (C_f\,\Delta T)^{0.5}\right]^2} \tag{4.6}$$

The average value of the difference between the vapour and liquid specific volumes $\mathbf{v_{fg}}$ at the set and relief pressures can be determined from

$$v_{fg} = \frac{1}{\rho_g} - \frac{1}{\rho_f} \tag{4.7}$$

At constant pressure there is no temperature rise and equation (4.6) simplifies to

$$W = \frac{m_R^2\, q_{av}\, v_{fg}}{V\ h_{fg}} = \frac{Q_T\, m_R\, v_{fg}}{V\ h_{fg}} \tag{4.8}$$

For the external heating case Leung's method equation (4.6) can be used with a modified value of heat release rate per unit mass **q** as follows [1 A5.7]

$$q_{mod} = q_{av} + \frac{2Q_H}{m_R} \tag{4.9}$$

Alternatively the modified form of Leung's equation [10] is used where we have

$$T_m - T_s = \frac{Q_T}{GA\ C_v}\left[\ln\left(\frac{m_0}{V}\frac{Q_T}{GA}\frac{v_{fg}}{h_{fg}}\right) - 1\right] + \frac{V\ h_{fg}}{m_0 C_v v_{fg}} \tag{4.10}$$

It is recommended that average values of the physical properties and heat input rate, between the vent opening pressure and accumulated pressure, be used. Liquid specific heat at constant pressure $\mathbf{C_f}$ can be used in place of specific heat at constant volume $\mathbf{C_v}$.
For single phase liquid venting due to thermal expansion refer to section 4.4.1

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4.0 Sizing Methods Basis

4.3 Heat Models

4.3.1 External Fire [2]

Heat input rates as a result of external fires have received extensive investigation by several organisations including API, NFPA and OSHA.

The operating pressure determines the applicable standard as follows:

API 520 [3] / API 521 [4]	operating pressure > 15 psig
API 2000 [5]	operating pressure ≤ 15 psig

In API 520/API 521 the heat input $\mathbf{Q_H}$ is determined from:-

With adequate drainage and fire fighting equipment	$\mathbf{Q_H = 21000\ FA^{0.82}}$	(4.11)
Without adequate drainage and fire fighting equipment	$\mathbf{Q_H = 34500\ FA^{0.82}}$	(4.12)

In API 2000 for low pressure storage tanks the heat input $\mathbf{Q_H}$ is determined from:-

In the range $0.4 \times 10^6 < \mathbf{Q_H} < 4.0 \times 10^6$ Btu/hr	$\mathbf{Q_H = 20000\ A}$	(4.13)
In the range $4.0 \times 10^6 < \mathbf{Q_H} < 9.95 \times 10^6$ Btu/hr	$\mathbf{Q_H = 199300\ A^{0.566}}$	(4.14)

Where,

$\mathbf{Q_H}$	= total heat absorption	Btu/hr
$\mathbf{A}$	= total wetted surface	ft^2
$\mathbf{F}$	= environmental factor	(API 521 Table 5)

The exposed area to be used is determined by the agency governing the equipment operation and the criteria for selection is detailed in previous section 3.9.

Single phase vapour venting design basis for non-foamy systems is usually considered the appropriate vent flow model for external fire cases unless the vessel is completely liquid filled or the entrainment velocity [7] $\mathbf{U_E}$ is exceeded where

$$\mathbf{U_E \approx 3.0 \left[\frac{\sigma\ g\ \rho_f}{\rho_g^2} \right]^{0.25}} \qquad (4.15)$$

4.3 Heat Models

4.3.2 Chemical Reaction [1] [2]

For relief device sizing involving runaway exothermic reactions experimental data is required from reaction screening techniques to establish the vessel model from reaction behaviour (vapour pressure, gassy or hybrid) and the vent flow model from the reaction type (tempered or untempered).

For most exothermic runaway reactions, the reaction rate which determines the heat release rate, increases exponentially with temperature. Hence for tempered systems the relief pressure should be set as low as practicable consistent with operational requirements for inerting procedures and pressure transfers.

For tempered runaway reactions (vapour pressure or hybrid) calorimetry can determine the tempering temperature at the relief pressure and if it can be established that the reaction will continue to temper until completion the Omega method can be used based on a homogeneous two phase system (vessel and vent flow model).

CHEMCAD process modelling and simulation software can be used to predict the tempering temperature at the relief pressure and temperature at the accumulated pressure.

Alternatively if the rate of temperature rise at the relief device set pressure $(\mathbf{dT/dt})_s$ and the maximum rate of temperature rise at the maximum design overpressure $(\mathbf{dT/dt})_m$ are known the average value of heat release rate per unit mass of reacting mixture $\mathbf{q_{av}}$ can be calculated from

$$\mathbf{q_{av} = 0.5\ C_p\left[\left(\frac{dT}{dt}\right)_s + \left(\frac{dT}{dt}\right)_m\right]} \qquad (4.16)$$

where $\mathbf{C_p}$ is the average liquid specific heat (kJ/kg K)

Leung's method for vapour pressure systems can now be used based on a homogeneous two phase system. The ERM vent model may also be used if friction can be ignored such as in the case for safety valves.

For untempered runaway reactions (gas evolution) the reaction rate depends primarily on the temperature and not on the pressure. The operation of the relief system cannot control the rate of runaway reaction but simply acts to remove material from the reactor. The vessel model with the most vapour/liquid disengagement ie churn turbulent, will normally provide the worst case. Relief device sizing is not based on a heat model, as such, but uses calorimetry to establish the peak gas evolution rate $\mathbf{Q_G}$ (m^3/s) by determining the peak rate of pressure rise $(\mathbf{dP/dt})_e$, the corresponding temperature $\mathbf{T_e}$ and the rate of temperature rise $(\mathbf{dT/dt})_e$ suitably adjusted for thermal inertia, where subscript **e** refers to experimental parameters and $\mathbf{T_c}$ the containment vessel temperature.

4.3 Heat Models

For closed tests the following relationship is used where $\mathbf{m_R}$ is the vessel reactant mass

$$Q_G = \left[\left(\frac{V}{P}\frac{dP}{dt}\right) - \left(\frac{V}{T}\frac{dT}{dt}\right)\right]_e \left[\frac{T_e}{T_c}\frac{m_R}{m_e}\right] \tag{4.17}$$

For open tests the above can be simplified to

$$Q_G = \left[\left(\frac{V}{P}\frac{dP}{dt}\right)\right]_e \left[\frac{T_e}{T_c}\frac{m_R}{m_e}\right] \tag{4.18}$$

If the containment vessel temperature $\mathbf{T_c}$ is not given it can be approximated by

$$T_c = 0.5\ (T_e + T_{amb}) \tag{4.19}$$

The required relief rate **W** (kg/s) can now be determined from

$$W = Q_G \frac{m_R}{V} \tag{4.20}$$

This relief rate is now used in the Heat Model option field "Specify Vent Flowrate"

In the absence of any calorimetric data the heat of reaction should be established from the components heats of formation using the following relationship

$$\sum \Delta H_{Products} = \sum \Delta H_{Reactants} \tag{4.21}$$

This heat of reaction is now used in the Heat Model option field "Specify Heat Rate".
It should be noted that a facility is available to add an additional Heat Input which could be as in the case of a services failure or external fire.

4.3.3 Abnormal Operation

Heat input rates can be user defined from a knowledge of the thermal characteristics of the reaction system and associated jacket services. This is particularly useful when considering operational and control system failure modes such as in the case of maximum heat input from vessel jacket services systems.

Control valve or pressure regulator failure is a common design case for determining the relief rate with the flowrate being based on the maximum differential pressure attainable across the valve seat.

Another common application is to protect equipment against overpressure due to thermal expansion of a trapped liquid.

4.0 Sizing Methods Basis

4.4 Sizing Methods for Single Phase Flow

4.4.1 Single Phase Liquid [3]

The sizing method presented is based on API 520 7th edition which should be referred to for more specific details and the relevant capacity factor correction graphs.

When capacity certification is required per ASME Section VIII, Div L

$$A = \frac{Q_{gpm}}{38\ C_d K_w K_c K_v}\sqrt{\frac{\rho_{SG}}{p_i - p_o}} \quad \text{in US customary units} \quad (4.22)$$

Where

A = required effective discharge area (in^2)
Q_{gpm} = flow rate (US gpm)
C_d = manufacturers rated coefficient of discharge, 0.65 as default
K_w = back pressure correction factor, 1.0 if atmospheric, otherwise refer [3]
K_c = combination correction factor, 1.0 if no rupture disc or 0.9 with rupture disc
K_v = viscosity correction factor from $\left(0.9935 + \frac{2.878}{Re^{0.5}} + \frac{342.75}{Re^{1.5}}\right)^{-1}$ (4.23)
ρ_{SG} = specific gravity of liquid at flowing temperature to water at standard conditions
p_i = upstream relieving pressure (psig)
p_o = back pressure (psig)

When capacity certification is not required the method assumes a C_d **of 0.62** and an overpressure of 25%.

$$A = \frac{Q_{gpm}}{38\ C_d K_w K_c K_v K_p}\sqrt{\frac{\rho_{SG}}{1.25p - p_o}} \quad (4.24)$$

Where
K_p = over pressure correction factor, 1.0 at 25% overpressure otherwise refer [3]

A common application is to protect equipment against overpressure due to thermal expansion of a trapped liquid. Expansion rates can be determined [4] from the following:

$$Q_{gpm} = \frac{B\ Q_H}{500\ G\ C} \quad (4.25)$$

Where
B = cubical expansion coefficient per ºF at expected temperature
Q_H= total heat transfer rate Btu/h
G = specific gravity reference water=1.00 at 60ºF
C = specific heat trapped fluid Btu/lbºF

If the blocked in liquid has a vapour pressure higher than the relief pressure then ERS design is based on vapour generation rate.

4.4 Sizing Methods for Single Phase Flow

4.4.2 Single Phase Gas or Vapour [3]

The sizing method presented is based on API 520 7th edition which should be referred to for more specific details and the relevant capacity factor correction graphs.

Sizing equation for critical flow in US customary units

$$A = \frac{W}{C\ K_d P_1 K_b K_c}\sqrt{\frac{T\ Z}{M_W}} \tag{4.26}$$

Where

$$C = 520\sqrt{k\left[\frac{2}{k+1}\right]^{\frac{k+1}{k-1}}} \tag{4.27}$$

$$k = \frac{C_p}{C_v} = \frac{M_W C_p}{M_W C_p - 1.99} \tag{4.28}$$

A = required effective discharge area (in^2)
W = flow rate (lb/hr)
$\mathbf{C_p}$ = specific heat (Btu/lb °F) If **k** cannot be estimated use **C** = 315 or refer Table 8 [3]
$\mathbf{K_d}$ = effective coefficient of discharge. For preliminary sizing use the following
= 0.975 for pressure relief device with or without rupture disc
= 0.62 for rupture disc
$\mathbf{P_1}$ = upstream relieving pressure (psia)
= Set pressure(psig) x (1+ accumulation) + atmospheric pressure
Accumulation = 0.1 for unfired emergency relief
= 0.21 for fire condition
= 0.33 for piping
$\mathbf{K_b}$ = back pressure correction factor from manufacturers data for balanced bellows valves only. Conventional and pilot operated valves $\mathbf{K_b}$ = 1.0.

Refer Figure 35 [3] or estimate using

$$K_b = \frac{735\ F\ \sqrt{1-r}}{C} \tag{4.29}$$

$\mathbf{K_c}$ = combination correction factor,
= 1.0 if no rupture disc
= 0.9 when rupture disc installed in combination with pressure relief valve
T = relieving temperature of inlet gas or vapour (°R = °F + 460)
Z = compressibility factor of gas or vapour at inlet relieving conditions
$\mathbf{M_W}$ = molecular weight of gas or vapour at inlet relieving conditions

Sizing equation for subcritical flow in US customary units with above nomenclature

$$A = \frac{W}{735\ F\ K_d K_c}\sqrt{\frac{T\ Z}{M_W P_1 (P_1 - P_2)}} \tag{4.30}$$

$$F = \sqrt{\left(\frac{k}{k-1}\right)(r)^{2/k}\left[\frac{1-r^{(k-1)/k}}{1-r}\right]} \qquad r = \frac{P_2}{P_1} \tag{4.31}$$

$\mathbf{P_2}$ = back pressure (psia)
Balanced bellows relief valves operating in the subcritical region may be sized using the critical flow formulae.

4.4 Sizing Methods for Single Phase Flow

4.4.3 Dry Steam [3]

The sizing method presented is based on API 520 7th edition which should be referred to for more specific details and the relevant capacity factor correction graphs.

Sizing equation for steam service at critical flow in US customary units

$$\mathbf{A} = \frac{\mathbf{W}}{\mathbf{51.5\ P_1 K_d K_b K_c K_N K_{SH}}} \qquad (4.32)$$

Where

$\mathbf{K_N}$ = correction factor for Napier equation
= 1.0 where $\mathbf{P_1} \leq 1500$ psia
$= \frac{\mathbf{0.1906 \times P_1 - 1000}}{\mathbf{0.2292 \times P_1 - 1061}}$ where $\mathbf{P_1} > 1500$ psia and ≤ 3200 psia

$\mathbf{K_{SH}}$ = superheat steam correction factor from Table 9 [3]
= 1.0 for saturated steam at any pressure

4.5 Sizing Methods for Two Phase Vapour Liquid Flow (2, p 27)

4.5.1 Method for Predicting Onset of Two Phase Vapour Liquid Flow

Determine the vapour capacity **F** (lb/h) of the relief device at the desired set pressure. The superficial vapour velocity $j_{g\infty}$ is then calculated using the following relationship:

$$j_{g\infty} = \frac{F}{3600\ \rho_g A_R} \tag{4.33}$$

where ρ_g is the vapour density (lb/ft^3) and A_R is the vessel cross sectional area (ft^2)

The bubble rise velocity U_∞ (ft/s) is now calculated using the following relationship:

$$U_\infty = \frac{K\left[32.174\left(2.2046E-03\right)\sigma\left(\rho_f-\rho_g\right)\right]^{0.25}}{\rho_f^{0.5}} \tag{4.34}$$

Where **K** is **1.53** for churn turbulent fluid and **1.18** for bubbly fluid
The dimensionless superficial gas/vapour velocity ψ due to flow is given by:

$$\psi_F = \frac{j_{gx}}{U_\infty} \tag{4.35}$$

The dimensionless superficial vapour velocity at which two phase-vapour-liquid flow commences is now determined from the following:

Churn turbulent fluid

$$\psi = \frac{2\alpha}{1 - C_0\alpha} \tag{4.36}$$

Bubbly fluid

$$\psi = \frac{\alpha\left(1-\alpha\right)^2}{\left(1-\alpha^{-3}\right)\left(1-C_0\alpha\right)} \tag{4.37}$$

Where C_0 the correlating parameter is given by:

Vessel Fluid Model	C_0 Best Estimate	C_0 Conservative
Churn-turbulent	1.5	1.0
Bubbly	1.2	1.0
Homogeneous		1.0

Note that

$$\alpha = \frac{V - V_f}{V} \tag{4.38}$$

Where **V** is the total reactor volume and $\mathbf{V_f}$ is the volume of liquid in the reactor.

In summary we can now predict the onset of two phase-vapour-liquid flow from:

Two phase-vapour-liquid venting is predicted if $\psi_F >= \psi$

All vapour venting is predicted if $\psi_F < \psi$

4.5 Sizing Methods for Two Phase Flow

4.5.2 Coupling Equation and Derivations (2, 6 - Memo 1/96)

The relief of a two phase-vapour-liquid fluid is to be achieved in accordance with the overpressure requirements of the relevant design code or standard.
When the onset of two phase vapour liquid flow is predicted the coupling equation (2, p5) is solved to determine the flowing quality of vapour at the vent inlet.
The coupling equation is a vapour material balance at the vent inlet as follows:

Total weight rate of vapour entering the vent =
Vapour disengaging the liquid surface + Vapour in the aerated liquid passing to the vent

$$X_0 A\ G = j'_{g\infty}\rho_{g,s} A_R + X_m\left(A\ G - j'_{g\infty}\rho_{g,s} A_R\right) \quad (4.39)$$

Where:

$j'_{g\infty}$ = superficial vapour velocity at liquid surface necessary to just swell the liquid to the top of the vessel (ft/s)

$\rho_{g,s}$ = vapour density at relief pressure and corresponding saturation temperature (lb/ft^3)

X_m = stagnation quality at liquid surface, where $X_m = \dfrac{\alpha_m \rho_{g,s}}{\alpha_m \rho_{g,s} + (1-\alpha_m)\rho_f}$ (4.40)

α_m = void fraction at the liquid surface

For Churn-Turbulent Vessel Model the condition at the relief device inlet, denoted by subscript 1, is used where

$$\alpha_{21} = \frac{2\ a}{(1 + C_0 a)} \quad (4.41) \qquad v_{21} = v_{f0} + X_1(v_{fg}) \quad (4.42) \qquad v_{fG} = \frac{1}{\rho_g} - \frac{1}{\rho_f} \quad (4.43)$$

a = average void fraction in the swelled liquid for Churn-Turbulent Vessel Model
Co = Correlation parameter

For Homogeneous Vessel Model it can be shown

$$\alpha_0 = \frac{X_0 v_{g0}}{v_{20}} \quad (4.44) \qquad X_0 = \frac{\alpha_0 \rho_{g0}}{\alpha_0 \rho_{g0} + (1-\alpha_0)\rho_{f0}} = \frac{\alpha_0 \rho_{g0}}{\rho_{20}} \quad (4.45)$$

$$v_{20} = X_0 v_{g0} + (1 - X_0) v_{f0} = \frac{V}{m_0} = \frac{1}{\rho_{20}} \quad (4.46)$$

Alternatively in the terms of initial vessel conditions with vapour volume $\mathbf{V_g}$, vessel gross volume **V** and reactant mass $\mathbf{m_0}$ we have

$$\alpha_0 = \frac{V_g}{V} = 1 - \frac{V_f}{V} = \frac{V}{m_0} \quad \text{the average vessel void fraction} \quad (4.47)$$

The vent mass flux **G** (lb/ft^{2}s) can be determined from frictionless flow:

$$G = \left[2 g_c \rho_{21}(P - P_B)144\right]^{1/2} \text{ where vent two phase density } \rho_{21} = \frac{1}{X_0/\rho_{g,s} + (1-X_0)/\rho_f} \quad (4.48)$$

Two phase flow vent rate (ft^3/s) $\qquad Q_{20} = \dfrac{G\ A}{\rho_{20}}$ (4.49)

Gas generation rate (ft^3/s) $\qquad Q_G = j_{g\infty} A_R$ (4.50)

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4.5 Sizing Methods for Two Phase Flow

4.5.3 Omega Method for Two Phase Vapour Liquid Flow (2, 6 - Memo 1/ 96)

Omega ω is a correlating parameter in the following equation of state

$$\left(\frac{v}{v_0}-1\right)=\omega\left(\frac{p_0}{p}-1\right) \quad (4.51)$$

Where $\omega < 1$ gassy two phase, $\omega = 1$ gas only single phase and $\omega > 1$ flashing two phase

The method is applicable when the following conditions are satisfied

- Homogeneous two phase turbulent flow
- If two phase flashing mixture, vapour liquid equilibrium is maintained and expansion process is isenthalpic
- If two phase non flashing mixture thermal equilibrium is maintained between phases and expansion process is isentropic
- Friction factor constant
- Conditions of applicability $T / T_{Tc} < 0.9$ and $P / P_{Tc} < 0.5$

Flowing quality X_0 (vapour wt flow/total wt flow) is determined by solving the coupling equation (4.4.1) where subscript 0 refers to stagnation conditions in the upstream vessel

The Omega method is used to calculate the dimensionless mass flux G_c^* and the critical pressure ratio η_c using Figure A8.2 (1) or by curve fitting

For vapour pressure and hybrid systems

$$\omega = \frac{\alpha_0}{k} + (1-\alpha_0)\rho_{f0} C_f T_0 P_{v0}\left(\frac{v_{fg0}}{h_{fg0}}\right)^2 \quad (4.52)$$

Vapour volume $v_{g0} = \dfrac{X_0}{\rho_{g0}}$ (4.53) Liquid volume $v_{f0} = \dfrac{(1-X_0)}{\rho_{f0}}$ (4.54)

$\alpha_0 = \dfrac{v_{g0}}{v_{g0}+v_{f0}}$ (4.55) $v_{fg0} = \dfrac{1}{\rho_{g0}} - \dfrac{1}{\rho_{f0}}$ (4.56) $\alpha_0 = v_{f0} + X_0 v_{20}$ (4.57)

Note that the term based on stagnation liquid conditions can also be represented by inlet two phase density or specific volume as follows

$$(1-\alpha_0)\rho_{f0} = \rho_{21} = \frac{1}{v_{21}} \quad (4.58)$$

For gassy systems the above reduces to $\omega = \dfrac{\alpha_0}{k}$ (4.59)

$$\eta_c = 0.6055 + 0.1356 \times \log \omega - 0.0131 (\ln \omega)^2 \quad (4.60)$$

If $\omega < 4$ then G_c^* is calculated from $G_c^* = \dfrac{0.66}{\omega^{0.39}}$ (4.61)

If $\omega >= 4$ then G_c^* is calculated from $G_c^* = \dfrac{\eta_c}{\sqrt{\omega}}$ (4.62)

Calculate choked nozzle flow G_c from $G_c = G_c^*\sqrt{\dfrac{P_0}{v_0}}$ (4.63) Relief area $A=W/G$ (4.64)

4.5 Sizing Methods for Two Phase Flow

4.5.4 Henry-Fauske's Homogeneous Non-equilibrium HNE Model [1 A53, 2 p64]

This method is applicable to two-phase mixtures treated as a single pseudo-component with vapour/liquid equilibrium being maintained in the reactor during the relief process. It assumes ideal vapour/liquid equilibrium, vapour phase is an ideal gas, frictionless nozzle, turbulent flow, no heat gain or loss and overpressure in the range 10-30% of absolute relief pressure.

No vapour disengagement ie system inherently foamy

HNE model uses the following relationship for relief system sizing where all pseudo-component properties are at the relief conditions.

$$A = \frac{1}{2} \frac{m_R \left(dT/dt\right)_R}{F\left(P_m - P_R\right)} \sqrt{\frac{C_{fR}}{T_R}} \tag{4.65}$$

where C_{fR} is liquid specific heat at the relief pressure and $(P_m - P_R)$ is the maximum accumulated pressure and relief pressure difference.

The frictional correction factor **F** values are shown below as a function of vent line equivalent length

L_E/D	F
0	1
50	0.87
100	0.78
200	0.68
400	0.57
600	0.5

Vapour / liquid disengagement

The area calculated from the equation above is multiplied by the disengagement factor

$$\frac{\alpha_D - \alpha_R}{1 - \alpha} \tag{4.66}$$

where α_R is void fraction in reactor at relief pressure and α_D is void fraction at disengagement. Note application of this correction will result in a reduced relief area.

4.5.5 Equilibrium Rate ERM Model [1 9.4.2 , 2 p68]

Special case of HNE critical flow model obtained by setting parameter N equal to unity. The model is only applicable to flashing two phase flow with vapour phase treated as an ideal gas ie tempered vapour pressure systems. It assumes saturated liquid entering the vent, no flashing (non-equilibrium flow) until the choke point and then flashing at equilibrium rate (N=1) at the choke point.
The simplified ERM correlation is given by

$$G = \left(\frac{dP}{dT}\right)_0 \left(\frac{T_0}{C_{f0}}\right)^{0.5} \tag{4.67}$$

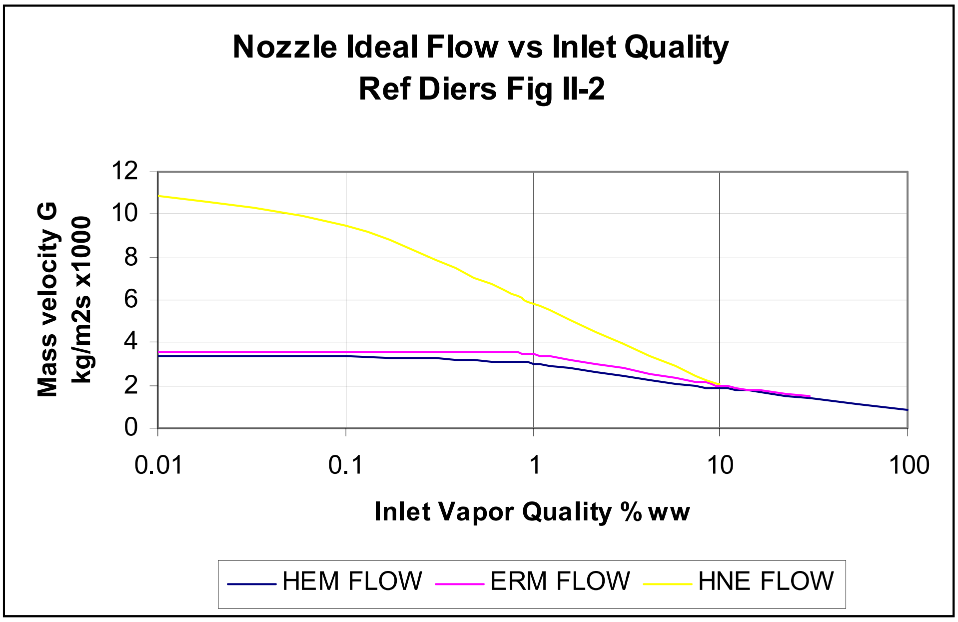
CHEMCAD component data base can be used to evaluate the slope of the vapour pressure vs temperature curve at stagnation condition.

For single component systems or treating as a single pseudo-component mixture with a narrow boiling range and using the isenthalpic rate of vaporization at the critical condition we obtain the simplified ERM non-equilibrium correlation which uses stagnation conditions subscript “0” in the reactor.

$$G = \frac{h_{fg0}}{v_{fg0}\left(C_{f0}T_0\right)^{0.5}} \tag{4.68}$$

The method finds its main application in the sizing of relief valves on vapour pressure systems with an appropriate discharge coefficient applied.

The mass velocity in an ideal nozzle will depend on the model selected and the flowing quality as shown by Diers.[p 64]



5.0 Nomenclature

Definition of symbols used unless defined otherwise in main text for specific equations.
SI units are the preferred but consistent units should be used as appropriate.
Some correlation constants require units to be used as defined in main text.

A	cross-sectional area of relief device (m^2)
A_R	cross-sectional area of vessel (m^2)
B	cubical expansion coefficient
C_d	relief device discharge coefficient
C_f	liquid specific heat capacity (J/kg°K)
C_0	correlating parameter two phase flow
C_p	average liquid specific heat at constant pressure (J/kg°K)
C_v	liquid specific heat at constant volume (J/kg°K)
D	relief system (or pipe) diameter (m)
F	environmental factor
f	Fanning friction factor
G	two phase mass flowrate per unit flow area (kg/m^2s)
G_c	choked nozzle flow (kg/m^2s)
g_c	gravitational constant consistent units
h_{fg}	latent heat of vaporisation (J/kg)
$j_{g\infty}$	superficial vapour velocity consistent units
K_c	relief device combination factor
K_N	correction factor for Napier equation
K_{SH}	correction factor for superheat steam
K_p	overpressure correction factor
K_v	viscosity correction factor
K_w	back pressure correction factor
L	line length (m)
M_W	vapour or gas molecular weight
m_e	mass in experimental calorimeter (kg)
m_0	initial mass in the reactor (kg)
m_R	mass in reactor at relief pressure (kg)
P_{atm}	atmospheric pressure (N/m^2 abs)
P_f	flowing pressure (N/m^2 abs)
P_m	maximum accumulated pressure (N/m^2 abs)
P_{mG}	maximum accumulated pressure (N/m^2 gauge)
P_R	relief pressure (N/m^2 abs)
P_{RG}	relief pressure (N/m^2 gauge)
Q_H	relief system external heat input (J/s)
Q_G	peak rate of permanent gas evolution (m^3/s)
Q_T	total heat release rate of reacting mixture (J/s)
q_{av}	average heat release rate per unit mass reacting mixture (J/kg s)
q_{mod}	average heat release rate per unit mass reacting mixture modified for external heating (J/kg s)
Re	Reynolds number
T_{amb}	ambient temperature (°K)
T_c	containment vessel temperature (°K)

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5.0 Nomenclature

$\mathbf{T_f}$	flowing temperature (°K)
$\mathbf{U_E}$	entrainment velocity (m/s)
$\mathbf{V}$	volume of reactor (m^3)
$\mathbf{V_f}$	volume of liquid in reactor (m^3)
$\mathbf{V_g}$	volume of vapour in reactor (m^3)
$\mathbf{v_0}$	specific volume at stagnation conditions at inlet to relief system (m^3/kg)
$\mathbf{v_f}$	liquid specific volume (m^3/kg)
$\mathbf{v_{fg}}$	difference between vapour and liquid specific volumes (m^3/kg)
$\mathbf{W}$	mass flowrate (kg/s)
$\mathbf{X}$	vapour or gas mass fraction
$\mathbf{X_m}$	stagnation quality at liquid surface
$\mathbf{X_0}$	mass fraction of gas/vapour in a two phase mixture at stagnation conditions
$\mathbf{Z}$	vapour or gas compressibility
α	vapour or gas volume(void) fraction
α_m	void fraction at liquid surface
α_0	initial void fraction in vessel
β	ratio of small pipe diameter to large pipe diameter
ε_R	roughness factor (ft)
ρ_G	gas density (kg/m^3)
ρ_g	gas or vapour density (kg/m^3)
ρ_f	liquid density (kg/m^3)
ρ_{SG}	specific gravity of flowing fluid
ρ_{21}	two phase flowing density at relief device inlet (kg/m^3)
μ_{TP}	two phase dynamic viscosity (Ns/m^2)
μ_g	gas/vapour dynamic viscosity (Ns/m^2)
μ_f	liquid dynamic viscosity (Ns/m^2)
σ	liquid surface tension (N/m)
ψ	superficial vapour velocity (dimensionless)
ω	correlating parameter
ΔH	heat of formation in consistent units
ΔP_S	relief device set pressure minus back pressure (N/m^2 abs)
ΔP_I	inlet pipe pressure drop (N/m^2 abs)
ΔP_O	outlet pipe pressure drop (N/m^2 abs)
(dT/dt)	adiabatic temperature rise (°K/s)
$(dP/dt)_e$	peak rate of pressure rise in experimental calorimeter (N/m^2s)
$(dT/dt)_s$	temperature rise at set pressure (°K/s)
$(dT/dt)_m$	temperature rise at maximum design overpressure (°K/s)

Subscripts

This is not a comprehensive list but is included for clarification only

0	condition at initial or stagnation
1	condition at inlet
2	two phase condition
21	two phase condition at inlet
R	relief condition
f	liquid
g	gas or vapour

Appendix I
CHEMCAD Relief Sizing Tool Design Note
Two Phase Flow Relief Sizing by Ming der Lu

Method Two Phase	**Results**			
Parameter	Mass Flux		**Flowrates**	Area P_{acc}
Units	Kg/m^2s	Ft^3/s	Kg/h	M^2
Omega	5485.3	0.80	50955	0.00257
Leung 1 (2*Q Ext Heat)	5897.2	0.95	60060	0.0028
Leung Ext Heat	5897.2	0.78	50955	0.0024
Superficial flow for vent	5907.6	2.44	154957	0.00728
CC-5 Relief (HEM)	7603	1.18	74995	0.00274
CC-5 Relief (ERM)	11678	1.81	115195	0.00274

Notes:
HEM two phase density 38.836 lb/ft^3 (all methods agree for HEM case).
Flowrate based on superficial velocity and vessel csa is 2.44 ft^3/s
DIERS [3, p400] recommend a derating factor for safety valve discharge coefficients using the ASME Efficiency Factor 0.9, so factor 0.975 for safety valve becomes 0.8775.

Leung and Omega methods propose use of average vapour density calculated from set pressure and accumulated pressures for tempered systems. For the above Leung external heat case the area increases for 0.0024 to 0.0026. For a 21% accumulation; an increase of 8.3%.

Two Phase tempered runaway reaction sizing using Leung's equation [12, p1625, Eq 20 & 11, p371, Eq 5-64]
For gassy systems G should be evaluated at maximum accumulated pressure which minimizes the relief area.

$$W = \frac{m\ Q}{\left[\left(\frac{V}{m}\frac{h_{fg}}{v_{fg}}\right)^{0.5} + \left(C_f \Delta T\right)^{0.5}\right]^2}$$

which reduces to $W = \frac{m\ Q_T\, v_{fg}}{V\ h_{fg}}$ at const P

The sizing flowrates, except for the superficial velocity case above, have been determined using the following modification to Leungs method for the external heating case *using average values of physical properties between vent opening and maximum pressure allowable. G evaluated at accumulated pressure or maximum disc burst pressure including tolerances. Q_T is evaluated using half the reactant mass m_0.*

$$T_m - T_s = \frac{Q_T}{GA\ C_v}\left[\ln\left(\frac{m_0}{V}\frac{Q_T}{GA}\frac{v_{fg}}{h_{fg}}\right) - 1\right] + \frac{V\ h_{fg}}{m_0 C_v v_{fg}}$$

Diers VI-A5-1 Flashing choked flow-modified ERM equation (metric form)
For low quality $X_0 < 0.02$ or $\alpha_0 < 0.5$aand negligible friction ie relief via safety valve directly to atmosphere or bursting disc with short discharge pipe

$$G_{TP} = (0.9)\frac{h_{fg}}{v_{fg}}\left[\frac{1}{C_P T}\right]^{0.5} 0.8775$$

Diers VI-A5-2 Flashing choked flow-Generalised HEQ correlation (metric form)
For any quality, where G^* is evaluated using the ω method

$$G_{TP} = \left(\frac{P_0}{v_{20}}\right)^{0.5}\left(G^*\right)\text{Factor}$$

Compare to Diers II-38 $G_{c,homogeneous} = \eta_c\left(\frac{P_0}{a\ v_0}\right)^{0.5}$

Two phase non-tempered reactions are sized for VSP using DIERS [3, page 431] Leung's Analytical Method III (Hybrid/Non-tempered or Gassy Non-tempered Reaction. For RSST using Vent Sizing for Gassy and Hybrid Systems by Leung, equation 8 page 301.

Appendix II
CHEMCAD Relief Sizing Tool User Notes

1. Specify the inlet stream composition at a nominal flow, say 1.0 in your units, value not important, as relief device needs inlet conditions at zero flow, stagnation conditions by setting stream pressure at relief device absolute set pressure and vapour fraction at 0 flash at bubble point.

2. Select inlet stream(single click), go to Sizing > Relief Device. Inlet/Outlet page relief device flow result is transferred to blowdown header inlet stream to allow back pressure to be checked.
3. Example is sizing an existing rupture disc mounted on a vessel of dimensions shown; note discharge coefficient (note rupture discs with vacuum supports C_d 0.625 x 0.6 factor = 0.375).

4. Vessel level / vapor volume fraction important as it determines onset of two phase flow for two phase models eg HEM, ERM, HNE, HFZ. If single phase model selected program will warn of onset of two phase flow.

5. Set pressure is normally same as stream pressure; overpressure % selected depends on codes.
6. Design method >use short cut, rigorous integral is for dynamic relief requiring Dynamics Module

7.Latent heat > rigorous minimum gives conservative result (increases relief flow) note vapour pressure relief sizing in its simplest form is based on Flow W = Heat Q / Latent Heat L

8. Vessel model > Churn Turbulent for fire cases, Bubbly for foaming cases and Homogeneous vessel for reactions; note homogeneous vessel is worst case and forces two phase flow with vapour fraction depending on level or inlet vapour fraction selected. The maximum allowable operating level is sometimes limited by maximum nozzle size on vessel.

9. Heat model depends on your design case; note you can add extra heat from reactions for the external fire cases. If you have lab data use tempered or non-tempered runaway models. Non-tempered runaway is the worst case scenario. For hybrid non-tempered or gassy non-tempered reactions use VSP results. For Gassy and Hybrid systems use RSST results.

10. Use other options to suit specific application noting that ignoring top / bottom head (changes exposed surface area), F = 1 for bare vessel MNL043A for other values; site insurance cover may dictate values used.
11. Set inlet / outlet piping; note you can define inlet stream from a complex flow sheet (many relief devices) and specify relief device outlet stream.

Additional Notes
If sizing relief devices, without vessel, set up a nominal flat bottomed vessel. This is method for liquid expansion thermal relief.
For Heat Model-Non-tempered Runaway select Vent Flow Model HFZ (Homogeneous Frozen)
Vessel Model HEM forces two phase flow.
Reducing maximum operating level or increasing vapor fraction reduces the relief device diameter. Maximum operating level in the vessel should be agreed.
Increasing allowable over pressure, depending on vessel design pressure and local codes, will reduce relief diameter.
For external fire heat model, say API 520. For worst case use inadequate fire protection and no vessel insulation i.e. F=1. We normally take no benefit for reactor jackets.
Client to advise what additional heat input to allow for any coincident reaction that might be taking place. Entry field available.

Shortcut and Rigorous Integral Method
Shortcut method is a steady procedure using customer set parameters
Rigorous method tries to model relief from a dynamic vessel.
Calculates area that would be required for 'all vapor' vent and then uses this as area for relief device on a dvsl, configured with user vent model and heat model. Run the dvsl, using two phase mass flux, check the pressure in the vessel and determine if it is less than the maximum pressure. If the vessel pressure rises above maximum, the vent area is increased and run again. When the calculations find an above / below bound for the minimum value, we change to bisection method and find the minimum area Sometimes the area calculated for 'vapor vent rate' is sufficient to relieve the vessel.

Appendix III
General Guidance Notes

Thermodynamics

Enthalpy decreases with rising pressure, increasing relief flow.
Latent heat decreases with increasing temperature, increasing relief flow.
Liquid specific heat increases with increasing temperature, increasing heat rates.

Choked flow through an orifice is isenthalpic. Where the presence of both liquid and vapor phases in the vent stream can be treated as a vapor-liquid mixture at equilibrium little error is introduced by carrying out the flash computation at constant enthalpy.

A constant entropy (isentropic) process is adiabatic and reversible. Flow through a nozzle approximates to an isentropic behaviour where there is little time for heat transfer and the process is frictionless. A relief valve, unlike other valves, approximates to an isentropic expansion, because it contains a nozzle and not an orifice. An isentropic expansion will be accompanied by a decrease in enthalpy which will result in a higher predicted nozzle mass velocity than that for an isenthalpic case.

Relief device sizing and rating is based on conditions at inlet being at stagnation with the pressure at relief device set pressure and bubble point vapor fraction of 0. For relief device sizing and rating not involving an actual vessel a pseudo flat bottomed vessel with an arbitrary level or void fraction is used in CHEMCAD.

Vent Flow Model Selection Criteria
Homogeneous - Equilibrium Model (HEM)

Non-slip model where vapor and liquid have same composition and velocity. HEM gives the best predictions for most cases of two-phase relief flows from tempered runaway reactions,subject to limitations in the following cases:

- Flows with nonvolatile liquids such as water or crude oil
- Flows of near-saturated mixtures
- Single phase liquids with high gas loading
- Dense phase releases
- Releases through different nozzle geometries

Homogeneous - Frozen Model (HFZ)

Flow of a non-volatile liquid and an insoluble gas phase ie nonflashing (frozen) flow from non-tempered runaway reactions

Equilibrium - Rate Model (ERM)

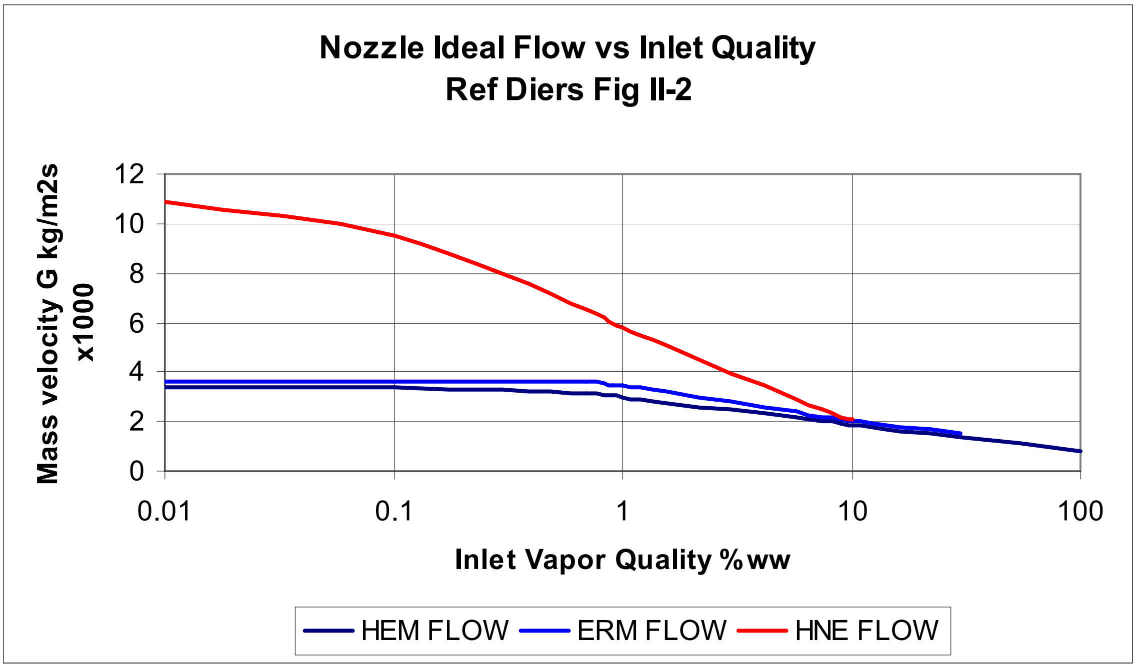
Single component systems and narrow boiling range mixtures using the isenthalpic rate of vaporization at the critical condition rather than isentropic value.

Henry Fauske's Homogeneous Non-equilibrium Model (HNE)

Vent model HNE should only be used if the process conditions under study match precisely all model requirements. HNE assumes ideal vapour/liquid equilibrium, vapour phase is an ideal gas, frictionless nozzle, turbulent flow, no heat gain or loss and overpressure in the range 10-30% of absolute relief pressure.

Vent Flow Model

The mass velocity in an ideal nozzle will depend on the model selected and the flowing quality as shown by Diers.[p 64] The graph below and the results for Diers Test in Appendix II-E shows the predicted flows through an ideal nozzle (C_d=1).



DIERS TEST APPENDIX II-E (page 113) RESULT SUMMARY		
VENT MODEL	**$G_{CHEMCAD}$ lb/sec-ft^2**	**G_{DIERS} lb/sec-ft^2**
HNE	881	856
ERM	794.5	778
HFZ	Two component nonvolatile liquid noncondensible gas system	
HEM	642	643

Note that all models converge at inlet vapor quality >10% w/w.
The HNE vent model results in a smaller diameter relief device. predicting a device **G** capacity 33% greater than the HEM vent model.

Vent header sizing

Relief vent header flow is typically choked and two phase in the annular flow regime with the friction factor being pressure dependent. The Churchill friction factor correlation, applicable to the laminar, transition and turbulent flow regimes is preferred to Fanning and Swamee Jain methods.

The Beggs and Brill method is preferred to the use of Baker flow regime maps for pressure drop correlation. Isothermal flow conditions provide a more conservative result over adiabatic conditions. The Beggs and Brill method is applicable for isothermal flow only and considers the effect of two liquid phases on pressure recovery due to elevation change. Generally more accurate than Baker method for systems with elevation change. If adiabatic flow occurs or flow with cooling or heating, the Beggs and Brill method will overpredict the pressure drop.

FIGURE 1

ERS SIZING STRATEGY

ERS SIZING STRATEGY

FIGURE 1
PID/JEE/02

NOTES

(1) HEM vessel model 2PH flow same vessel void fraction (X=m/V)
(2) Untempered hybrid refer HSE A6.2.2
(3) All gassy reactions are untempered. Single phase refer HSE A6.2.2
(4) Tempered single phase systems refer HSE A6.2.1
(5) See Technical Paper references 1&2